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Gas/liquid dispersions with a SMX static mixer in the laminar regime

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Abstract

A Sulzer SMX mixer was used to disperse gas into viscous, Newtonian and non-Newtonian fluids. The investigation covered the effect of the dispersed phase volume fraction, the viscosity of the continuous phase, the mixer length and the power draw. The flow regime was kept laminar in all the experiments. The dispersion of gas was carried out with gas concentrations between 1% and 7% in volume. Using the "process viscosity" concept, it was possible to collapse all the measured sizes on a single master curve by using the energy consumption in the mixer as the common variable between the experiments. Comparison was made with a Kenics mixer. The SMX mixer was found to be better adapted to the dispersion task due to its internal structure.

Keywords: SMX static mixer; Multiphase flow; Gas–liquid dispersion; Laminar flow; Newtonian; Non-Newtonian

1. Introduction

The process of dispersing a fluid (liquid, gas) or a solid into a liquid continuous phase is probably one of the most widely spread mixing process in the industry. Despite this popularity, the equipment used for dispersion in continuous laminar processes are not that well known and the knowledge is still, in large parts, owned by the equipment builders. There exist a certain number of studies reporting capacity augmentation, higher process efficiencies, or reaction yield when using static mixing devices with two or more fluid phases in laminar regime (e.g., Schneider et al., 1988; Stringaro and Luder, 1991; Yeon and Choi, 1996). This work is aimed at quantifying the dispersion capacity of the SMX mixer (Sulzer) in the laminar regime and compare it to the performance of other mixers.

Most of the work dealing with laminar dispersion whether is based on Grace (1982) or makes direct references to it. The center of his work is the variation of the critical capillary number for dispersion of drops as a function of the viscosity ratio of the phases. The capillary number is defined as the ratio of the Weber number to the Reynolds number namely:

$$Ca = \frac{\eta u}{\sigma} = \frac{\tau}{(\sigma/L)}. \quad (1)$$

From the dispersed phase point of view, this number can be seen as the ratio of the "destructive" viscous forces responsible for the rupture of the secondary phase over the "protective" surface tension forces opposing the deformation. The inertial forces do not dominate due to the laminar flow regime. In this expression, η is the continuous phase viscosity, u a characteristic velocity and σ is the interfacial tension. The results of Grace were generated for a simple shear flow (Couette apparatus) and for a 2D extensional shear field (four rolls). All these results were obtained for one drop in steady state deformations, which necessitate extremely long times to reach equilibrium. While being very instructive about the break-up mechanisms fundamentals, these experiments do not represent industrial static mixer conditions. Static mixers intrinsically generate transient conditions while the capillary number study by Grace made use of steady state conditions only. The fact that only one drop is studied is also a strong deviation from the industrial situation where billions of drops can be present and interact with each other.

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At lower viscosity ratios $p < 0.1$ ($p = \eta_{\text{dispersed}}/\eta_{\text{continuous}}$), in a simple shear field, very high stresses must be generated in order to break a bubble or a drop. With viscosity ratios above 3.0, steady simple shear flows cannot generate stresses high enough to lead to break-up. With extensional shear flows however, the capillary numbers remain within reasonable values for a very wide range of viscosity ratios. According to these results, flow fields with strong extensional components are preferred for dispersion applications (Rallison, 1984).

The flow type can be defined as the ratio of the deformation (D) to the rotational component (Ω) of the tensor:

$$\alpha = \frac{|D|}{|D| + |\Omega|} \quad \text{where } D + \Omega = \frac{1}{2}\left[(\nabla v + \nabla v^{\mathrm{T}}) + (\nabla v - \nabla v^{\mathrm{T}})\right], \tag{2}$$

α is helpful in representing the type of two-dimensional flow used: 0 = pure rotational; 0.5 = simple shear; 1 = pure extensional. Depending on the complexity of the geometry used, any intermediate value of α is possible. While this parameter is not objective, that is it cannot be used to compare totally different geometries such as an agitated tank and a static mixer (Astarita, 1979), it is still useful to compare similar geometries like two different static mixer designs or two mixing tanks against each other (e.g., De La Villéon et al., 1998).

