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An experimental investigation on the onset from bubbling to turbulent fluidization regime in micro-structured fluidized beds

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ABSTRACT

Membrane-assisted fluidized beds have been investigated for a number of industrially important applications. Recently, micro-structured membrane-assisted fluidized beds have been proposed and particular for efficient hydrogen production in particular for maximising the volumetric production rate by maximising the installed specific membrane areas. Another important advantage of micro-structuring became evident from computational studies. CFD calculations have shown that micro-structured fluidized beds can be operated in the turbulent fluidization regime at much lower superficial gas velocities compared with larger fluidized bed [1]. This work focuses on an experimental validation of the transition velocity in micro-structured fluidized beds. The influence of the reactor scale and the particle diameter on the transition gas velocity from the bubbling to the turbulent fluidization regime in micro-structure fluidized beds was measured in pseudo 2D column using different experimental techniques: in particular, the pressure drop fluctuation measured by pressure sensors and the local solid holdup fluctuation measured by digital image analysis on images recorded by high speed camera. It was found that smaller reactors filled with particles of smaller diameter can be operated in the turbulent fluidization at a relatively lower gas velocity compared with larger reactors. Interestingly, the different measurement techniques result in somewhat different transition velocities, i.e. the pressure drop fluctuation method generated higher critical velocities than the local solids' holdup fluctuation method. This difference is discussed in this paper. The experimental results confirmed that micro-structured fluidized beds can indeed be operated in the turbulent fluidization regime with lower fluidization velocities thus anticipating increased energy efficiencies and enhanced mass transfer rates.

1. Introduction

With the increase in concerns about anthropogenic greenhouse gas emissions and their effects on climate change and the increasing costs of fossil fuels, more and more attention is being devoted towards the development of highly energy efficient reactor concepts. From this point of view, membrane reactors for hydrogen production have been proposed as potential candidates for both pre-combustion CO_2 capture and for small scale ultra-pure H_2 production units for fuel cell applications. On the one hands, more stable and highly permeable membranes are being developed that bring membrane reactors for pure H_2 production closer to industrial exploitation. On the other hand, more permeable and selective membranes drive the development towards more advanced reactor concepts such as membrane-assisted fluidized bed reactors, which may show advantages over conventional packed-bed membrane reactors, such as lower membrane area required and more uniform temperature profiles [2]. Usually, fluidized beds are classified based on the fluidization regime at which they are operated [3]. Membrane-assisted fluidized beds would be better operated in the turbulent fluidization to avoid the unnecessary bubble-to-emulsion phase mass transfer limitation that could decrease the efficiency of the membrane separation, thus requiring more expensive membrane area for a targeted H_2 recovery. Wang et al. [1] have shown by numerical simulations that a micro-structured fluidized bed membrane reactor has both the advantages that more membrane area is installed per unit of reactor volume, and also that the micro-structured reactor can be operated in the turbulent fluidization regime with much lower gas flow rates compared with bigger scale fluidized bed reactors. An experimental proof of these finding would drive the research towards the development of micro-structured fluidized bed membrane reactors because of the anticipated improved mass transfer characteristics. However, some controversies still exist in literature as whether the turbulent fluidization regime starts immediately after the bubbling regime or at a higher gas velocity. To distinguish the difference between the flow regimes, a common way is to define a so-called transition or critical velocity. Different interpretations of transition velocity as well as different techniques to identify it have been proposed in literature [1–6]. Several factors can affect the transition velocity e.g. the particle size, gas and solid properties, total static bed height, reactor size, experimental methods used, the way

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to interpret the results as well as the bed geometry [1,4,6,7]. There are several ways to explain the mechanism of the transition flow regime: (i) change in bubble behaviour where bigger bubbles are replaced by smaller bubbles resulting in a more uniform voidage and (ii) the change in solids' dynamics like the break-up of dispersed bubbles into a more continuous and less dense emulsion phase.

It was thus suggested that different fluidization regimes can exist at different positions inside the bed at the same time [4]. For example, the turbulent fluidization can be obtained at a higher superficial gas velocity at the bottom bed resulting in the formation of many small bubbles. In contrast to the bottom part of the bed, the turbulent fluidization regime can be obtained at a lower superficial velocity at the top part of the bed due to the presence of bigger coalesced bubbles that are broken into smaller voids resulting in more uniform fluidization.

This work will experimentally investigate for the first time the fluidization regime transition in smaller reactors to mimic the micro-structured fluidized beds previously studied by Wang et al. [1]. In particular, two different experimental techniques, viz. the pressure drop fluctuation method and the local solid holdup fluctuation method have been used and the results will be compared. The local solids' fraction inside the fluidized beds has been measured non-invasively via Digital Image Analysis on images recorded with a high speed camera. The solid holdup distribution in the whole flow field and the bubble size distribution have been investigated in great detail.

