Hydrodynamics of Two-Phase Cocurrent Downflow through Packed Beds

Part I. Macroscopic Model

Pressure drop and liquid saturation are two important design parameters in cocurrent gas-liquid downflow through packed beds. A macroscopic model based on momentum balance is formulated for the condition of no radial pressure gradients. The model includes the effect of bubble formation on the pressure drop and holdup and is compared with the experimental data of the earlier investigators and of the present study. The model provides a functional form for correlating pressure drop and liquid saturation but some parameters have to be determined by fitting the experimental data.

V. G. RAO, M. S. ANANTH and
Y. B. G. VARMA

Department of Chemical Engineering
Indian Institute of Technology
Madras-600 036, India

SCOPE

Several studies have been reported on the hydrodynamics of two-phase cocurrent downflow in packed beds which is of extensive use in industrial practice ranging from synthesis of chemicals to waste water treatment. The approach for the prediction of pressure drop and liquid saturation in those studies has been mostly empirical and the first mechanical model due to Sweeney (1967) used the momentum balance and the absence of radial pressure gradients; the model, neglects all dynamic interactions between the phases.

A macroscopic model is formulated in the present study within the framework of the momentum balance using the experimentally observed condition of no radial pressure gradients. This formulation contains three parameters: two of them account for the effect of reduction in cross-sectional area available for flow of each phase due to the presence of the other, while the third accounts for the effect of bubble formation, breakage and reformation. Three well-defined regions of flow are identified and the model is satisfactorily applied to each of the regions separately.

CONCLUSIONS AND SIGNIFICANCE

The measurement of pressure drop and liquid saturation in gas-liquid cocurrent downflow through packed beds has been reported in literature and correlations that are presented to predict the aforementioned parameters are empirical and cannot be extrapolated. Although distinct flow regions have been identified, none of the earlier approaches have taken into account the fact that the contacting mechanism between the phases is different in the different regions of flow.

In the present treatment, the formulation is generalized, the assumptions are explicitly stated and an attempt has been made to take into account the interaction between the phases through bubble formation, breakage and reformation. The model also presents a logical framework for the inclusion of other dynamic interactions such as friction at the interface and entrainment and is thus more comprehensive in its presentation than the model due to Sweeney (1967). The model parameters are determined independently for each of the identified regions of flow. However, *a priori* values are provided for five parameters out of nine on physical grounds, as well as order of magnitude values for the remaining four. Actual values for these four have to be determined by fitting experimental data to the theory.

MATHEMATICAL FORMULATION

The hydrodynamics of two-phase flow-through packed beds has been theoretically considered by Sweeney (1967) taking geometric interaction between the liquid and the gas. This aspect is re-examined in this paper keeping in mind the different hydrodynamic regions of flow. Secondly, the dynamic interaction between liquid and gas is taken into account in modelling the flow by introducing the rate of energy dissipation through formation and breakage of bubbles. The model is compared with the experimental two-phase pressure drop data of the earlier investigations and of the present study.

Assuming isothermality, either the momentum balance or the mechanical energy balance can be used to calculate the pressure drop which can be looked upon as either the force per unit area of cross section required to overcome frictional forces or the energy per unit volume lost due to dissipation. Figure 1 is a simplified schematic diagram of the system with the following assumptions:

i) Each phase is treated as a continuum.

ii) The flow is steady with the voidage and holdup being constant and uniform.

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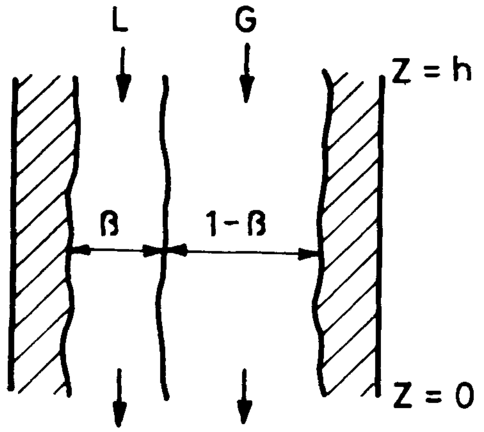

Figure 1. Formulation of the flow of the phases.

iii) Acceleration effects are negligible; i.e., the momentum and the kinetic energy of the fluid in each phase is the same for all values of Z ($0 \leq Z \leq h$).

The mechanical energy equation for the gas is given by

$$\Delta P_{lg}(1-\beta)SV_g' + ghS(1-\beta)\rho_g V_g' = \tau_{lg}A_{lg}V_{ig}' - \phi_g \quad (1)$$

Similarly for the liquid,

$$\Delta P_{lg}\beta SV_l' + ghS\beta\rho_l V_l' = -\tau_{lg}A_{lg}V_{il}' - \phi_l \quad (2)$$

where ϕ is the rate of energy dissipation. In the absence of slip, interfacial velocities, V_{ig}' and V_{il}' are equal. Hence,

$$\Delta P_{lg}[\beta V_l' + (1-\beta)V_g'] + [\beta\rho_1 V_1' + (1-\beta)\rho_g V_g']gh = -\frac{1}{S}(\phi_1 + \phi_g) \quad (3)$$

Apart from the dissipation terms, Eq. 3 contains two measurable unknowns ΔP and β, requiring an additional equation to calculate the two quantities. The derivation of Eqs. 1 and 3 assumes that the pressure does not vary across the cross-section, implying that ΔP in two-phase flow equals the pressure drop in each of the two phases. For downward flow, this equality may be expressed as:

$$\Delta P_{lg} = \Delta P_l = \Delta P_g = (\Delta P_{fl} - \rho_l gh) = (\Delta P_{fg} - \rho_g gh) \quad (4)$$

The frictional dissipation per unit area can be conveniently written in terms of the superficial mass velocities L and G and the frictional pressure losses of the individual phases:

$$-\frac{1}{S}(\phi_l + \phi_g) = \Delta P_{fg}\frac{G}{\rho_g} + \Delta P_{fl}\frac{L}{\rho_l} \quad (5)$$

The *LHS* of Eq. 5 is the rate of energy dissipation. So is the right-hand side of Eq. 5, but the frictional pressure loss terms, which are related to measurable qualities in Eq. 4, are amenable to mathematical modelling.

GEOMETRIC INTERACTION MODEL

A simple model for the frictional pressure losses, hereafter referred to as the geometric interaction model, can be formulated on the basis of the following assumptions:

(i) The two phases affect each other only to the extent of reduction in cross-sectional area for flow. Thus, if the liquid saturation (assumed uniform) in two-phase flow is β, the area for flow of the liquid phase is β times that in single-phase flow and so the velocity is $(1/\beta)$ times that in the latter.

(ii) The area of contact for friction in two-phase flow is an as-yet unspecified factor α'' times that in single-phase flow.

(iii) The friction factor for each phase in two-phase flow is a constant multiple, α', of that in single-phase flow.

From these assumptions and from a simple overall momentum balance it follows that

$$\frac{\Delta P_{fl}}{\Delta P_{fl}^{o}}\frac{(\text{Cross-Sectional Area})_{lg}}{(\text{Cross-Sectional Area})_s} = \frac{(\text{Shear Stress})_{lg}}{(\text{Shear Stress})_s} \times \frac{(\text{Area of Contact})_{lg}}{(\text{Area of Contact})_s}$$

$$\frac{(\text{Shear Stress})_{lg}}{(\text{Shear Stress})_s} = \frac{(\rho_l V_l'^2/2)_{lg}\psi_{lg}'}{(\rho_l V_l'^2/2)_s\psi_s'} = \frac{\alpha_l'}{\beta^2} \quad (6)$$

Combining the above gives the simple result that

$$\frac{\Delta P_{fl}}{\Delta P_{fl}^{o}} = \frac{\alpha_l}{\beta^3} \quad (7)$$

where $\alpha = \alpha'\,\alpha''$

Thus, the frictional pressure drop for flow of liquid in the presence of the gas is larger than that for flow of liquid alone through the same bed by a factor of $(1/\beta^3)$ times α_l which accounts for the difference in the effective area of contact between liquid and packing in the presence and in the absence of gas flow and for the change in the friction factor. Similarly for the gas, it follows that

$$\frac{\Delta P_{fg}}{\Delta P_{fg}^{o}} = \frac{\alpha_g}{(1-\beta)^3} \quad (7a)$$

The single-phase frictional pressure drops, ΔP_{fl}^{o} and ΔP_{fg}^{o} are assumed to be known and may be obtained using Ergun's or Tallmadge's expressions at the same mass flow rates and for flow through the same packed bed as for the two-phase flow.

Sweeney (1967), assuming that the liquid flows over the packing wetting it entirely and that the gas flows relative to the liquid over the "wet" packing, obtained the following expressions:

$$\alpha_1 = 1$$

$$\alpha_g = YZ^2\frac{C_1Y[C_2\mu_g(1-\epsilon)a_sZ^2]/V_g\rho_g}{C_1 + [C_2\mu_g(1-\epsilon)a_s]/V_g\rho_g} \quad (8)$$

where

C_1, C_2 = constants, characteristic of the packing
Y = factor accounting for the relative velocity of the gas with respect to the liquid
Z = geometric factor to account for the increase in effective size of packing due to the liquid film over it

Charpentier et al. (1969), however, note that $\alpha_g \simeq 1$ over a wide range of operating conditions, which is reasonable for the treatment of low interaction regions of flow because the absolute velocity of the gas is usually much greater than that of the liquid ($Y \simeq 1$) and the increase in the effective size of the packing due to wetting by liquid is negligible ($Z \simeq 1$).

The aforementioned geometric model is, however, inadequate in the high interaction regime because of friction at the interface, bubble formation, increased agitation, and drop formation or entrainment.

Friction at the interface can be partly accounted for in the empirical factors α_1 and α_g. However, when bubble formation occurs, the gas phase is no longer continuous and the mechanical energy balance, Eq. 3, does not strictly apply. However, Eq. 3 may be retained, and ΔP_1 and ΔP_g may be interpreted as those in an equivalent homogeneous phase. To account for the dynamic interaction, it is necessary to first calculate the additional pressure drop due to formation, breakage and reformation of bubbles.

