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Residence time distribution, axial liquid dispersion and dynamic–static liquid mass transfer in trickle flow reactor containing β-SiC open-cell foams

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ABSTRACT

Open-cell foams are a promising new support for gas–liquid contacting and heterogeneous catalysis. Using the Piston-Dispersion-Exchange model, axial dispersion and dynamic–static liquid mass transfer coefficient are estimated from residence time distribution responses measured under co-current down flow configuration.

The specific morphological parameters and the high porosity involved within open-cell foams result in low Bodenstein numbers ($0.05 < Bo < 0.2$ for $2.5 < Re < 20$) in comparison to conventional packed beds ($0.2 < Bo < 0.7$ for $5 < Re < 23$). Furthermore, open-cell foam improves liquid mass transfer coefficient between the dynamic and the static zones under trickle flow conditions.

1. Introduction

Solid open-cell foams constitute a class of porous materials with low density and improved thermal, mechanical, electrical and acoustic properties [1]. They have been used for a long time in designing aircraft wing structures in the aerospace industry, core structure for high strength panels, and also in compact heat exchangers. More recently, open-cell foams (with or without carbon nanofibers decorating superstructure) have been considered as potential candidates for catalyst support in the heterogeneous catalysis field due to their high external surface area combined with a low pressure drop [2–5]. Actually, these structured supports are expected to bring a new way to intensify process operations involving multiphase flows [6], such as fuels synthesis [7,8] and reactive distillation [9,10].

Up to date, several reports have been published in the literature dealing with gas–liquid flow, liquid holdup, and pressure drop and transport properties behavior on different commercial open-cell foams [11–21]. These studies clearly showed that the frictional pressure drop measured through open-cell foam beds is always small in comparison to classical packed beds (especially in comparison to the spherical packed bed). It is also shown that the liquid holdup within open-cell foams remains higher than the one measured for other packed beds under similar reaction conditions [22].

A large variety of open-cell foams with different intrinsic properties are available nowadays for several applications. The previous applications of silicon carbide open-cell foam with a cubic structure (β-SiC) in a fixed bed configuration for multiphase reactors have shown very satisfactory results in terms of reaction selectivity [23] and transport properties [10]. Therefore, investigation of transport phenomena within β-SiC open-cell foams remains an important field of fundamental research and the results can be extrapolated to other kind of open-cell foams as a function of the downstream application. In this context, the present paper investigates the axial liquid dispersion and mass transfer, under trickle flow conditions, for different β-SiC foams.

In the most common case, axial dispersion is studied by injecting a tracer, as a single pulse on the whole cross section (or upstream) of the packed bed, and then measuring the outlet response. In this work, the experimental data reported by Edouard et al. [18] are coupled with the Piston-Dispersion-Exchange model (PDE, presented by Iliuta et al. [24,25]) in order to estimate the axial dispersion coefficient in dynamic liquid zone and the mass transfer between the dynamic and static zone. The residence time distribution curves are obtained in co-current down flow trickle bed reactor filled with spherical particles or β-SiC open-cell foams. Finally, the results obtained in the present work are compared with those available in the literature.

2. Experimental

2.1. Support characteristics

The open-cell silicon carbide foams (Fig. 1) used in this work are manufactured by Sicat Company (www.sicatcataly.com). The several packed bed morphological parameters are measured and

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Nomenclature

Bo	Bodenstein number for liquid axial dispersion
D_{ax}	dispersion coefficient, $m^2\ s^{-1}$
d_{peq}	equivalent particle diameter, m
$E(t)$	residence time distribution function
Ga	Galileo number of liquid phase
H	length of the packed bed, m
ka	mass-transfer coefficient between the liquid zones, s^{-1}
Re	Reynolds number of liquid phase referred to superficial velocity
t	time, s
U_L	liquid superficial velocity (based on an empty cross section of the reactor), $m\ s^{-1}$
U_g	gas superficial velocity (based on an empty cross section of the reactor), $m\ s^{-1}$

Greek symbols

ε_{LT}	total liquid holdup
ε_{Ld}	dynamic or free-draining liquid holdup
ε_{Lst}	static liquid holdup
ε	overall porosity of the packed bed
μ_L	liquid dynamic viscosity, Pa s
ρ	liquid density, $kg\ m^{-3}$

compared with values given by Sicat Company (Table 1). The specific surface area of the sample is estimated with equation proposed by Lacroix et al. [26].

