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Dispersion in fully developed flow through regular porous structures: Experiments with wire-mesh sensors

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ABSTRACT

Within this study the wire-mesh sensor is proposed as a suitable device to investigate radial and axial dispersion in tubular reactors. Axial dispersion in turbulent flow through a regular highly porous structure is addressed and the effect of reactor length on the estimated axial dispersion coefficient is discussed. We state that the gradual increase of turbulence intensity in the entrance section of the porous structure is an effect which leads to a length dependence of the dispersion coefficient. Furthermore, we present measurements of radial mixing by the wire-mesh sensor and compare it to reference measurements using laser induced fluorescence. By that we discuss the possible bias from non-homogeneous tracer distribution between the electrodes. Despite the differences of the measurement principles we found a good qualitative agreement between the results.

1. Introduction

The potential to achieve constant product quality with increased safety compared to batch reactors led to a new packed tubular reactor approach which is produced by selective laser sintering [1]. The reactor has an inner diameter of 7 mm and contains a packing with a porosity of 84%. This allows to run chemical reactions with reasonable high flowrates at comparable low pressure drop. A challenge for the design of continuous reactors is the characterization of the residence time distribution (RTD).

The RTD can be measured by tracer pulse experiments. Depending on the application, authors used various different tracers like radioactive substances to measure in pipelines [2] and fluorescent dyes to measure in microreactors [3]. Often used tracer substances are aqueous salt solutions which can be tracked by their conductivity. We decided to use potassium chloride as tracer in water in combination with a wire-mesh sensor for measurement of conductivity.

In this publication we present a new method to characterize the residence time distribution from spatially resolved information. In a further step the technique could be applied to two-phase flow RTD measurements as the wire-mesh sensor is well established for measurements in water–air two-phase flow [4]. The experiments performed with a water flow demonstrate the potential of the measurement technique.

With the resulting data the length dependence of the axial dispersion coefficient D_L in structured packings is discussed.

2. Literature review

The following section attempts to review the most important literature on the length dependence of the dispersion coefficient and on flow in the entrance section of a pipe and of porous structures.

Danckwerts wrote in one of the first publications on the dispersion model: “D should, of course, be independent of L for a given u” [5]. In the same year Taylor showed that in laminar flow through a tube, the axial dispersion can be described by the dispersion model, provided the length of the pipe exceeds a certain minimum [6]. When this condition is not fulfilled the dispersion is non-Fickian and cannot be described by the dispersion model. When the dispersion model is applied to laminar pipe flow without fulfilling this condition, a dispersion coefficient varying with the length is obtained.

Levenspiel and Smith mentioned two conditions for the dispersion model to hold: uniform velocity profile or long enough measurement section [7]. Later studies [8,9] on the length effect identified the laminar boundary layer on the surface of a solid as an effect which cannot be described by the dispersion model.

Another important fact to be considered is that the mathematics behind the dispersion model with open–open boundary conditions only lead to a proper residence time distribution if $\frac{D_L}{u \cdot L} < 0.01$ ([10],

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p. 302). This is primarily because it is unclear what happens at the boundary. It is possible that some of the observed trends of length effects are resulting from a violation of this condition.

The continuing research on this topic shows that no closed theory was developed which allows to exclude all length effects. Delagado mentioned in his review article [11], in a much weaker form than Danckwerts [5], that the dispersion coefficient *should be* independent of the length, if an experimental method is valid. Nevertheless he then referred to Han et al. [12], who gave a criterion on the minimum length of a measurement section in order to reach a constant value of dispersion coefficient. A satisfactory explanation why the dispersion coefficient changes with the length is however not given. Han et al. only mentioned that this behavior is attributed to turbulence effects [12].

In summary it can be stated that the length effect is well known in literature but a conclusive theory is still missing. We will later propose that also the build-up of turbulence (or decay, depending on geometry) in the entrance section of a porous structure leads to a length dependent dispersion coefficient. For that we reflect on some results from literature on developing flow through pipes and porous structures.

In pipes, developing turbulent flow is a transition from a boundary layer type flow at the entrance to a fully developed flow downstream [13]. In turbulent flow the entrance section contains six flow regimes, for which theoretical models were developed [14]. Despite the identification of flow regimes many researchers investigated the length that it takes for the flow to be fully developed. This knowledge is of great importance for the design of rigs for internal flow investigations. Barbin and Jones observed that developed turbulent flow is not attained in a pipe length corresponding to 40 diameters [15]. Weir et al. states that for developed flow it takes longer than usually assumed, in their case more than 70 diameters [16]. Further they observed that the entrance length largely depends on inlet flow conditions. Recent studies present numerical treatments of the entrance flow. Kumara et al. found that for the laminar case after 31.78 diameters the centerline velocity reached 99% of its final value, what they defined as fully developed [13]. In the turbulent case they found that, after an overshoot of centerline velocity at about 30 diameters, developed turbulent flow is attained at about 65 diameters.