The rate of work done due to bubble formation is a product of the rate of bubble formation and the work done in forming a single bubble, with both the quantities depending on the bubble radius, r. There are three forces acting on the bubble (Figure 2); the net pressure force acting downwards is balanced by the surface tension and the buoyancy forces:

$$\frac{\pi d^2}{4}\left[\left(\frac{\Delta P}{\Delta Z}\right)_{flg} d\right] = \pi d_o\sigma + \frac{\pi d^3}{6}(\rho_l - \rho_g)g \quad (9)$$

The bubble diameter is, therefore, given by

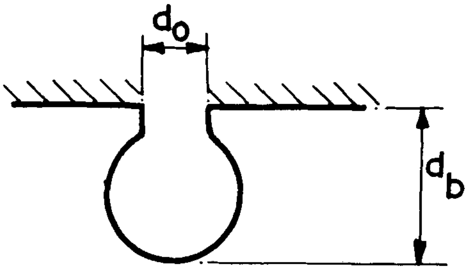

Figure 2. Formulation of the bubble formation.

$$\frac{d}{d_o} = \left[\frac{6s}{(3/2)\delta_{lg} - 1 + (\rho_g/\rho_l)}\right]^{1/3} \quad (10)$$

where $s = \sigma/\rho_l g d_o^2$ and $\delta_{lg} = \Delta P_{flg}/hg\rho_l$. The factor ρ_g/ρ_l is in most cases negligible and may be omitted without loss of accuracy. The orifice diameter, d_o, in Eq. 10 should be replaced by a length characteristic of the space between particles. In a packed bed, bubbles form in the small gaps in-between particles of packing. An estimate of the linear dimension of these gaps can be obtained as follows: in random packing, the average number of spherical particles along any diameter of the tube is usually one less than the maximum possible number, which is the integral part of D/d_p. Hence, the equivalent length, d_o, characteristic of bubble formation may be taken as:

$$d_o = [D - Nd_p]/N \quad (11)$$

where D is the column diameter and N is an integer calculated as the integral part of (D/d_p) minus one. It must be emphasized that this equation only provides a reasonable estimate for d_o. Changes in the value of d_o will be seen to affect the fitted value of α_b only slightly because the pressure drop due to bubble formation varies effectively as $(\alpha_b/d_o{}^{1/3})$.

The rate of bubble formation is given by the volumetric gas rate divided by the bubble volume, assuming equal size for all the bubbles. Bubbles may be assumed to form whenever the absolute velocity of the gas exceeds that of the liquid. Hence, the volumetric flow rate of the bubble-forming gas is given by

$$\left[\frac{G}{\rho_g\epsilon(1-\beta)} - \frac{L}{\rho_1\epsilon\beta}\right]\frac{\pi}{4}D^2\epsilon(1-\beta) \simeq \frac{G}{\rho_g}\frac{\pi}{4}D^2 \quad (12)$$

The above approximation is based on the fact that under normal operating conditions (Charpentier et al. 1969):

$$\frac{G}{\rho_g\epsilon(1-\beta)} \gg \frac{L}{\rho_l\epsilon\beta}$$

The number of bubbles per unit time is then given by

$$\dot{N}_b = \frac{G}{\rho_g}\left(\frac{\pi}{4}D^2\right)\Big/\left(\frac{\pi}{6}d^3\right) \quad (13)$$

The rate of work done additionally for bubble formation is given by $\dot{N}_b\pi d^2\sigma$, which on substitution gives

$$\alpha_b\frac{h}{d_p}\frac{6\pi}{4}\frac{G}{\rho_g}D^2\frac{\sigma}{d_o}\left[\frac{1.5\delta_{lg}{}^{-1}}{6s}\right]^{1/3}$$

where the factor $\alpha_b h/d_p$ is introduced to take into account the frequency of breakage and reformation; the maximum number of times a bubble may be expected to break and reform is once every time it traverses a particle in the packing. The empirical constant α_b accounts for the fact that in practice bubble may not break and reform quite so often. α_b is expected to be between 0 and 1.

The additional pressure drop of the gas due to bubble formation is obtained by dividing the aforementioned rate of work done by $(\pi/4\,D^2)(G/\rho_g)$. The frictional pressure drop for flow of gas from the present model can now be written as

$$\Delta P_{fg} = \frac{\alpha_g\Delta P_{fg}^o}{(1-\beta)^3} + 6\alpha_b\frac{\sigma}{d_o}\left[\frac{1.5\delta_{lg}{}^{-1}}{6s}\right]^{1/3}\frac{h}{d_p} \quad (14)$$

It is possible to take into account entrainment of the liquid in the gas phase for calculating ΔP_{fl} by using the analysis due to Tattersen et al. (1977) for predicting the mean size of drops entrained by a cocurrent flow of gas. It is, however, found that the present model for α_l together with Eqs. 7 and 14 describes the data of pressure drop and liquid saturation adequately.

The factor α_l is the ratio of the equivalent area of contact between liquid and packing in two-phase flow to that in single-phase flow. At low liquid flow rates ($L \leq 5$), the liquid saturation is unaffected by the gas flow rate ($G \leq 1$) supporting that $\alpha_l = 1$. This is true for gas continuous region, provided the gas flow rate is not too high to cause significant entrainment. It can be further assumed similarly that $\alpha_g = 1$ and $\alpha_b = 0$, since there is no bubble formation in this region. The resulting model for ΔP and β in the gas continuous region is essentially the same as Sweeney's (1967).

At intermediate and high liquid flow rates, the gas flow significantly reduces the liquid saturation. At low gas flow rates, i.e., in dispersed bubble flow, the bubble formation and coalescence result in filling up of many of the voids, thus reducing the area of contact between liquid and packing when compared with single-phase flow. It is assumed here that $\alpha_l'' = 1$ and that this reduction is equal to the holdup itself, i.e., that $\alpha_l = \beta$. The factor α_g should be large in this region, as bubbles increase the interfacial area for frictional losses and alter the friction factor as well. α_b should be maximum for this region subject to the constraint that it lies between 0 and 1. α_g and α_b are determined by fitting the experimental data with the theory.

