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A ONE DIMENSIONAL TWO-FLUID MODEL FOR BUBBLY FLOWS IN FIXED BEDS

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ABSTRACT

Pressure drop and gas void fraction are important parameters for the design of multiphase packed bed reactors which are widely used in petrochemical industry. Several experimental studies have been devoted to the hydrodynamics of two-phase cocurrent upflow or downflow through fixed beds, and various correlations of limited range of validity are available in the literature. However, there is not yet a clear agreement on the form of the momentum equations to be used in such systems. Early attempts devoted to the pressure drop estimate were based on an extension of the Lockhart-Martinelli approach (Sweeney 1967), Rao et al. 1983). More recently, Attou at al. (1999) proposed the first serious attempt to adapt the Eulerian two-fluid model to cocurrent bubbly flows through packed beds. From an analysis of their proposal, it happens that the basic mechanical equilibrium for the gas phase needs to be reconsidered. In this scope, we derived a new model on the basis of the so-called hybrid approach initially developed for bubbly flows in ducts in absence of shear-induced turbulence (Achard and Cartellier 2000). As a first application, we considered a mean unidirectional flow of a bubbly mixture through a porous medium composed of beads uniform in size. For steady and fully established flows, and assuming a flat void fraction (α) profile, the resulting momentum equations for each phase write:

Liquid phase : $$-\frac{dp}{dz} = \rho_L g + f_{LS} - \frac{f_{LG}}{1-\alpha} \quad (1)$$

Gas phase : $$-\frac{dp}{dz} = \rho_G g + f_{LS} + \frac{f_{LG}}{\alpha} \quad (2)$$

where f_{LS} is the resultant of the liquid shear stress exerted on beads surface and on exterior walls, and where the quantity $f_{LG} = \alpha F^* / V_p$ represents the interaction force density between the gas and the liquid (F^* is the mean force on bubbles and $V_p = 4\pi a^3/3$ denotes the bubble volume, a being the bubble radius). The main difference with the model derived by Attou et al. is the presence of the f_{LS} term in the gas phase equation. Without this term, the relative velocity of bubbles would be controlled by the axial pressure gradient dP/dz even in non accelerating flows which is unphysical. On the opposite, in the present model (1-2) the relative movement of bubbles is simply due to buoyancy. The set of equations (1-2) provides a mean to exploit the experimental data to derive the required closures, namely the evolution of the friction f_{LS} with the gas content and that of the momentum exchange between phases f_{LG}. Notably, from (1) and (2), one gets

$$f_{LG} = \alpha(1-\alpha)(\rho_L - \rho_G)g \quad (3)$$

In order to establish reliable closures, available experimental data of the literature are currently revisited under this framework. For the friction term, which is the principal contribution to the pressure drop, the usual closure law for f_{LS} as given by an Ergun equation adapted to two-phase flows is under analysis. For the interfacial momentum transfer, the objective is to evaluate an "apparent" drag coefficient defined

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as $C_d = F^*/[\rho_L U_r^2 \pi a^2 / 2]$ where the mean relative velocity U_r is defined as the difference between the mean gas and liquid velocities averaged over a volume. Indeed, paralleling an approach already exploited for bubbly flows in ducts (Rivière and Cartellier 1999), it happens that the mean void fraction can be derived from equations (1) and (2) assuming a flat void fraction profile :

$$\frac{\beta}{(1-\beta)} - \frac{\alpha}{(1-\alpha)} = (4\pi/3)\,\alpha\,(1-\alpha)\,\frac{\left[\frac{g\,\delta^2}{V_{SL}\,\nu_c}\right]\left(\frac{a}{\delta}\right)^2}{f_d} \quad (4)$$

where δ is the typical size of the pores and where $f_d = (\pi/2)\,Re_p\,C_d$ is expected to be a function of the bubble size, the porosity ε and the void fraction. To extract f_d or C_d from (4), a characteristic bubble size must be specified. As shown Fig.1, the bubble size is controlled by the bed geometry and evolves between 0.2 δ and 3 δ in the dilute limit (Bordas et al. (2001)). Analysis of the existing data will be presented based on these size estimates, and comparison will be performed of this "apparent" drag with values measured for isolated bubbles in fixed beds (Fig.2).

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