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Correlations for predicting single phase and two-phase flow pressure drop in pebble bed flow channels

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ABSTRACT

An experimental study was conducted on the pressure drop of the single phase and the air–water twophase flow in the bed of rectangular cross sections densely filled with uniform spheres. Three kinds of glass spheres with different equivalent diameters (3 mm, 6 mm, and 8 mm) were used for the establishment of the test sections. The Reynolds number in the experiment ranged from a dozen to thousands for the singlephase flow and from hundreds to tens of thousands for the two-phase flow. In the present flow-regime model, the bed was subdivided into a near-wall region and a central region in order to take the wall effect into account to improve the prediction at low tube-to-particle diameter ratios. Improved correlations are obtained based on the previous study to consider the single-phase flow pressure drops for finite pebble beds with spherical particles and nonspherical particles by fitting the coefficients of that equation to both the database and the present experiment. The correlation is consistent with the observed physical behavior which explains its comparatively good agreement with the experimental data. A new empirical correlation for the prediction of two-phase flow pressure drops was proposed based on the gas phase relative permeability as a function of the gas phase saturation and the void fraction. The correlation fit well for both experimental data of spherical particles and nonspherical particles.

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1. Introduction

An accurate prediction of the pressure drop for single- and twophase flow through porous media composed of stationary granular particles is critical in the design and operation of the pebble bed reactor which will be one of the most promising reactors in safety and efficiency.

Since the well-known works of Darcy or Forchheimer, in particular, most of the researches concerning the single phase flow pressure drop in packed beds were either based on an overall analysis of the bed as a continuum or as a porous medium affected by the porosity distribution which was a function of the shape and size of the packing materials, bed geometric ratio and porosity profiles (Nemec and Levec, 2005a; Montillet et al., 2007). Ergun's equation stated that the pressure drop of the flow through a bed of spherical particles with a uniform size was a result of kinetic and viscous losses (Ergun and Orning, 1949). Bernsdorf et al. (2000) concluded that a significant error occurred when only shear forces were taken into account. Most researchers believed that Ergun constants were determined empirically for each bed for they were not only dependent on the particle geometry but varied from one macroscopic bed to another (made of the same particles) owing to the different structures of the packing within the bed after repacking (Jiang et al., 2000). Nevertheless, Nemec and Levec (2005a) believed that there existed some principles for the values. Recent theoretical works developed different approaches to the modeling of the pressure drop (Montillet et al., 2007; Çarpinlioğlu and Özahi, 2008).

The hydrodynamics of the two-phase flow pressure drop in packed beds was traditionally studied from an empirical point of view. Previous researchers suggested the prediction of pressure drop and liquid holdup by means of empirical correlations based on dimensional analysis and visual observations over a relatively narrow range of experimental conditions (Wammes and Westerterp, 1991; Ratnam et al., 1993). Four groups of models in general were reported and are concluded here as shown in Table 1. The first group followed the Lockhart-Martinelli correlations in a horizontal pipe and used the parameter as well as the two-phase multipliers (Goto and Gaspillo, 1992), and the definitions of the parameters were modified in some of the cases (Ellman et al., 1988). In the second group the pressure drop was related to the gas and liquid phases through relative permeability (Boyer et al., 2007). The third set was based on a mechanic approach in which mass and momentum balances were built for gas and liquid phases with closure laws for interfacial forces mainly between liquid and solid and between liquid and gas (Tung and Dhir, 1988; Boyer et al., 2007; Schmidt, 2007). Different from the above three groups of models in which a uni-

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Nome	nclature
А, В	coefficients of Ergun-type equation
D	hydraulic diameter of packed bed (m)
d _e	equivalent particle diameter $[6V_p/S_p](m)$
Ga _s *	modified Galileo number for <i>s</i> phase $[Ga_s^* =$
	$(\rho_s/\mu_s)^2 g(d_e \varepsilon/(1-\varepsilon))^3$
g	gravitational acceleration (m s ⁻²)
j _s	superficial velocity of s phase (m s ⁻¹)
k _s	viscous relative permeability for s phase
k _{st}	inertial relative permeability for s phase
L	length of the packed bed (m)
п	exponent in Eq. (12)
Res	Reynolds number [Re _s = $\rho_s j_s d_e / \mu_s$]
Re _s *	modified Reynolds number for s phase $[Re_s^* =$
4 D	$\rho_{sJs} d_e / [\mu_s (1 - \varepsilon)]]$
ΔP_{S}	total pressure drop including gravitational contri-
n	prossure (Pa)
р n	$pressure (ra)$ $copillary pressure [n - n - n] (P_2)$
p_c	surface of particle (m^2)
Sp V.	volume of particle (m^3)
X^{p}	constant in Eq. (15)
Greek l	etters
α	void fraction
ε	bed porosity
σ	surface tension (N m ⁻¹)
μ	dynamic viscosity of s phase (Pa s)
ρ	density of s phase (kg/m ³)
ψ	pressure drop
Subscri	pts
S	gas or liquid phase
g	gas phase
1	liquid phase
р	particle
w	wall

form correlation was obtained for the whole operating range, the fourth set was presented with different correlations for different flow patterns based on the theoretic analysis of the characteristic of the flow patterns. More recent experimental works focused mainly on proposing a more refined cartography of flow patterns (Bai et al., 2010). To review the models quantitatively, the correlations of the first three groups of models were compared with the experimental data collected from references. And the result of the fourth model was cited here directly from the original reference. All the resulting mean relative errors are listed in Table 1. The smallest mean relative error occurred with the fourth set of models.

The objective of the present work is to acquire a profound knowledge about the single-phase and two-phase flow pressure drops as well as the effect of the parameters and the bed specifications. Three kinds of particles with different equivalent diameters (3 mm, 6 mm, and 8 mm) are used for the establishment of the test section. According to the experimental data from the present work and references available, two new empirical correlations for single-phase and two-phase flow pressure drop, respectively, are proposed based on different models with greater soundness and accuracy.

Table 1

Mean relative error of four groups of models.

Model no.	Reference	Mean relative e	rror (%)
		Pressure drop	Liquid holdup
	Larkins et al. (1961)	30	50
	Charpentier et al. (1969)	118	11
1	Midoux et al. (1976)	127	36
	Wammes and Westerterp (1991)	71	17
	Larachi et al. (1991)	73	17
	Sáez and Carbonell (1985)	30	20
	Sáez et al. (1986)	20-30	-
2	Nemec et al. (2001)	40	10
	Lakota et al. (2002)	40	10
	Nemec and Levec (2005b)	41	7
	Holub et al. (1993)	40	17
	Al-Dahhan et al. (1998)	30	13
3	Attou et al. (1999)	60	17
	Narasimhan et al. (2002)	30	13
	Boyer et al. (2007)	20	20
	Turpin and Huntington (1967)	38 ^a , 31 ^a , 19 ^a	-
4	Varma et al. (1997)	14 ^a , 12 ^a , 2.7 ^a	-

^a RMSD.

2. Pressure drop models

2.1. Single-phase flow pressure drop

The pressure drop of single-phase flow in infinite packed beds is known to be dependent on Reynolds number, bed porosity ε , particle geometry and particle size distribution. The pressure drop is the result of the frictional losses and the inertia characterized by the linear dependence of flow velocity and the quadratic dependence of flow velocity, respectively (Vafai and Tien, 1981; Çarpinlioğlu and Özahi, 2008).

$$\frac{\Delta P_s}{L} = A \frac{(1-\varepsilon)^2}{\varepsilon^3} \mu_s \frac{j_s}{d_e^2} + B \frac{(1-\varepsilon)}{\varepsilon^3} \rho_s \frac{j_s^2}{d_e} \tag{1}$$

where ΔP_s is the overall pressure drop including gravitational contribution of *s* phase, *L* the length of the packed bed, *j_s* the superficial velocity of *s* phase and *d_e* is the equivalent particle diameter; *A* and *B* are constants. Ziolkowska and Ziolkowski (1988) defined the streamline flow, the transitional flow and the turbulent flow. In order to cover the whole range of Reynolds numbers from the streamline flow to the turbulent flow, the dimensionless pressure drop for *s* phase, Ψ_s , is written in terms of modified Reynolds and Galileo numbers as follows (Eisfeld and Schnitzlein, 2001; Niven, 2002; Nemec and Levec, 2005a).