As the gas flow rate is increased at intermediate and high liquid flow rates, the flow is characterized by alternate pulses of low and high density. The bubble formation in the high density pulse reduces α_l, though the reduction is less than in dispersed bubble flow. The increased gas flow rate, on the other hand, causes spreading of the liquid film over a large area of packing. The effect of bubbles in reducing α_l is expected to be more significant with α_l being less than unity. By trial and error, a value of 0.75 for α_l was chosen as one which gives predictions in agreement with experiment. The values of α_g and α_b should be less than that for dispersed bubble flow, with the former being greater than 1. In summary, of the nine parameters in the model, values are assigned *a priori* to five parameters as follows:

$\alpha_1 = 1, \alpha_g = 1, \alpha_b = 0$	Gas Continuous Flow	
$\alpha_1 = 0.75$	Pulse Flow	(15)
$\alpha_1 = \beta$	Dispersed Bubble Flow	

DETERMINATION OF MODEL PARAMETERS

The two-phase pressure drop and the liquid saturation are obtained using Eqs. 4 and 6. The single-phase pressure drop is calculated using Ergun's equation (1952):

$$\left(\frac{\Delta P_f^o}{h}\right) = 150\frac{\mu V}{d_p^2}\frac{(1-\epsilon)^2}{\epsilon^3} + 1.75\frac{V^2}{d_p}\frac{1-\epsilon}{\epsilon^3} \quad (16)$$

The final equation for ΔP_{flg} and β are obtained by substituting the model equations, Eqs. 7 and 14, into Eqs. 4 and 6. These equations are conveniently written in dimensionless form as:

$$\delta_{lg} - \beta - \alpha_1\frac{\delta_1^o}{\beta^3} + 1 = 0 \quad (17)$$

$$\delta_{lg} - \beta + \gamma_1 + \gamma_g\frac{\rho_g}{\rho_1} - \left[\alpha_g\frac{\delta_g^o}{(1-\beta)^3} + \alpha_b(6s)^{2/3}(1.5\,\delta_{lg}^{-1})^{1/3}(d_o/d_p)\right]\gamma_g - \alpha_1\frac{\delta_1^o}{\beta^3}\gamma_1 = 0 \quad (18)$$

where $\delta_{lg} = \Delta P_{flg}/\rho_l gh$
$\gamma_l = (L/\rho_l)/[(L/\rho_1) + (G/\rho_g)]$
$\gamma_g = 1 - \gamma_l$

The terms of order ρ_g/ρ_l are neglected in comparison with δ or 1.

TABLE 1. MAGNITUDE OF DIFFERENT CONTRIBUTIONS TO THE TOTAL FRICTIONAL PRESSURE DROP AS PREDICTED BY THE PRESENT MODEL FOR TYPICAL MASS FLOW RATES OF THE PHASES
(6.72 Ceramic Spheres; d_p = 6.72 mm; ϵ = 0.373; d_t = 92.4 mm)

			Dimensionless pressure drop		
L	G	Flow Pattern	Geometric Liq.-Phase	Interaction Gas Phase	Contr. for Bubble Formation
90.818	0.3053	DBF	1.8373	3.0960	1.2846
74.306	0.6598	DBF	0.9903	4.7733	1.6315
37.153	0.3053	DBF	0.4012	0.8225	1.0380
57.794	0.6569	DBF	0.6338	3.5389	1.5263
33.025	0.1613	DBF	0.4692	0.2924	0.8055
37.153	1.7987	PF	0.1712	5.2341	0.6580
16.512	0.3053	PF	0.0889	0.2197	0.2034
8.256	1.7987	PF	0.0218	2.4259	0.5066
16.512	1.7989	PF	0.0526	3.1960	0.5597
24.769	1.7986	PF	0.0926	3.9849	0.6029
4.128	0.3053	TF	0.0181	0.1290	
4.128	0.4661	TF	0.0133	0.2577	
8.256	2.7048	SF	0.0243	5.5947	
4.128	2.1494	SF	0.0095	3.1156	
4.128	2.6565	SF	0.0101	4.4066	

DBF = dispersed bubble flow; PF = pulse flow; SF = spray flow; TF = transitional flow.

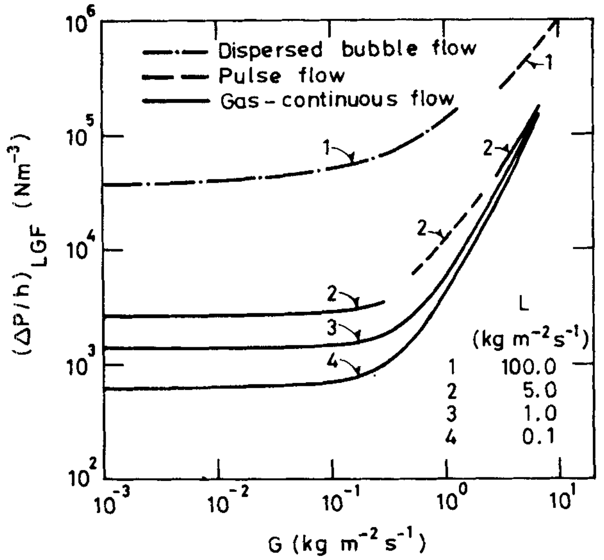

Figure 3. Variation in pressure drop with mass flow rates of the phases as predicted by the present model.

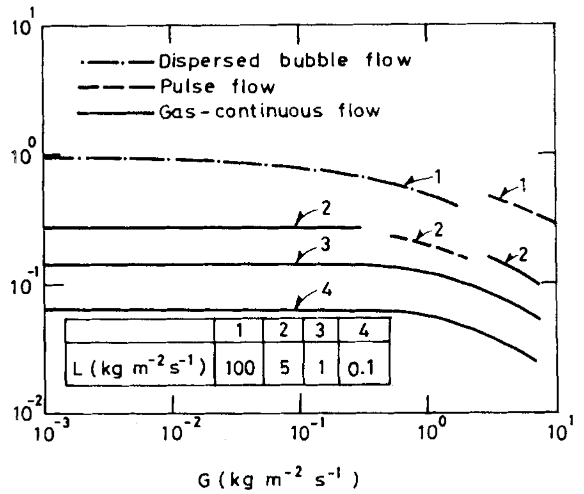

Figure 4. Variation in total liquid saturation with mass flow rates of the phases as predicted by the present model.

PARAMETER ESTIMATION AND PROCEDURE FOR CALCULATION OF ΔP and β

Data on two-phase flow are always reported either as ΔP or as β alone for given operating conditions. Simultaneous measurements of Δp and β are very tedious and are never reported. This poses severe constraints on the determination of parameters in Eqs. 17 and 18. In this work, only two-phase pressure drop data are used to determine α_g and α_b in the dispersed bubble flow and pulse flow regions. The model predictions for liquid saturation are, however, compared subsequently with experiment. Secondly, some of the values assigned to α_l, α_g and α_b *a priori* in Eq. 15 are admittedly open to discussion. However, as is clear from the form of Eqs. 17 and 18, this procedure permits the evaluation of the remaining parameters unambiguously using a simple linear least-squares program.

Using the experimental pressure drop, liquid saturation is calculated from Eq. 17. Then a linear least-squares analysis of Eq. 18 using experimental δ_{lg} and the calculated β gives the "best" values of the constants α_g and α_b for each region as:

Dispersed Bubble Flow	$\alpha_l = \beta$	$\alpha_g = 5.132$	$\alpha_b = 0.446$
Pulse Flow	$= 0.75$	$= 1.329$	$= 0.174$
Gas-Continuous Flow	$= 1.0$	$= 1.0$	$= 0$

Once the parameters are estimated, δ_{lg} and β are calculated from input data on packing characteristics, fluid properties and the flow rates of the phases solving Eqs. 17 and 18 using an interactive procedure.

Table 1 shows the relative magnitude of the different contributions to the total frictional pressure drop in two-phase flow for typical values of G and L in three different regions. It is clear that the contribution to the total pressure drop due to bubble formation is small but significant in both dispersed bubble and pulse flow regions.

MODEL SIMULATION

The model is simulated on the computer IBM 370/155 for air-water cocurrent downflow through a packed bed of 92.4 mm i.d. filled with a spherical ceramic packing of 6.72 mm diameter. (Details of the experiment and the data are being published separately.) The results of the simulation are shown in Figures 3 and 4, respectively, for pressure drop and liquid saturation. It is assumed in the theoretical development that the region of operation is known *a priori*. The experimentally observed flow pattern, Figure 5, is used to calculate the curves shown in Figures 3 and 4, wherein the flow regions identified by different types of lines with blanks indicating the diffuse (experimentally observed) transition between regions.

At low liquid rates ($L \leq 0.1$), Figure 3 shows that the calculated pressure drop is nearly independent of the gas flow rate for $G \leq 0.1$. There is a steep rise in the two-phase pressure drop for $G > 0.2$. Figure 4 shows that the total liquid saturation at low liquid rates is similarly independent of the gas rate for $G \leq 0.5$, thereafter decreasing as G increases.

At intermediate liquid flow rates ($L = 5$), the pressure drop and liquid saturation are nearly independent of the gas flow rate in trickle flow region ($0 < G < 0.37$). In the pulse flow ($0.37 < G < 2.3$) and spray flow regions ($G > 2.3$), the pressure drop increases rapidly with increase in gas flow rate while the liquid saturation decreases. The greater slope for the pressure drop curve in the spray flow region indicates the greater effect of the gas flow rate in this region.

At high liquid rates (L = 100), the pressure drop and liquid saturation are independent of gas flow rate in dispersed bubble flow region ($0 < G < 1.8$) for $G < 3 \times 10^{-3}$. Thereafter, as G is increased, the two-phase pressure drop and liquid saturation decrease at first slowly ($3 \times 10^{-3} < G < 0.3$) and then rapidly ($G > 0.3$). The greater slope for the pressure drop curve in the pulse flow region

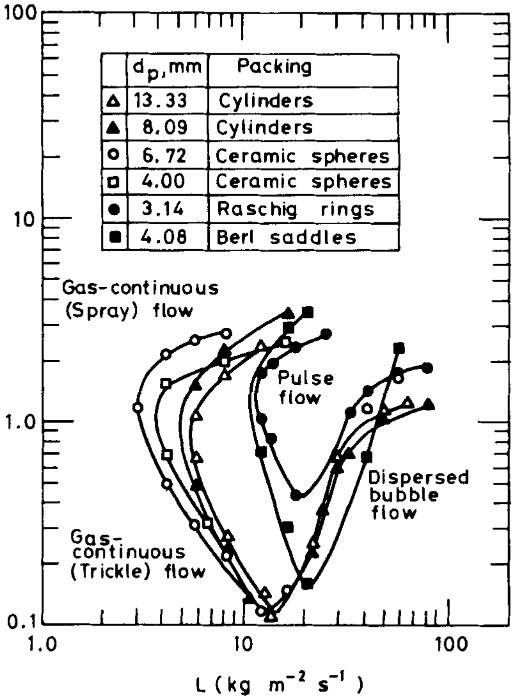

Figure 5. Flow pattern of the phases (shown as suggested by Sato et al., 1973).