2.2. Reactor residence time distribution and liquid holdup measurements

The hydrodynamic characterization of the flow pattern through open-cell foam in a packed bed configuration in the gas–liquid reactor is performed by tracing the liquid phase. The reactor has an inner diameter of 0.037 m and a total height of 0.93 m. It was filled with 0.003 m spherical glass particles or β-SiC open-cell foam to reach a total bed height of 0.9 m. Gas (air) and liquid (water) were injected through a spray shower on the top of the reactor in order to ensure a homogeneous fluid distribution at the reactor inlet. The gas flow rate was measured by a flowmeter. Different gas flow rates were studied for different superficial liquid velocities. These fluid velocities are based on the empty reactor cross section and varied between $7.7 \times 10^{-2}\ m\ s^{-1}$ and $22 \times 10^{-2}\ m\ s^{-1}$ for air and $1.6 \times 10^{-3}\ m\ s^{-1}$ and $7.8 \times 10^{-3}\ m\ s^{-1}$ for water. Using a KCl solution as a flow tracer, conductivity measurements were recorded by cell conductivity (Omega model CDCN-91) in order to get access to the residence time distribution related to the different packed bed configurations. A short tracer pulse was injected in the inlet. The first conductivity cell was placed on the top of the reactor to measure the inlet pulse injection and a second one was placed 0.03 m above the bottom of the reactor to measure the outlet response. In order to ensure the experiment reproducibility and the complete wetting of the packed bed, the column was first completely filled with liquid and fully drained. Then, the required gas and liquid flow rates were set and the column regime was stabilized for about 20 min under dynamic flow before starting the measurements.

The liquid holdup is the fractional volume of the fixed bed that is occupied by the liquid phase. In general, the total liquid holdup is divided into two phases (Eq. (1))

$$\varepsilon_{LT} = \varepsilon_{Ld} + \varepsilon_{Lst} \tag{1}$$

where ε_{Ld} and ε_{Lst} denote, respectively, the dynamic component and the static component.

The dynamic liquid holdup is the free flowing liquid fraction whereas the static liquid holdup is the stagnant liquid fraction retained by capillary forces on and inside the solid structure after column draining. These parameters depend on the operating conditions and the geometrical properties of the packed bed. The total liquid holdup (ε_{LT}) in the column is evaluated using residence time distribution (RTD) analysis and the dynamic liquid holdup (ε_{Ld}) is obtained by a drainage method (see for instance Yang et al. [27]) with an experimental uncertainty of about ± 0.01.

The static liquid holdup values are obtained by Eq. (1) and are assumed to be constant during experimental operation [18,27–30]. The main results obtained in terms of total and dynamic liquid holdup are reported in Table 2.

For diluted solutions, the KCl concentration is linearly proportional to the measured conductivity. The conductivity signals at both the inlet and the outlet are normalized to yield normalized concentrations. Then, residence time distribution function $E(t)$ can be calculated from the conductivity data according to Eq. (2).

$$E(t) = \frac{I(t)}{\int_0^\infty I(t)\ dt} \tag{2}$$

where $I(t)$ denotes the solution conductivity.

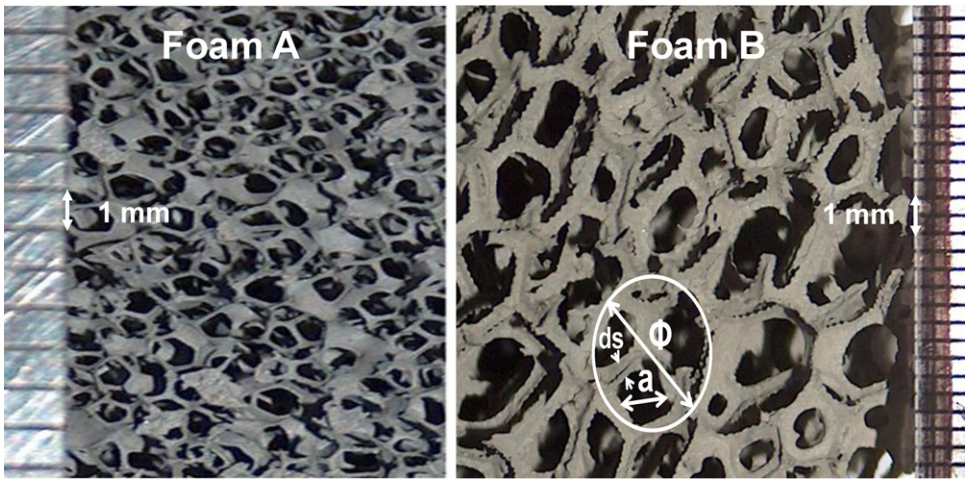

Fig. 1. β-SiC open-cell foams and the characteristic lengths involved within an elementary cell.