The viscosity of the dispersed phase is a major parameter in the dispersion phenomenon. In the turbulent regime, it has been shown that the width of the bubbles/drops distribution increases with the viscosity ratio (Berkman and Calabrese, 1988; Middleman, 1974). Recently, there is some work reporting dispersion results in the turbulent regime (Legrand et al., 2001) using static mixers. The Reynolds number ($Re = \rho \bar{v} D/\mu$) is part of every size correlation in this regime, generally of the form: $D = D(Re^a)$. The exponent a ranges between 0.1 and 0.2 with variations due to the mixer type.

The laminar regime was theoretically treated by using the energy repartition within a swirl (Middleman, 1974). At this length scale, the drop environment is limited to the movement inside a single swirl and is therefore laminar. The relation is of the form

$$\frac{D_{32}}{D_t} = C_2 \left(\frac{\sigma}{\mu v}\right)^3 \frac{D v \rho}{\mu}. \tag{3}$$

According to this relation, the average diameter D_{32} is inversely proportional to the viscosity of the dispersed phase and the average velocity of the flow ($1/\mu^4$ and $1/v^2$). The interfacial tension also plays a major role (σ^3). There are no available experimental data to validate this theory.

It is quite clear that the dispersion process is rather complex and it is also evident that the laminar regime is the least popular candidate for this type of process. To our knowledge, there exists only one work on laminar dispersion in a static mixer (Grace, 1982) in which only the Kenics mixer was considered.

The behavior of the size reduction in the mixer can be explained by a kinetic phenomenon. With the dispersed phase passing through the obstacles formed by the element, the drops break in a repetitive process until the kinetic equilibrium is reached between breakup and coalescence. The point where this equilibrium is reached is a function of the dissipated energy and the liquid or gas system involved. The mixing length to establish the equilibrium can be directly compared to the mixing time in tanks (Meyer et al., 1988; Villermaux and Falk, 1994). During the design process of a new application, it is hence of prime importance to evaluate the required number of elements properly.

Table 1
Experimental values of constants C_1, C_2 and C_3 (Eq. (4) with conditions

	Middleman	Al-Taweel
C_1	0.95	1.0
C_2	5.1	4.9
C_3	0.40	0.19
Mixers	3, 6, 21 Kenics	4, 8, 12, 16, 24, 36 in-liner
Fluids	Benzene in water	Kerosene in water
Volume fraction	0.5 à 1%	1%
Interfacial tension	4 mN/m	—

Grace, in addition to the investigation of the break-up of single drops, used static mixers in his experiments with unsteady flows. His objective was to build a parallel between abrupt changes in flow conditions and the repetitive transient conditions met in static mixers. He showed that variations in the shear rate could lead to the breakup of drops and bubbles without even reaching the critical capillary number. This finding is very instructive about the capacity of a static mixer to generate an efficient size reduction.

The effect of the number of mixing elements can be modeled with the following (Middleman, 1974):

$$\frac{D_{32}}{D_{32(N\ \max)}} = C_1 + C_2\, \mathrm{e}^{(-N^* C_3)}. \tag{4}$$

Here, the value of $D_{32(N\max)}$ is simply the minimum size achieved with the maximum number of elements used in the experiments. The values of constants C_1–C_3 have been calculated here based on the results generated in a Kenics (Middleman, 1974) and an in-liner mixer from Lightnin (Al-Taweel and Walker, 1983). They are presented in Table 1. The in-liner mixer is very close in design to the Kenics mixer with their difference laying in the lamellae that are composed of small and partial twisted lamellas instead of an entire 90° or 180° twist. The values of 0.19 vs. 0.40 shown in the same table are also of significant importance for their dispersion capacity characterization and will be discussed later in this paper along with the SMX results.

Based on the principle that the Kenics mixers leads to an interface generation proportional to 2^N (N the number of elements) within the laminar flow regime, an experimental investigation shows that this relation is true only after a minimum number of elements (Bigio and MacLaren, 1991). This minimum number of elements would increase with the volume fraction of dispersed phase and the viscosity ratio. Neither details nor data are provided in the paper.

The dispersion of a fluid into a second one requires some sort of an injection point where the first contact occurs between