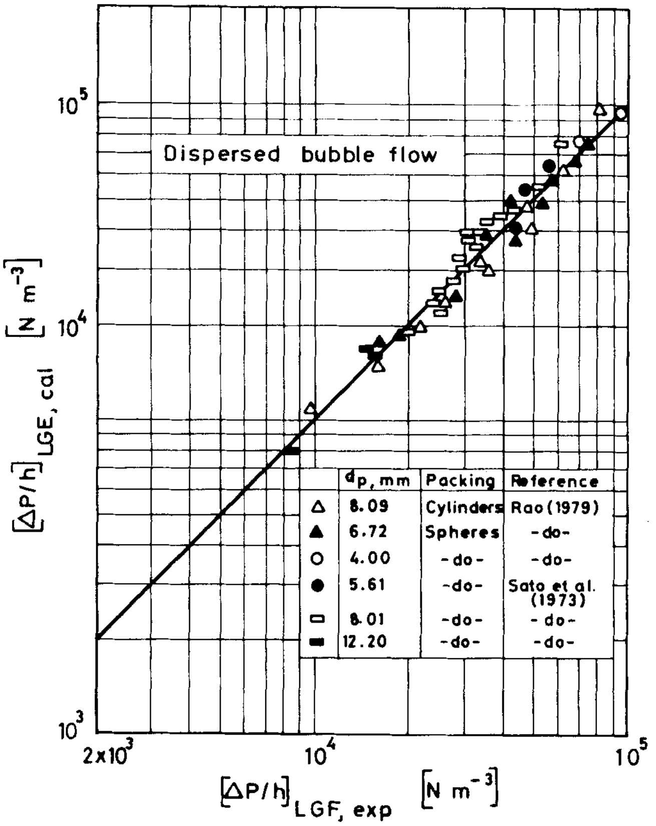

Figure 6. Comparison of two-phase experimental pressure drop data for dispersed bubble flow with the data obtained using the present model.

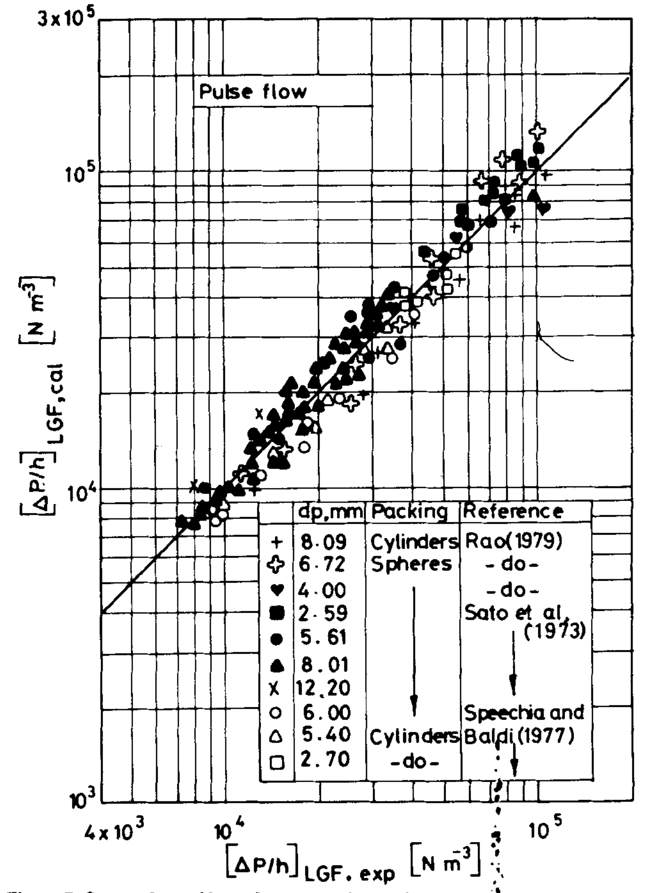

Figure 7. Comparison of two-phase experimental pressure drop data for pulse flow with the data obtained using the present model.

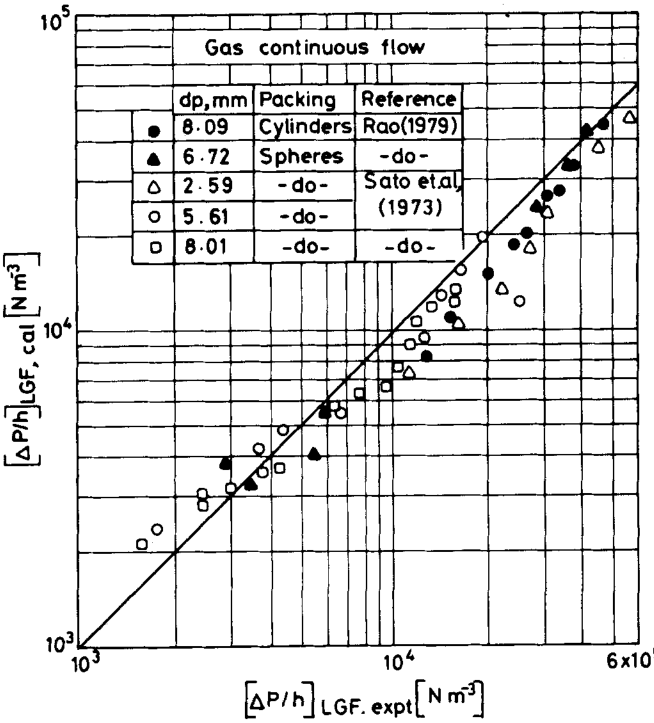

Figure 8. Comparison of the two-phase experimental pressure drop data for gas continuous flow with the data obtained using the present model.

