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Study of reactor model of fluidized catalyst bed utilizing hydrogenation of carbon dioxide

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Abstract

Fluidized catalyst bed reactors are used in many chemical processes. However, despite the fact that there are many reactions in which fluidized beds are considered to be suitable as reactors, they are currently not fully utilized. One of the reasons is that scale up and design methods are not sufficiently established. In many cases, it is hard to explain the results of chemical reactions by reactor models. Namely, the models have not been effectively utilized [1, 2]. Levenspiel [3] pointed out that the uncertainty of the bubble diameter of the fluidized bed made the design and scale-up difficult. In addition, Tsutsui [4] has emphasized the role of the direct contacting particles that are the particles in contact with a gas concentration of bubbles. In this study, the hydrogenation of carbon dioxide was performed as a model reaction. Based on the experimental results, the axial distribution of the contact efficiency and the roles of contact efficiency of direct contacting particles are studied to examine the fluidized bed model that can accurately predict the reactor performance.

The reactor was made of a glass column with an inner diameter of 45 mm and height of 1.5 m. Because a transparent heater was coated on the outer surface of the column, it is possible to observe the inner surface of the column during the bed was fluidized by reactant gases and reactions were performed. The hydrogenation of carbon dioxide was performed over the catalyst of 20 wt% Ni-La_2O_3-Pt /Al_2O_3. The content of Pt was 0.05 wt%. The carrier was the porous alumina particles. The particle diameter and particle density were 55 μm and 880 kg/m^3, respectively. The reaction was performed by changing the gas velocity, the reaction temperature and the settled bed height.

The conversion was slightly higher for the lower gas velocity at the same temperature. At the same gas velocity, the conversion increased with settled bed height. This is due to the increase in the contact time between the reaction gas and the catalyst particles. The overall reaction rate constant K_{OR} was obtained by using the experimental results. The higher settled bed height was, the lower K_{OR} became. This is considered to be that the bubble diameter is small in the bottom of the bed and the mass transfer capacitance coefficient increased due to the large interface area between the bubble and the emulsion phase.

Figs. 1 and **2** show the values of $k_{ob}a_b$ and v that correspond to the value of K_{OR} obtained in the experiments for each condition. The values of $k_{ob}a_b$ increased with decreasing settled bed height. This was caused by the bubble growth for the bed with high settled bed height.

In addition, the fraction of the direct contacting particles increased with settled bed height. At the upper part of the bed, several small bubbles gather and rise. This situation can be seen as a large bubble containing much particles in it [9]. The above tendency can be explained by considering these particles as the direct contacting particles.

Both the parameters, $k_{ob}a_b$ and v increased with gas velocity as shown in these figures. When the gas velocity increased, the number of bubbles increased, and the interface area also increased. Therefore, these parameters increased with gas velocity. In the case of this study, when the fraction of the direct contacting particles was 2–15%, the experimental results could be explained. Therefore, the role of the direct contacting particles cannot be ignored.

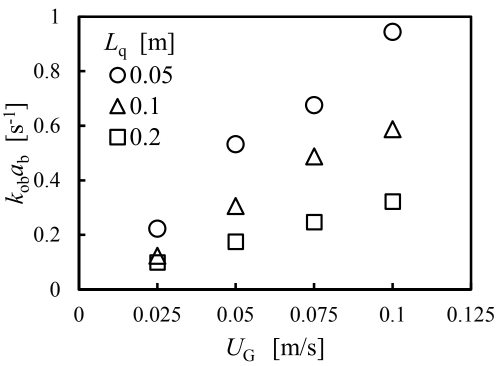

Fig. 1 Effect of gas velocity on the mass transfer capacitance coefficient

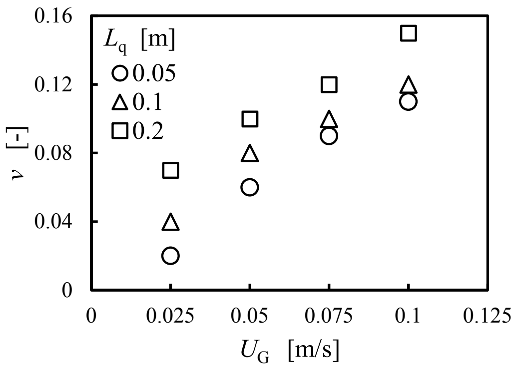

Fig. 2 Effect of gas velocity on the fraction of the direct contacting particles

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