Chapter V

DISCUSSIONS

5.1 Considerations on the Design of Experimental Apparatus

The main interest in studying gas absorption in fluidized-bed column is the determination of mass transfer coefficient. In order to improve the efficiency of mass transfer coefficient, essential parts of the experimental apparatus were carefully designed and chosen according to their functions as follows.

Fluidized-bed column has a relatively large distance between the grids. Under these conditions, the spheres were in turbulent motion and will not migrate as a bed to the top grid. Thus, an essentially self-relieving packing is provided.

Non uniformity in the movement or fluidization of the spheres was the experimental difficulty. The spheres tended to move entirely in up flow on one side and entirely in packed down flow on the other. This behavior was less likely if the gasdistributor as shown in Figure 3.4 were used. This type of distributor also counteracted the normal tendency of higher gas flow in the center of the bed and to increase the flow of gas in the outer portion of the bed.

The water was sprayed from a special type water distributor as shown in Figure 3.5. At low water flow rate, this distributor prevented water to combine together and fall down the center of the column, so water will distribute uniformly over the bed.

The air velocities were controlled by a butterfly valve. Butterfly valve occuplied less space in the line than any other type of valve so it will not cause to much pressure drop in the air line, and provides enough gas velocities in the column.

5.2 Hydrodynamic Properties of the Fluidized-bed Column

5.2.1 Effect of Gas Velocity on Hydraulic Resistance of Bed (ΔP_b)

The effect of gas velocity on hydraulic resistance of bed were illustrated in Figure 4.1 and 4.2. Visual observation and curve representing variation of the bed resistance with gas velocity at difference liquid rate showed that there were two hydrodynamic state. In the first state (stationary packing), there was a sharp increase in ΔP_b. At surperficial gas velocity higher than 110 cm/sec, there were slightly increase of pressure drop with gas velocities. The minimum fluidization were difficult to determine from ΔP_b directly because the spheres were rather large and light weight than those normally found in conventional fluidized-bed and those studied by Balabekov (11) and hence no smooth fluidization could be expected.

5.2.2 Effect of Static Packing Height and Superficial Liquid Velocity on Hydraulic Resistance of Bed (ΔP_b)

It can be observed from Figure 4.4 that ΔP_b increased linearly with static bed height. Visual observation indicated that the uniform and intensive agitation of the packing without sharp fluctuation of the dynamic bed height were obtained in columns with low static packing height. As the static packing height increased the dynamic bed height increased more fluctuation and when the static packing is higher than diameter of the column, the spheres tended to move as a floating bed type and there was a remarkable decrease in gas flow at sharply increase pressure drop,but this limited bed height could be increased as increasing liquid velocity. ΔP_b also increased linearly with increasing liquid velocities as shown in Figure 4.3. Liquid velocities had little effect on ΔP_b that may caused by the fact that increased liquid velocities would slightly increased the amount of liquid retained by the packing. Increasing liquid velocities also increased more uniform and intensive agitation of the fluidized-bed

5.2.3 Minimum Fluidization Velocity (G_{mf})

Because it was difficult to determine the minimum fluidization velocity from the pressure drop of the bed, Chen and Douglas (13) had defined G_{mf} determined from the bed height.

The results of G_{mf} from this study and calculated were shown in Table C-6. In Chen and Douglas 's equation only the effect of packing diameter and liquid velocity had been studied. In this study, the effects of bed height on G_{mf} were also studied. The results are shown in Figure 4.9 and can be concluded that G_{mf} are not varied with static packing height and decrease with increasing liquid velocities. When the results were compared with Chen and Douglas's correlation (2.26) the result agree well with their correlation as shown in Figure 5.1.

5.2.4 Effect of Gas Velocities, Liquid Velocities and Bed Heights on Gas Hold-up (ϵ_G)

The effects of gas velocities, liquid velocities bed heights on gas hold-up were shown in Table C-8. It was observed that ϵ_G was nearly independent of liquid velocities and bed heights but increased with increasing gas velocities. Comparison between ϵ_G observed and ϵ_G calculated from Kito, et al.'s empirical equation (2.28) were shown in Figure 5.2. The experimental result correlated well with empirical equation.

5.2.5 Effect of Gas Velocities, Liquid Velocities and Bed Heights on Liquid Hold-up (ϵ_L)

The amount of liquid retained per unit cross-sectional area of the bed is shown in Table C-7. It is evident that in the range of studying, heights of clear liquid were not affected by gas velocities but were affected by liquid velocities and bed heights.

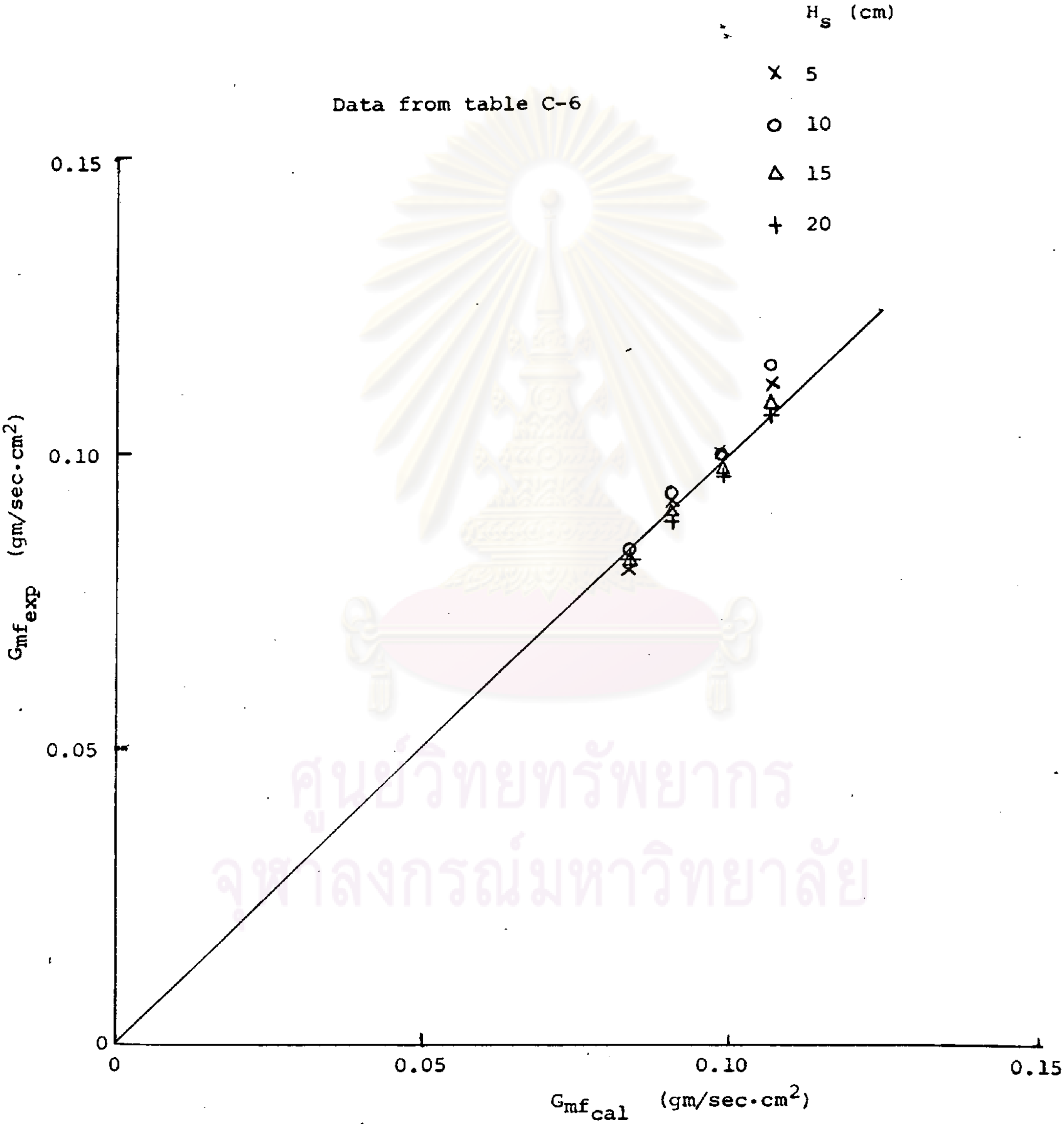

Fig.5.1 Comparison between $G_{mf_{exp}}$ and $G_{mf_{cal}}$

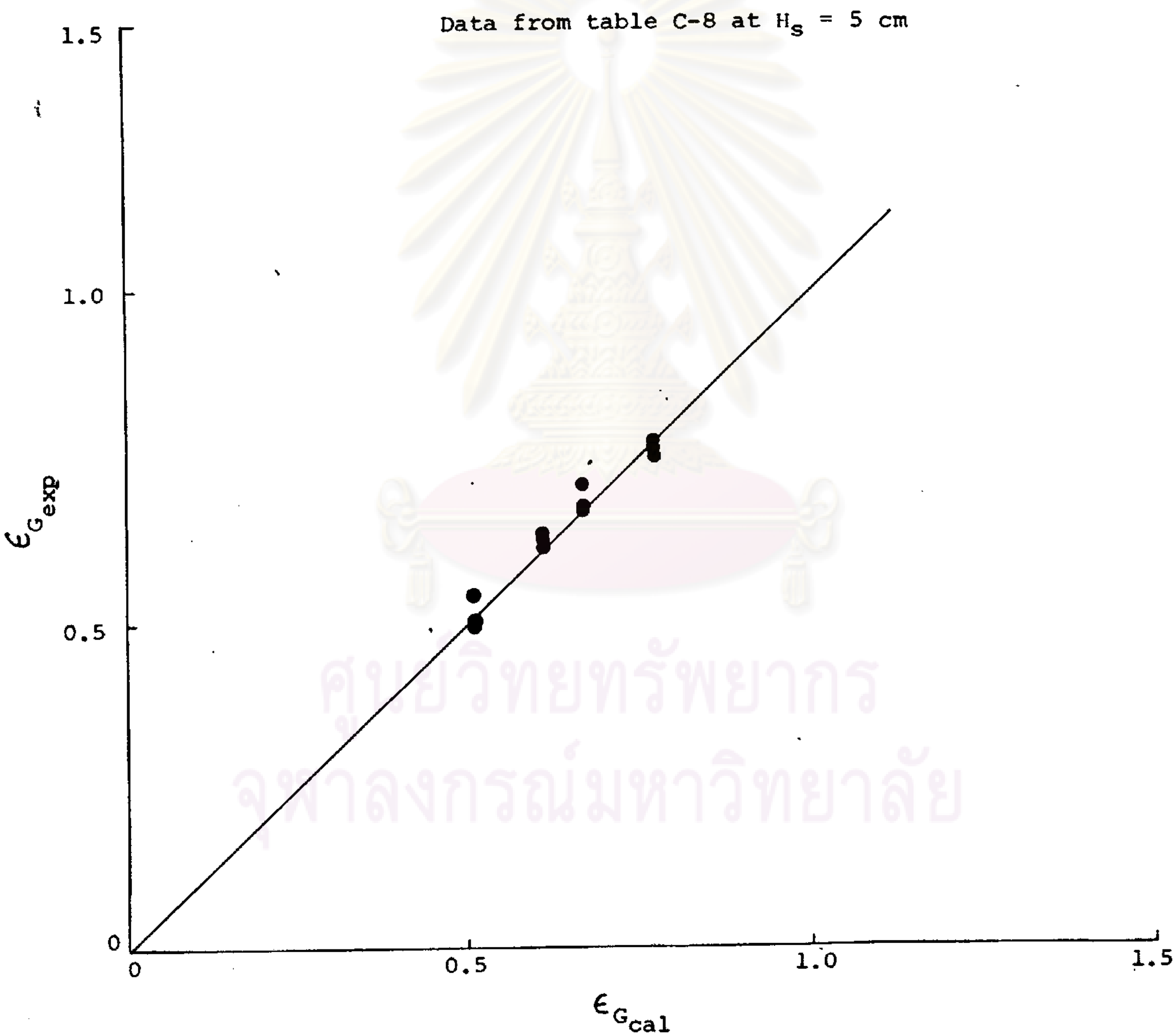

Fig.5.2 Comparison between $\epsilon_{G_{exp}}$ and $\epsilon_{G_{cal}}$

Visual observation of the present operation indicated the absence of flooding for any set of experimental conditions used. There fore, the gas velocities approximately 0.4573 gm/sec cm^2 (390 cm/sec) in this study was much below the loading velocity of this fluidized bed column. As H_L did not depended on gas velocities, the values of liquid hold-up, ϵ_{SL}, that defined as the ratio of H_L/H_S were also independent of the gas velocity. The resulls of ϵ_{SL} are shown in Table C-10. ϵ_{SL} calculated from Kito, et al.'s correlation (2.31) were also compared with experimental values. In Figure 5.3 all the experimental data of ϵ_L were plotted in accordance with Kito, et al.'s correlation. So, correlation for ϵ_L also applied well in the region that were studied.

5.3 Gas Absorption in a Fluidized-bed Column

At the first part, the hydrodynamic properties of fluidized-bed column were discussed and it is evidence that the system could be run at high liquid and gas velocity with low pressure drop. In this part, dependence of mass transfer coefficient of ammonia on liquid velocity, gas velocity, bed height and mole fraction of ammonia were studied. Although initial studied of ammonia-air-water system were pridicted on the hypothesis that it was gas phase controlling, recent studies had indicated that the liquid may provide 40 percent of the resistance at 25°C, thus the high value of mass transfer coefficient were expected for high liquid and gas velocity

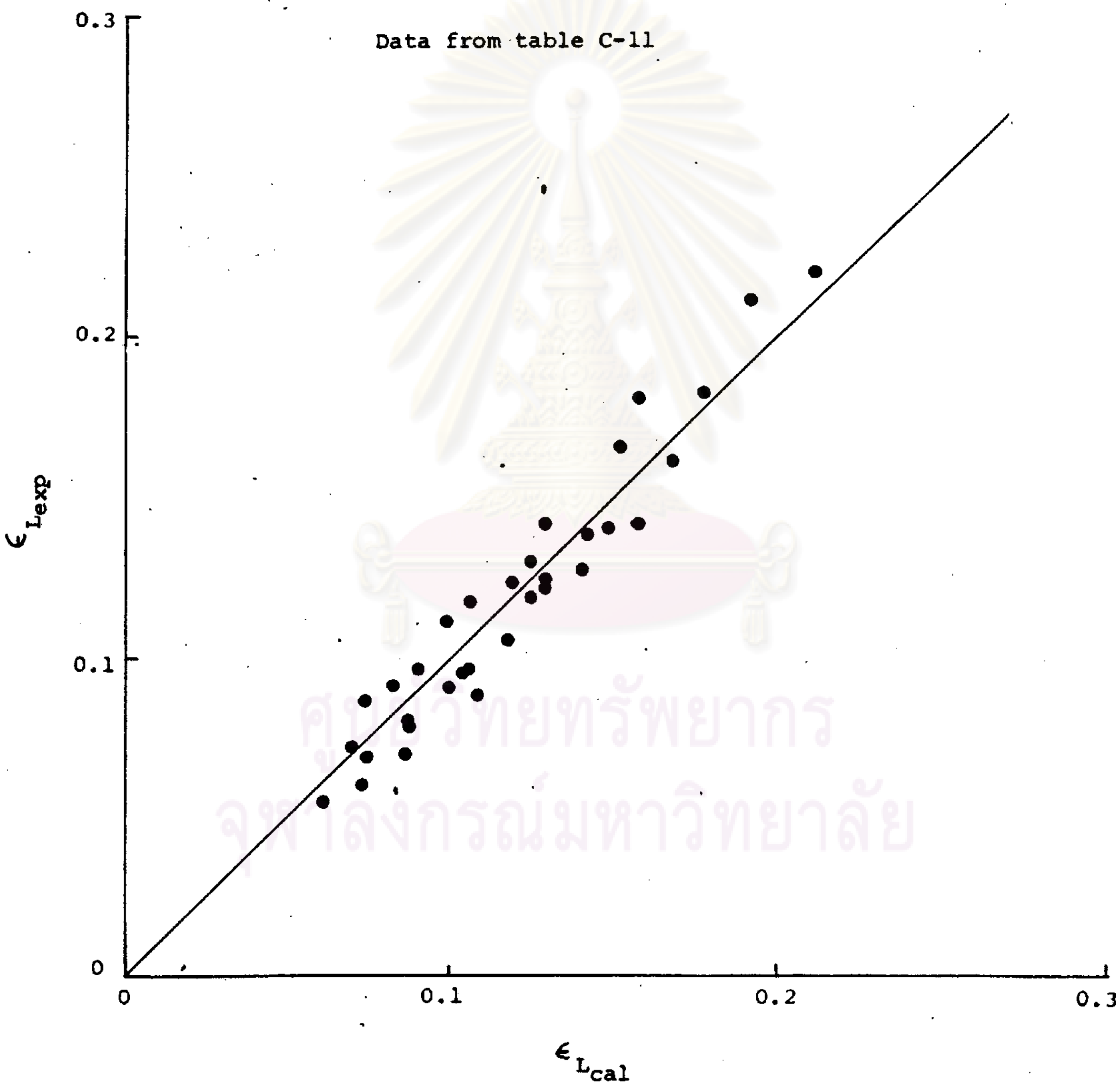

Fig.5.3 Comparison between $\epsilon_{L_{exp}}$ and $\epsilon_{L_{cal}}$

of turbulent bed system. The results were also compared with fixed-bed absorbers.