($G > 1.8$) for $G > 0.3$ indicates the greater effect of the gas flow rate on the two-phase pressure drop in pulse flow compared to that in the dispersed bubble flow.

In general, the coupling between Eqs. 17 and 18 is such that, in any given region whenever the pressure drop increases, the total liquid saturation decreases.

THE MODEL VS. EXPERIMENTAL DATA

Figures 6 to 8 compare the predicted and experimental pressure drop, using Eqs. 17 and 18 with the experimental values of the present study (Rao, 1979) and that of Sato et al. (1973) and Specchia

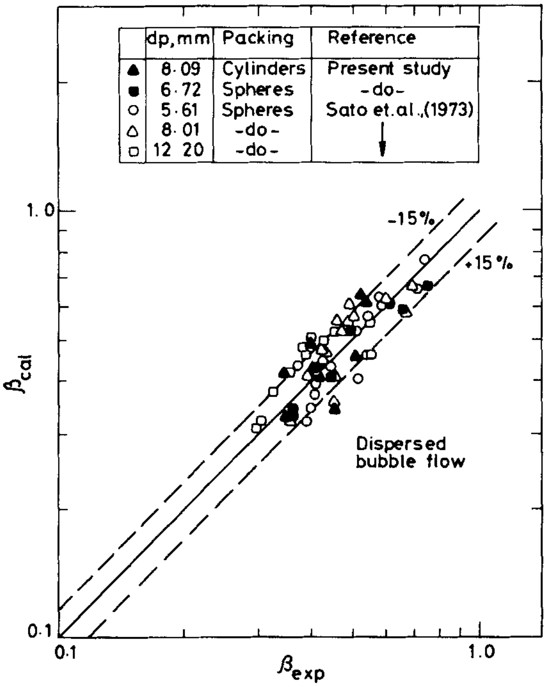

Figure 9. Comparison of the experimental data on total liquid saturation for dispersed bubble flow with the data obtained using the present model.

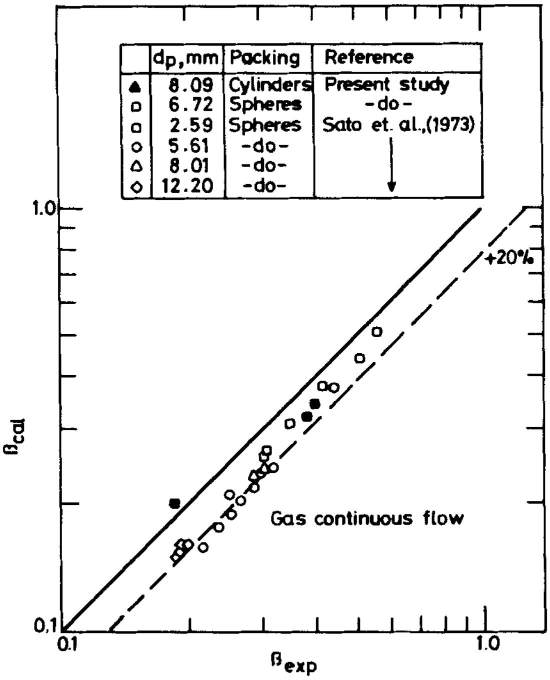

Figure 11. Comparison of the experimental data on total liquid saturation for gas continuous flow with the data obtained using the present model.

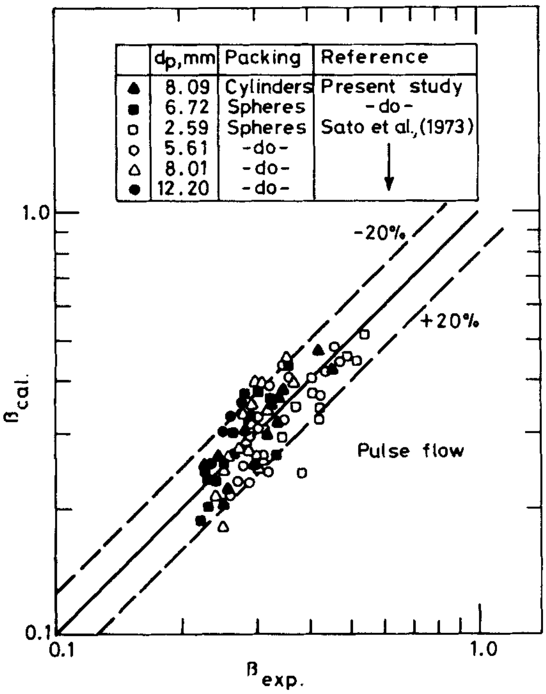

Figure 10. Comparison of the experimental data on total liquid saturation for pulse flow with the data obtained using the present model.

and Baldi (1977) for spherical and cylindrical packing of different geometry for the three identified flow regions, i.e., dispersed bubble flow, pulse flow, and gas-continuous flow. The model predicts the total liquid saturation simultaneously with the pressure drop. Figures 9 to 11 compare the theoretical and measured total liquid saturation; the overall standard deviations in respect of the pressure drop and liquid saturation are in Table 2.

The theoretical model proposed in the present study appears to predict values of the pressure drop and total liquid saturation satisfactorily for the three regions of cocurrent gas-liquid downflow through packed beds. The model requires, however, a knowledge of single-phase pressure drop to predict the two-phase pressure drop; the former, however, is easily obtained using either Ergun's

TABLE 2. OVERALL STANDARD DEVIATIONS

Flow Pattern	Two-phase Pressure Drop: Total No. of Experimental Data	σ	Total Liquid Saturation: Total No. of Experimental Data	σ
Dispersed Bubble Flow	56	13.0	60	13.7
Pulse Flow	175	18.2	94	15.3
Gas-Continuous Flow	48	23.9	28	18.6

or Tallmadge's expressions, the use of either of the expressions alters the parameters α_l, α_g and α_b very slightly, but does not alter the overall agreement between the model and the experiment.

In summary, the present model makes three significant contributions:

a) It provides a framework within which ΔP_{fl} and ΔP_{fg} can be modelled independently, while still satisfying the mechanical energy balance.

b) It shows that bubble work, hitherto assumed to be negligible, can be important, if one considers breakage and reformation of bubbles in packed beds and particularly when the average bubble size is small.

c) The model can be more reliably extrapolated than empirical correlations.

NOTATION

a_s = specific surface area of the particle, m^{-1}
A = interfacial area, m^2
C_1, C_2 = constants in Eq. 8
d = bubble diameter, m
d_o = orifice diameter or equivalent length, m
d_p = effective particle diameter, m
D = column diameter, m
g = acceleration due to gravity, $m{\cdot}s^{-2}$
G = superficial mass velocity of gas, $kg{\cdot}m^{-2}{\cdot}s^{-1}$
h = height of packed bed, m
L = superficial mass velocity of liquid, $kg{\cdot}m^{-2}{\cdot}s^{-1}$
$\dot{N}_b$ = number of bubbles per unit time, s^{-1}
ΔP = pressure drop, $N{\cdot}m^{-2}$

s = factor defined as in Eq. 10 [$=\sigma/\rho_l g d_o^2$]
S = $(\pi/4)D^2\epsilon$, free cross-sectional area
V = superficial velocity, $m \cdot s^{-1}$
V' = absolute velocity, $m \cdot s^{-1}$
Y,Z = factors defined in Eq. 8

Greek Letters

α = empirical constant introduced in Eqs. 7 and 7a
α', α'' = empirical constants
α_b = nonideality factor, $0 \leq \alpha_b \leq 1$
β = total liquid saturation based on void volume
δ = dimensionless pressure drop [$=\Delta P/\rho g h$]
ϵ = porosity
γ_l = $[L/\rho_l)/(L/\rho_l) + (G/\rho_g)]$
μ = viscosity, $kg \cdot m^{-1} \cdot s^{-1}$
ρ = density, $kg \cdot m^{-3}$
σ = surface tension, $N \cdot m^{-1}$; standard deviation
τ = shear stress, $N \cdot m^{-2}$
ϕ = Energy dissipation, $N \cdot m \cdot s^{-1}$
ψ' = friction factor $[\Delta P/(\rho V^2/2)](d_p/h)$

Subscripts

f = frictional
g = gas
i = interface
l = liquid
lg = two-phase
sp = single phase
w = water

Superscripts

o = single phase

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Part II: Experiment and Correlations

Two-phase pressure drop and dynamic and total liquid saturation are experimentally determined for air-water system under cocurrent downflow through packed beds using packings differing widely in geometry. The experimental data of the present study as well as that available in literature is satisfactorily correlated in terms of: (a) Lockhart-Martinelli parameters; and (b) the Reynolds numbers defined for the respective phases and the bed porosity, taking into account the flow behavior of the phases through the packed bed.

V. G. RAO, M. S. ANANTH and Y. B. G. VARMA

Department of Chemical Engineering
Indian Institute of Technology
Madras-600 036, India

SCOPE

Gas-liquid cocurrent downflow is commonly encountered in chemical engineering practice as, for example, in synthesis of chemicals, hydrogenation and hydrodesulpurization of petroleum products, waste water treatment etc. Several studies have been reported in literature for foaming and nonfoaming systems in respect to the flow pattern delineation, the two-phase pressure drop, and the total and dynamic liquid saturation. The correlations in respect to pressure drop and liquid saturation are either based on Lockhart-Martinelli concept or in terms of directly measurable variables or the corresponding dimensionless groups and cover the entire region of operation.

Although distinct flow regions have been identified and the flow interaction mechanism is different in each identified region, the fact has not been considered in detail in the earlier investigations in developing the correlations except for delineation of the flow into poor and high interaction regimes (Midoux et al., 1976; Specchia and Baldi, 1977). This aspect, which is found to be of importance from the theoretical study, is included in the present investigation to correlate the experimental data more satisfactorily.

CONCLUSIONS AND SIGNIFICANCE

The pressure drop and dynamic and total liquid saturation are the important parameters in the design of equipment for gas-liquid cocurrent downflow through packed beds which is of common use in chemical engineering practice. The system presents different flow pattern depending on the flow rates of the phases and the packing characteristics and thus gives rise to different contacting mechanism between the phases in the different regions of flow.

This aspect has been considered in the present study, while formulating the correlations for the two-phase pressure drop and liquid saturation separately for each identified flow region, viz., the gas-continuous flow, the pulse flow, and the dispersed bubble flow. The correlations are presented in terms of: (i)