$$\psi_s = \frac{\Delta P_s}{\rho_s j_s^2} \frac{d_e}{L} = A \frac{\mathrm{Re}_s^*}{\mathrm{Ga}_s^*} + B \frac{\mathrm{Re}_s^{*2}}{\mathrm{Ga}_s^*}$$
(2)

$$\operatorname{Re}_{s}^{*} = \frac{\rho_{s} j_{s} d_{e}}{\mu_{s}(1-\varepsilon)}$$
(3)

$$Ga_{s}^{*} = \left(\frac{\rho_{s}}{\mu_{s}}\right)^{2} g\left(\frac{d_{e}\varepsilon}{1-\varepsilon}\right)^{3}$$

$$\tag{4}$$

$$d_e = \frac{6V_p}{S_p} \tag{5}$$

where the subscript *s* represents the liquid-phase *l* or the gas-phase *g*, respectively; ρ_s , Re_s*, Ga_s* and μ_s are the density, the modified Reynolds number, the modified Galileo number and the dynamic viscosity for *s* phase, respectively, *g* the gravitational acceleration, and d_e is the equivalent particle diameter; V_p and S_p represent the particle volume and the particle surface.

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TP

two-phase flow

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Through a linear regression analysis for the different data sets, Leva et al. (1950) and Eisfeld and Schnitzlein (2001) assumed that the Ergun-type pressure drop equation was valid within the overall range of the investigated parameters. However, this correlation fails to reflect explicitly any influence due to the finite size of real packed beds. In practice any packing is bounded by confining walls, so that the influence of the tube-to-particle diameter ratio, D/d_e , on the pressure drop is expected. In other words, for the finite packed beds, there exists a wall zone where the average porosity is greater than in the core of the bed since the spatial distribution of particles conforms to the shape of the wall (Achenbach, 1995). For sphere particle bed, the strong dependence of the pressure drop on the porosity causes a nonuniform velocity distribution across the particle bed, since the disturbance of the statistical particle arrangement adjacent to the wall generates here a higher porosity than the average value in the bed (Achenbach, 1995).

2.2. Two-phase flow pressure drop

For two-phase flow, the microscopic equations of motion for the liquid and gas phase under tricking flow conditions were first implemented by Sáez and Carbonell (1985) and Sáez et al. (1986) and then extended by Nemec et al. (2001) and Nemec and Levec (2005b):

$$-\frac{\mathrm{d}p_l}{\mathrm{d}z} + \rho_l g = \left[\frac{A}{k_l} \frac{\mathrm{Re}_l^*}{\mathrm{Ga}_l^*} + \frac{B}{k_{lt}} \frac{\mathrm{Re}_l^{*2}}{\mathrm{Ga}_l^*}\right] \rho_l g \tag{6}$$

$$-\frac{\mathrm{d}p_g}{\mathrm{d}z} + \rho_g g = \left[\frac{A}{k_g}\frac{\mathrm{Re}_g^*}{\mathrm{Ga}_g^*} + \frac{B}{k_{gt}}\frac{\mathrm{Re}_g^{*2}}{\mathrm{Ga}_g^*}\right]\rho_g g \tag{7}$$

where k_s and k_{st} are defined as viscous and inertial relative permeability for *s* phase, respectively.

Previous researchers used the representation of the capillary pressure $p_c = p_g - p_l$ which has extensive use in the analysis of the multiphase flow through porous media to produce a relation between p_g and p_l . However, the validity of such a representation for packed beds is not established. This representation implies that the capillary pressure gradient is proportional to the gradient of the liquid fraction $(1 - \alpha)$. As the liquid fraction does not change significantly for a steady-state fully established flow, the capillary pressure gradient dp_c/dz is negligible with regard to the fluid pressure gradient (Attou et al., 1999):

$$\frac{\mathrm{d}p_g}{\mathrm{d}z} = \frac{\mathrm{d}p_l}{\mathrm{d}z} = \frac{\mathrm{d}p_{\mathrm{TP}}}{\mathrm{d}z} \tag{8}$$

where the subscript TP represents two-phase flow.