Similar studies in porous structures are rare. While the entrance flow in a pipe is expected to share some characteristics with entrance flow in a porous structure, there are also considerable differences, e.g. the periodic acceleration and deceleration of the flow in porous structures. Mokrani et al. investigated the flow over vortex generators mounted in a pipe [17]. Mean and statistic quantities of the flow were measured using laser Doppler velocimetry. They found that after four rows of vortex generators the flow reaches a steady periodic regime. A porous structure which is better comparable to our structure is investigated by Horneber et al. [18]. By numerical simulations the flow through a row of 8 units of a Kelvin cell is simulated. It is observed that fully developed flow, when judging from the development of mean velocity, is reached after very few cells. Butscher et al. [19] used particle image velocimetry with refractive index matching between the solid and the fluid to measure velocity fields inside a porous structure. In the resolved length scales the flow is developed after about two periodic units, what corresponds to 6 cells.

These studies imply that in porous structures the length until the flow is fully developed is much shorter (less than 10 cells) than in pipe flow (more than 50 diameters). The porous structure investigated in our study comprises 87 cells in a length of 0.2 m. We therefore conclude, based on comparison to the above mentioned literature, that the flow is fully developed after less than 0.2 m.

With this publication we contribute to the understanding of axial dispersion by stating that the build-up of turbulence in the entrance section of the porous structure leads to a length dependent dispersion coefficient. We further introduce the wire-mesh sensor as a measurement technique to assess both radial and axial dispersion in tubular reactors.

3. Theoretical background

3.1. Dispersion model

Idealized models are used to describe the residence time distribution of a real reactor. We use the dispersion model, which is a one dimensional model for a problem which has concentration and velocity fluctuations in three dimensions. The justification of this model has its origin in the work by Taylor [6]. He showed that even in laminar flow through a tube, where the velocity profile over the cross section is parabolic, the dispersion model is applicable provided that the time for convective transport is long compared to the time of decay during which radial variations of concentration are reduced to a fraction of their initial value through the action of molecular diffusion. Later the dispersion model was applied to turbulent flow [20]. By assuming the universal velocity profile over the cross section it was shown that a one-dimensional model can describe dispersion in turbulent flow. Today the dispersion model is, along with the tanks in series model, one of the two most important models of dispersion and can be used for turbulent flow in pipes and for flow in packed beds ([10], p. 293). The dispersion of mass in turbulent flow (stochastic movement of fluid) is similar to diffusion (stochastic movement of molecules). Therefore we use the dispersion model because it describes mass transfer with a coefficient (D_L) which is similar to the diffusion coefficient:

$$\frac{\partial c}{\partial t} = D_L \cdot \frac{\partial^2 c}{\partial x^2} - u \cdot \frac{\partial c}{\partial x} \tag{1}$$

The mass transfer is composed of advection with velocity u and dispersion with a coefficient D_L which contains diffusion and turbulent mixing.

Applying Eq. (1) to a tubular reactor of length L, there are different possible boundary conditions [10]. The open–open boundary condition takes an infinitely long section for calculation, motivated by the fact that there is also dispersion over the entry and the exit of the investigated section. The initial condition for Eq. (1) is to have a Dirac pulse at the inlet of a reactor at time zero. As known from literature [7], the solution to Eq. (1) for the outlet of the reactor ($x = L$) leads to the residence time distribution, given by:

$$E(t) = \frac{1}{\tau \cdot \sqrt{4 \cdot \pi \cdot (D_L/(u \cdot L)) \cdot t/\tau}} \cdot \exp\left(\frac{-(1 - t/\tau)^2}{4 \cdot (D_L/(u \cdot L)) \cdot t/\tau}\right) \tag{2}$$

The dispersion model describes the residence time distribution by only two parameters, i.e. the mean residence time τ and the dimensionless vessel dispersion number $\frac{D_L}{u \cdot L}$. A narrow residence time distribution is associated with a small value of $\frac{D_L}{u \cdot L}$.

3.2. Determination of RTD from pulse experiments

Injecting a Dirac pulse of tracer at the inlet of the reactor would allow the direct measurement of the residence time distribution at the outlet of the reactor. However, if the tracer pulse at the inlet is broad, the outlet signal is given as the convolution of the inlet signal and the residence time distribution:

$$c_{out}(t) = \int_0^t c_{in}(t^*) \cdot E(t - t^*) \cdot dt^* \tag{3}$$

When the concentration profiles are measured at the inlet and at the outlet, the residence time distribution can be determined by two methods: