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Original Article

An experimental study on two-phase flow resistances and interfacial drag in packed porous beds

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ABSTRACT

Motivated by reducing the uncertainties in quantification of debris bed coolability, this paper reports an experimental study on two-phase flow resistances and interfacial drag in packed porous beds. The experiments are performed on the DEBECO-LT (DEbris BEd COolability-Low Temperature) test facility which is constructed to investigate the adiabatic single and two phase flow in porous beds. The pressure drops are measured when air-water two phase flow passes through the porous beds packed with different size particles, and the effects of interfacial drag are studied especially. The results show that, for two phase flow through the beds packed with small size particles such as 1.5 mm and 2 mm spheres, the contribution of interfacial drag to the pressure drops is weak and ignorable, while the significant effects are conducted on the pressure drops of the beds with bigger size particles like 3 mm and 6 mm spheres, where the interfacial drag in beds with larger particles will result in a descent-ascent tendency in the pressure drop curves along with the fluid velocity, and the effect of interfacial drag should be considered in the debris coolability analysis models for beds with bigger size particles.

1. Introduction

Single and two-phase gas/liquid flow in packed porous media occur in many engineering applications, ranging from agricultural, biomedical, mechanical, chemical and petroleum engineering to food industry [1,2]. Specifically, during a severe accident of a light water reactor (LWR) with failure of cooling systems, a porous debris bed may be formed when melt corium relocates to a water pool in the lower head or in the cavity. The coolability of the debris bed therefore is of great importance in corium risk quantification, which is crucial to the stabilization and termination of a severe accident in LWR. Towards quantitative understanding of debris bed coolability, numerous experiments [3–9] have been carried out and a great number of analytical models and empirical correlations [10–17] are proposed, the central point is to provide the formulation of the friction laws for momentum equations of single and two-phase flow in the particulate beds. It is now generally accepted that satisfactory predictions of pressure drop of single phase flow in packed spheres beds can be achieved by the simple semi-empirical models like the Ergun equation [10].

$$-\frac{dp}{dz} = \frac{\mu}{K}J + \frac{\rho}{\eta}J^2 = \frac{150(1-\varepsilon)^2\mu}{d_p^2\varepsilon^3}J + \frac{1.75(1-\varepsilon)\rho}{d_p\varepsilon^3}J^2 \tag{1}$$

where dp/dz is the pressure gradient along the height of the bed, the first term of the right side is the viscous loss (proportional to velocity) and the second term is the inertial loss (proportional to velocity squared). μ is the dynamic viscosity of fluid, ρ is the density, J is the superficial velocity of fluid, the parameters K and η are called permeability and passability, respectively. 150 and 1.75 are called the Ergun constants, d is the diameter of particles, and ε is the bed porosity.

Contrary to single-phase flow, there exist a good number of models and correlations [11–17] to assess the pressure drops of two-phase flow in porous media, and their predictions are quite scattering [8]. Equation (2) shows the general expressions of some models and Table 1 lists the related parameters proposed by different researchers.

$$-\frac{dp_l}{dz} = \rho_l g + \frac{\mu_l}{K \cdot K_{r,l}}J_l + \frac{\rho_l}{\eta \cdot \eta_{r,l}}J_l \cdot |J_l| - \frac{F_i}{1-\alpha} \tag{2a}$$

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Table 1
Different models and their parameter expressions.

Models	Flow patterns	K_{rg}	η_{rg}	K_{rl}	η_{rl}	F_i
Lipinski [11]	All	α^3	α^3	$(1-\alpha)^3$	$(1-\alpha)^3$	0
Reed [12]	All	α^3	α^5	$(1-\alpha)^3$	$(1-\alpha)^5$	0
Hu & Theofanous [13]	All	α^3	α^6	$(1-\alpha)^3$	$(1-\alpha)^6$	0
Schulenberg & Műller [14]	All	α^3	$\alpha^6,\alpha > 0.3$; $0.1\alpha^4,\alpha \le 0.3$	$(1-\alpha)^3$	$(1-\alpha)^5$	Eq.(3)
Tung & Dhir [15]	Bubble flow & Slug flow	$\left(\frac{1-\varepsilon}{1-\varepsilon\alpha}\right)^{\frac{4}{3}}\alpha^4$	$\left(\frac{1-\varepsilon}{1-\varepsilon\alpha}\right)^{\frac{2}{3}}\alpha^4$	$(1-\alpha)^4$		Eq.(4)
	Annular flow	$\left(\frac{1-\varepsilon}{1-\varepsilon\alpha}\right)^{\frac{4}{3}}\alpha^3$	$\left(\frac{1-\varepsilon}{1-\varepsilon\alpha}\right)^{\frac{2}{3}}\alpha^3$	$(1-\alpha)^4$		Eq.(8)
Schmidt [17]	Bubble flow & Slug flow	$\left(\frac{1-\varepsilon}{1-\varepsilon\alpha}\right)^{\frac{4}{3}}\alpha^4$	$\left(\frac{1-\varepsilon}{1-\varepsilon\alpha}\right)^{\frac{2}{3}}\alpha^4$	$(1-\alpha)^4$		Eq.(4)
	Annular flow	$\left(\frac{1-\varepsilon}{1-\varepsilon\alpha}\right)^{\frac{4}{3}}\alpha^3$	$\left(\frac{1-\varepsilon}{1-\varepsilon\alpha}\right)^{\frac{2}{3}}\alpha^3$	$(1-\alpha)^4$		Eq.(9)