As is mentioned above, the dimensionless pressure drop in the single phase flow results from two independent contributions: the viscous term and the inertial term and they are dominant at low and high Reynolds number, respectively. In the general case of two-phase flow, a similar constitutive equation for the dimensionless pressure drop, $\Psi_{\rm TP}$, is proposed, in which each contribution to the force corresponding to the single-phase flow is scaled by a factor, the relative permeability. The relative permeability is defined as the ratio of the single-phase flow pressure drop (leaving the gravity pressure drop out of consideration) to the two-phase flow pressure drop obtained at the same interstitial velocity:

$$k_{\rm s} = \frac{\Delta p_{\rm s}}{\Delta p_{\rm TP}} \tag{9}$$

Based on Eq. (2), the dimensionless pressure drop for two-phase flow can be expressed as

$$\psi_{\rm TP} = \frac{A}{k_{\rm s}} \frac{{\rm Re}_{\rm s}^{\rm s}}{{\rm Ga}_{\rm s}^{\rm s}} + \frac{B}{k_{\rm st}} \frac{{\rm Re}_{\rm s}^{\rm s2}}{{\rm Ga}_{\rm s}^{\rm s}} \tag{10}$$

Table 2

1)	J	J									ć			ć			ć	ć				è	è	è							è	è		è		è	è	è	è	è														Į		1	1	1	1	1		-		4	•			Į		ļ	ļ		ļ		•		1	1	•	t		1	ſ]	l	1	1		ί	(X	2		t	ľ	1	1	1	l	ł	8	ć	t	1	5	5	1)	(1		l	(1	
	4	1	4	4	1	1	1	4	4	4	4	4	4	4	4	4	4	4	4	4	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1										4	•			Į		ļ	ļ		ļ		•		1	1	•	t		1	ſ]	l	1	1		ί	(X	2		t	ľ	1	1	1	l	ł	8	ć	t	1	5	5	1)	(1		((1	

Particle type	X
Spheres	4.37
Cylinders	6.54
Extrudates	3.31
Raschig rings	9.52

Viscous and inertial relative permeability are important in the study on multiphase flow through porous media (Scheidegger, 1974). It is assumed that the relative permeability of each phase remains the same in both viscous and turbulent flow regimes. Sáez and Carbonell (1985) produced experimental evidence to support the following assertion

$$k_s = k_{st} \tag{11}$$

It is still controversial that the relative permeability is independent of surface tension (Eötvös number), liquid and gas phase Reynolds and Galileo number. However, there is a dominating assumption in the previous works that the relative permeability is the function exclusively of liquid holdup. Based on this assumption, Sáez and Carbonell (1985), Attou et al. (1999), Nemec et al. (2001) and Lakota et al. (2002) fit data for holdup and pressure drops in beds packed with Raschig rings, Berl saddles, spheres, cylinders and carbon rings. Their works point to the insensitivity of the relative permeability–saturation curves to the particle shape (Sáez et al., 1986). Therefore, the similar assumption is employed in the present work instead of holdup by void fraction.

$$k_g = k_{gt} = \alpha^n \tag{12}$$

$$\psi_{\rm TP} = \frac{1}{k_g} \left[A \frac{{\rm Re}_g^*}{{\rm Ga}_g^*} + B \frac{{\rm Re}_g^{*2}}{{\rm Ga}_g^*} \right] \tag{13}$$

Sáez and Carbonell (1985) and Levec et al. (1986), for lack of the substantial experimental data for the analysis of gas-phase relative permeability, proposed empirical expressions for the gas-phase relative permeability and the exponent n in Eq. (12) was believed to be practically independent of the particle size, but strongly affected by the particle shape. Lakota et al. (2002) obtained an empirical expression for the gas-phase Reynolds number

$$n = X + 0.0478 \times (\text{Re}_g)^{0.774} \tag{14}$$

The constant X is dependent on the particle shape, as is shown in Table 2, which limits the application of Eq. (14).