2. Experimental setup

Three pseudo-2D fluidized bed columns with 80, 40 and 20 mm width, 600 mm height and 10 mm depth were constructed and used in this experimental work (see Fig. 1). The front walls were made by transparent glass for optical access (required for image recording), while the black back walls were made of black anodized metal to obtain a good contrast between the particle and gas phases. The distributors were made of porous metal with averaged pore size of 10 μm and 3 mm thickness. Two types of glass beads with a particle size distribution within d_p = 400–600 μm and d_p = 100–200 μm, (ρ_p = 2525 kg·m^{-3}) were used (Geldart B classification). The minimum fluidization of 0.21 m·s^{-1} and 0.020 m·s^{-1} was measured by the pressure drop method for two different particle sizes respectively. The fluidization gas (air for all the experiments) flow rate was controlled digital mass flow controllers (MFC—Bronkhorst B.V.) range 10–200 l·min^{-1}, depending on the fluidized bed size and particle diameter used. To avoid the electrostatic attraction between the walls and particles, air was humidified up to 60–70% of the relative humidity by passing through a bubble column before entering into the fluidized beds.

The fluidized beds were illuminated by LED lamps placed at the proper positions to avoid reflections. The images were recorded by a CMOS high speed camera (Lavision 2016 × 2016 pixel resolution at f_{max} = 1600 Hz), equipped with a 50 mm and a 200 mm lens, depending on the particle size used, to achieve a good resolution required for digital image analysis (at least 2 pixels per particle diameter). A frequency of 20 Hz and an exposure time of 0.15 ms were set for the camera. The images were recorded and stored in the memory of the camera (4GB) and transferred to the memory of the PC. About 5000 images were used in each series of the experiment to achieve good time-averaged results.

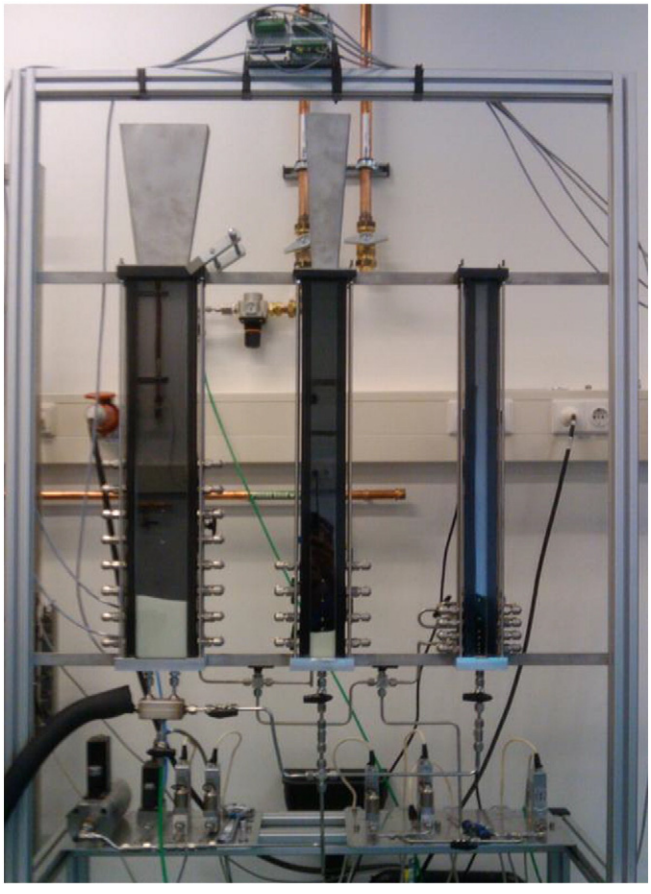

Fig. 1. Experimental setup of the three different fluidized beds: 80 mm, 40 mm and 20 mm bed widths from the left to right hand side.

3. Experimental techniques

3.1. Pressure drop fluctuation

The pressure fluctuation method is the most common technique that has been used to measure the transition velocity in gas–solid fluidized beds [1,3,4,8]. The pressure drop fluctuation can be measured either by the absolute pressure fluctuation or the differential pressure fluctuation. The absolute pressure fluctuation method refers to the measurement of the pressure fluctuations at a single location along the axial direction while the differential pressure method determines the transition velocity by measuring the fluctuations in the pressure difference at two different axial locations. In this study, several pressure sensors, purchased from Sensortechnics (±0.2 mbar), were positioned at different axial positions in the bed from the back wall of each column. The tips of the sensors were covered by filters to avoid particle blockage and connected to amplifiers for signal amplification. The outputs from the amplifiers were connected with a data acquisition box. Before using,

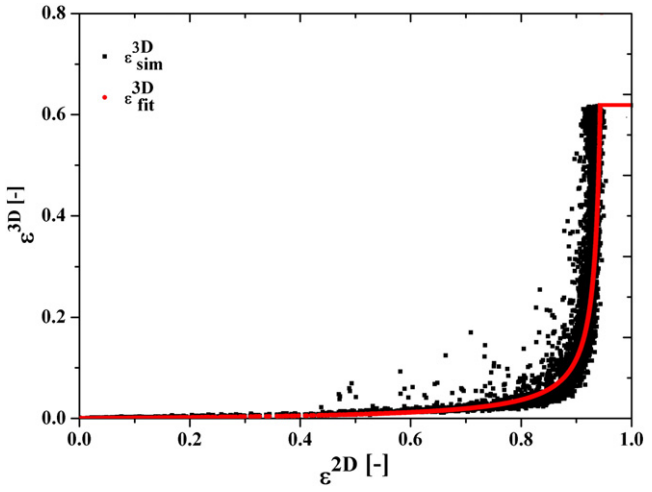

Fig. 2. The relationship between the 2D solids' fraction and the 3D solids' volume fraction (obtained from DEM simulation in the same conditions with the experiments).