$$-\frac{dp_g}{dz} = \rho_g g + \frac{\mu_g}{K \cdot K_{r,g}} J_g + \frac{\rho_g}{\eta \cdot \eta_{r,g}} J_g \cdot \left|J_g\right| + \frac{F_i}{\alpha} \tag{2b}$$

where l and g represent the liquid and gas phases respectively, and the parameters K_r and η_r are called relative permeability and relative passability respectively, F_i is called interfacial drag, α represents void fraction. Clearly, the total pressure drop consists of three terms: gravity force term, fluid-particles drag term and interfacial drag term. It can be seen from Table 1 that these models can be divided into two groups generally, according to whether the researchers considered the interfacial drag or not.

By introducing the relative permeability K_r and relative pass ability η_r, Lipinski [11] extended the Ergun equation [10] to the case of two-phase flow through the particulate beds. Such approach was also adopted in Reed model [12] and Hu & Theofanous [13], but different parameters were used. It should be noted that interfacial drag did not take account in These models ($F_i = 0$).

Tutu et al. [16] stressed the importance of gas-liquid interfacial drag of flow in porous beds with coarse particles, especially for large particle sizes ($d_p \geq 6$ mm), where the gas-liquid interfacial drag can be comparable with the gas-solid drag and can't be neglected. Based on measured experimental data and corresponding analysis, Schulenberg & Műller [14] took account of the interfacial drag between liquid and gas (related with buoyancy force, viscous force, inertial force and capillary force), which was expressed as following form.

$$F_i = 350(1-\alpha)^7 \alpha \frac{\rho_l K}{\eta\sigma}\left(\rho_l - \rho_g\right) g \left(\frac{J_g}{\alpha} - \frac{J_l}{1-\alpha}\right)^2 \tag{3}$$

Here σ is the surface tension.

One step further, Tung & Dhir [15] developed a hydrodynamic model including interfacial drag, based on flow regimes and their relationship with flow and porous layer configuration, to predict void fraction and pressure drops for two-phase flow through porous media. Depending on visual observation, they defined three flow patterns: bubble flow, slug flow and annular flow. The expressions of interfacial drag F_i were proposed for different flow patterns. For bubble and slug flow, the interfacial drag F_i was deduced as the following form:

$$F_i = C_1 \frac{\mu_l}{D_b^2 \varepsilon}(1-\alpha) J_r + C_2 \frac{(1-\alpha)\rho_l + \alpha\rho_g}{D_b^2 \varepsilon^2}(1-\alpha)^2 J_r . |J_r| \tag{4}$$

Where J_r is the relative velocity, D_b is the bubble diameter defined by

$$J_r = \frac{J_g}{\alpha} - \frac{J_l}{1-\alpha},\ D_b = 1.35\sqrt{\frac{\sigma}{\left(\rho_l - \rho_g\right) g}} \tag{5}$$

and the friction coefficients were given separately for bubble flow and slug flow

$$\text{Bubble flow } (0 < \alpha < \alpha_1): C_1 = 18\alpha, \quad C_2 = 0.34\alpha(1-\alpha)^4 \tag{6}$$

$$\text{Slug flow } (\alpha_2 < \alpha < \alpha_3): C_1 = 5.21\alpha, \quad C_2 = 0.92\alpha(1-\alpha)^4 \tag{7}$$

The interfacial drag for annular flow was expressed as:

$$F_i = \frac{\mu_g}{K . K_{r,g}}(1-\alpha) J_r + (1-\alpha)\alpha \frac{\rho_g}{\eta . \eta_{r,g}} J_r . |J_r| \tag{8}$$

Schmidt [17] modified Tung & Dhir model [15] by revising some expressions of parameters, such as the diameter of gas bubbles or slugs, the flow pattern bounds and the interfacial drag in annular flow, which all mainly affects the formulation of interfacial drag. For Annular flow, the interfacial drag F_i can be deduced as the following form:

$$F_i = \frac{\mu_g}{K K_{r,g}}(1-\alpha) J_r + (1-\alpha)\alpha \frac{\rho_g}{\eta\eta_{r,g}} |J_r| J_r \times \begin{cases} \left(\frac{d_p}{6}\right)^2 : d_p < 6\text{mm} \\ 1 : d_p > 6\text{mm} \end{cases} \tag{9}$$

From the models discussed above, one can see that the key point in the models is to provide the formulation of the friction laws for momentum equations of two-phase flow in porous beds, since it is believed that the debris coolability is mainly restricted by hydrodynamic limitations of two-phase flow through the debris bed [18]. However, some of the key parameters in above equations such as K_r and η_r are given different expressions by different researchers. What's more, the effects of interfacial drag F_i on the pressure drops of two phase flow are still unsettled. Therefore even for the same flow conditions, the calculated results by different models are different due to the inconsistent parameters. Recent work from Chikhi et.al (2016) [8] stated that there is no definitive conclusion on this subject by now. In order to verify debris coolability analytical models, and to better understand the effect of interfacial drag, experiments are conducted to study the flow characteristics of particulate beds with different sizes particles are performed in the

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