3. Experimental apparatus and procedure

3.1. Experimental apparatus

The experimental apparatus is shown in Fig. 1. The primary parts are: a plexiglass test section, a water reservoir, a centrifugal pump and two valves for the control of the mass flow rate. Two Dwyer glass rotameters with different measure ranges are used to measure the gas flow. The accuracy of the two glass rotameters is $0.05 \text{ m}^3/\text{h}$ and $0.03 \text{ m}^3/\text{h}$, respectively. And the Rosemount 3051 water flow meter is used to measure the water flow rate, and its accuracy is $\pm 0.075\%$ of full scale. The MPM498 static pressure transducer has accuracy within $\pm 0.5\%$ of full scale for the measurement of the inlet pressure of the test section. A T-type thermocouple is installed at the inlet of the test section to measure the bulk flow temperature and it has an accuracy of ± 0.1 °C and the measurement repetitiveness is within ± 0.1 °C.

As is shown in Fig. 2, a polycarbonate plate serves as the front plane for the visual observation, a layer of semisphere aluminum particles is carved on the aluminum plate, and another layer of

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Fig. 1. Schematic diagram of the experimental apparatus.



a Overall perspective

b Partial perspective

Fig. 2. Special aluminum plate with particle diameter of 6 mm.

Table 3

Properties of the three different packings.

Case no.	1	2	3
Packing shape		Rectangle	
Dimensions of the j	packing		
Length (mm)	181.90	311.80	361.33
Width (mm)	311.80	57.00	78.00
Height (mm)	3.95	7.90	10.53
d_e (mm)	3.00	6.00	8.00
ε	0.3408	0.3781	0.3976

sphere glass particles which are of the same diameter as the semisphere aluminum particle is filled between the aluminum plate and the polycarbonate plate. Experiments are conducted on three types of test sections with particles of three equivalent diameters (3 mm, 6 mm and 8 mm). The specifications of the three different channels are listed in Table 3.

3.2. Experimental procedures

A predetermined flow rate of water is established to flow through the porous layer first. Thus, the packed bed is initially covered with water to ensure complete wetting. When the liquid flow rate is stable, gas is introduced into the bed. The gas flow rate first increases to a predetermined maximum, and then reduces to a low level. On both the increased and reduced gas flow rate conditions, the pressure drop, the inlet temperature and pressure, the flow rate of gas and liquid are recorded with NI Data Acquisition System. The operating ranges are shown in Table 4.

4. Results and discussion

4.1. Single-phase flow pressure drop

Fig. 3 shows the dimensionless pressure drop Ψ_s vs. the Reynolds number for *s* phase. All data points scatter around a general trend. The coefficients of Ergun-type equation of the present work are presented in Table 5. According to Fig. 3, the Ergun-

Table 4	
Flow parameter range	in the experiment

d _e (mm)	Fluid type	Re_p	
		Min.	Max.
	Gas	42	3226
3.00	Water	286	2210
	Gas/water	521	4194
	Gas	19	1523
6.00	Water	463	4601
	Gas/water	573	9152
	Gas	13	1059
8.00	Water	451	4198

380

14,713

Gas/water



Fig. 3. Experimental single-phase pressure drop vs. Ergun-type equation.

Table 5
Coefficients of Ergun-type equation.

d _e (mm)	Fluid type	D/d _e	Α	В
3.00	Gas Water	2.3267	99.44 143.61	0.0032 0.0773
6.00	Gas Water	2.3125	466.31 1048.99	0.0616 0.2185
8.00	Gas Water	2.3197	526.89 1631.90	0.8451 0.1043

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Fig. 4. Comparison of the coefficients of *A* and *B* of Ergun-type equation between different experimental data.

type equation with the constants recommended by Ergun (A = 180, B = 1.8) disagrees with the experimental data.

Fig. 4 shows the modified Ergun constants *A* and *B* vs. the tubeto-particle diameter ratio, D/d_e proposed by previous studies and by the present work. The coefficient *A*, the pressure drop at low Reynolds number, increases with the decrease of the dimensionless particle diameter. In contrast, the coefficient *B*, the pressure drop at high Reynolds number, shows a reverse trend. As the influence of the wall effect on single-phase pressure drop is quite complicate, especially for small packed beds ($D/d_e < 10$), not only the porosity in the near wall region, but also the other variables should be taken into consideration to reflect the specific characteristic of each beds.

According to Eisfeld's theory, at low Reynolds numbers the effect of the wall friction reaches far into the packed beds, thus, dominating the creeping flow regime. In contrast, at high Reynolds numbers the wall friction is restricted to a small boundary layer, so that the porosity effect becomes prominent. The latter is less pronounced than the friction effect, because the boundary layer is considerably thin in this regime. The influence of the wall manifests itself in two ways: (1) these external boundaries offer an additional resistance due to the viscous friction at the wall (Carman, 1937). The friction, which increases the pressure drop, is not negligible in comparison with that caused by particles owing to the fact that the friction surface of the wall increases relative to the total bed surface corresponding to particles as D/d_p decreases (Nemec and Levec, 2005a); (2) a region of increased void fraction is formed near the wall, which extends approximately for half a particle diameter from the walls into the bed (Thadani and Peebles, 1966; Achenbach, 1995; Eisfeld and Schnitzlein, 2001). This bypass flow formed in this region is a major reason for the severe departures of the experimen-



Fig. 5. Schematic diagram of two-regime model.

tal results. The smaller the tube-to-particle diameter ratio, D/d_e is, the stronger is the influence on the pressure drop (Fand and Thinakaran, 1990; Achenbach, 1995). By analyzing the wall effect wall, the bed porosity and the particle shape, Nemec and Levec (2005a) proposed a correlation for $D/d_e > 10$. For the improvement in the prediction of the pressure drop at low tube-to-particle diameter ratios, the observed wall effect is taken into account in dealing with the respective coefficients A and B. In order to find the most promising approach, Eisfeld and Schnitzlein (2001) compared 24 published correlations, and concluded that the most promising one was that by Reichelt (1972). In the equation, the contribution of the confining walls to the hydraulic diameter is analyzed by the wall effect coefficient A_w . And the function B_w describes empirically the porosity effect of the wall at high Reynolds numbers (Eisfeld and Schnitzlein, 2001). The average porosity is employed to represent the porosity of the packed beds. For the packed beds where the wall effect is negligible $(D/d_e > 10)$, this assumption is justifiably right. However, when the beds are strongly affected by the wall $(D/d_e < 10)$, the nonuniformity of the porosity is to be taken into consideration.

For spherical particles, Benenati and Brosilow (1962) and Goodling (1983) measured the radial bed porosity distribution. The porosity decreases from the unity at the wall to a minimum at a distance of $d_e/2$ and then levels out with decreasing amplitudes to a constant value. This point is reached at a distance of nearly four sphere diameters from the wall (Achenbach, 1995). Similar experiments on the bed porosity distribution were conducted for other particle shapes used in catalyst techniques. And the similar conclusion was obtained (Achenbach, 1995; Roblee et al., 1958). For a preliminary analysis, the bed is subdivided into a near-wall region and a central region as is shown in Fig. 5. In the central region, the wall effect is neglected and the average porosity is employed to represent the porosity of the packed beds. As a result, in the central region, A_w and B_w equal to 1.

The following expression for the porosity of sphere particle bed can be derived through a simple geometrical model.

$$\varepsilon = a_0 d_{\rho}^{*2} + b_0 \tag{15}$$

where $a_0 = 0.78$ and $b_0 = 0.375$ (Carman, 1937). While in the near-wall region, the bed porosity, ε_w , is closely related to the characteristics of the packed beds (Achenbach, 1995).

$$\varepsilon_w = 63.6(d_e^{*-1} + 15)^{-2} + 0.43 \tag{16}$$

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Fig. 6. Comparison of the present model for coefficients of A_w and B_w of Ergun-type equation for single-phase flow pressure drop with different experiment data.

Based on the work of Reichelt (1972) and Eisfeld and Schnitzlein (2001), the wall effect coefficient, A_w and B_w , are expressed as follows

$$\psi_{s} = A_{0}A_{w}\frac{\mathrm{Re}_{s}^{*}}{\mathrm{Ga}_{s}^{*}} + B_{0}B_{w}\frac{\mathrm{Re}_{s}^{*2}}{\mathrm{Ga}_{s}^{*}}$$
(17)

In the near-wall region,

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$$A_{w} = \left[a + b\frac{d_{e}^{*}}{(1 - \varepsilon_{w})}\right]^{2}$$
(18)

$$B_{\rm w} = \left(cd_e^{*-2} + d\right)^{-2} \tag{19}$$

$$d_e^* = \frac{d_e}{D} \tag{20}$$

And in the central region,

$$A_W = B_W = 1 \tag{21}$$

In the above equations $A_0 = 180$, $B_0 = 1.8$, a = 0.8, b = 2, c = 3 and d = 1; d_e^* is the dimensionless particle diameter; *D* represents the hydraulic diameter of the packed beds.

For the small packed beds $(D/d_e < 10)$, as no central region exits, the wall effect coefficient, A_w and B_w , are determined by with Eqs. (18) and (19). While for the large packed beds $(D/d_e > 10)$, compared with the central region, the near-wall region can be neglected and A_w and B_w are determined by Eq. (21). The improved correlations shown in Fig. 6 are applicable for particles of spheres and cylinders and for other particles regardless of their shapes by fitting its coefficients to the given experimental database. The precision of this



Fig. 7. Comparison of the present model for two-phase flow pressure drop with the present experiment data.

correlation is influenced by the accuracy of the average porosity in the near-wall region, which is hard to given a uniform expression for various types of packed beds.

4.2. Two-phase flow pressure drop

Sáez et al. (1986) used the conduit models to calculate the gas phase relative permeability. It was found that for gas phase Reynolds numbers below a certain value, the gas phase relative permeability-saturation curves, dependent on the liquid Reynolds number, were a single function of the ratio of Reynolds to Galileo numbers of the gas. For large values of the gas phase Reynolds number, all the relative permeability curves converge to a single function of saturation (Sáez et al., 1986). If a fit similar to Eq. (12) was performed to the results obtained with the conduit models, n is a function exclusively of $\operatorname{Re}_g^*/\operatorname{Ga}_g^*$ (Sáez et al., 1986; Levec et al., 1986). Sáez and Carbonell (1985) and Levec et al. (1986) concluded that the exponent *n* is a function of the gas phase Reynolds number. However, in the present experiment, *n* proves to be correlated with the void fraction α , as is shown in Fig. 7, which is more conclusive than the previous conclusion since the influence of the two-phase are taken into consideration. The expression of *n* obtained by the present experiment is

$$n = 7.185\alpha + 1.41 \tag{22}$$

where the exponent n is a linear function of the void fraction.



Fig. 8. Comparison of the present model for two-phase flow pressure drop with the previous experimental data.

The correlation is compared to the previous experimental data of Sáez and Carbonell (1985), Levec et al. (1986), Nemec et al. (2001) and Lakota et al. (2002), as shown in Fig. 8. The correlation results in a mean relative deviation of $\pm 16\%$ in the prediction of pressure drops for 92 experimental data about the spherical particles and that of $\pm 23\%$ in the prediction of pressure drops for 33 data about the nonspherical particles (Rasching rings, Checked Marbles and Cylinders). Compared with the relative error of the four groups of models shown in Table 1, the present model is one of the models with the smallest mean relative error.

5. Conclusions

In finite packed beds the additional influence of the confining walls must be accounted for. In the present flow-regime model, the bed is subdivided into a near-wall region and a central region. Improved correlations are obtained based on the previous study to consider the single-phase flow pressure drops for finite pebble beds with spherical particles and nonspherical particles by fitting the coefficients of that equation to both the database and present experiment. The correlation is consistent with the observed physical behavior which explains its comparatively good agreement with the experimental data. The correlation can be used to predict the single phase pressure drop for both great tube-to-particle diameter ratio packed beds ($D/d_e > 10$) and small tube-to-particle diameter ratio packed beds ($D/d_e < 10$) by employing porosity correlation corresponding to the packed beds. The correlation is compared with the previous experimental data and it fits well for both experimental data of spherical particles and nonspherical particles. As the influence of the wall effect on single-phase pressure drop is quite complicate, especially for small packed beds ($D/d_e < 10$), future study is necessary for the accurate predict of the single-phase pressure drop with this model.

A new correlation for the prediction of two-phase flow pressure drops is proposed based on the gas phase relative permeability as a function of the gas phase saturation and the void fraction. The correlation is compared with the previous experimental data and the present experimental data. The correlation fits well for both the experimental data of spherical particles and nonspherical particles.

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