5.3.1 Effect of Gas Velocities on Mass Transfer Coefficient (K_Ga)

The effect of gas velocities on K_Ga were illustrated in Figure 4.10-4.17 Gas velocities were expressed in Froude number. It was observed that K_Ga increased as increasing Froude number. This could be explained by the mechanism of mass transfer. As increasing gas velocity the spheres were in more turbulent motion that increased it surface area and bring more fresh liquid from the interior to the surface. The average value of the slope was 0.495. So mass transfer coefficient expressed in terms of Sherwood Number, could be related to Froude number as below

$$Sh = e_1 \; Fr^{0.495} \qquad \text{----- (5.1)}$$

5.3.2 Effect of Liquid Velocities on Mass Transfer Coefficient (K_Ga)

The effect of liquid velocities on K_Ga were illustrated in Figure 4.18-4.21. Overall mass transfer coefficient also increased with increasing Reynold number with an average slope of 0.442. Increasing liquid velocities, the spheres were in more uniform and turbulent motion that increased mass transfer coefficient. The relation between Sherwood number and Reynold number could be expressed as

$$Sh = e_2 Re_L^{0.442} \quad \text{--------- (5.2)}$$

5.3.3 Effect of Bed Heights on Mass Transfer Coefficient ($K_G a$)

A rather interesting result of this study was the decreasing of mass transfer coefficient as increasing bed height. The results were illustrated in Figure 4.22-4.32. It had been explained in the first part that is increasing bed heights the fluidized bed increased more fluctuation that resulting in decreasing $K_G a$. The relation between Sherwood number and bed height $\frac{D_c}{H_s}$ were

$$Sh = e_3 \left(\frac{D_c}{H_s}\right)^{0.621} \quad \text{--------- (5.3)}$$

5.3.4 Effect of Mole Fraction of Ammonia on Mass Transfer Coefficient ($K_G a$)

The effect of mole fraction of ammonia between 6.41×10^{-3} to 10.86×10^{-3} were insignificant. Althrough some of the values of $K_G a$ tended to increase as increasing mole fraction of ammonia but most of the results rather fluctuated as increasing mole fraction of ammonia. So, it can be concluded that mole fraction of ammonia had not pay the big role on mass transfer coefficient in the range that were studied.

5.3.5 The Relation between Mass Transfer Coefficient and Experimental Dimensional Variables

From the results obtained, Mass Transfer Coefficient were affected by gas velocity, liquid velocity and bed height. The values of mass transfer coefficient expressed in term of Sherwood number were proprotional to $F_r^{0.495}$, $Re_L^{0.442}$ and $\frac{D_c}{H_s}^{0.621}$. Thus, the Sherwood number can be represented as follows.

$$Sh = e_4 \, F_r^{0.495} \, . \, Re_L^{0.442} \, . \, \frac{D_c}{H_s}^{0.621} \quad \text{-- (5.4)}$$

The value of e_4 determined from experimental results was 7.17×10^9. The temperature of the system, size of particle, diameter of column and liquid properties were kept constant and included in term of e_4. The correlation together with experimental results are shown in Figure 5.4.

5.3.6 Comparison of Height of Transfer Unit (H_{og}) of Fluidized-bed Absorber with Fixed-bed Absorbers

The results of this experiment from Table C-11 expressed in term of H_{og} were used to compare with that for fixed bed absorbers. The most extensive data are that of Fellinger (26) which is readily available in Perry (24). As shown in Figure 5.5, Fellenger's data show a similar steady decreased in height of transfer unit with increasing liquid rate and increased in height of transfer unit with increasing gas rate until the loading point was reached after which there was a sharp drop. In this fluidized-bed experiment, there were no conditions which analogous to loading

Data obtain from table C12-C17

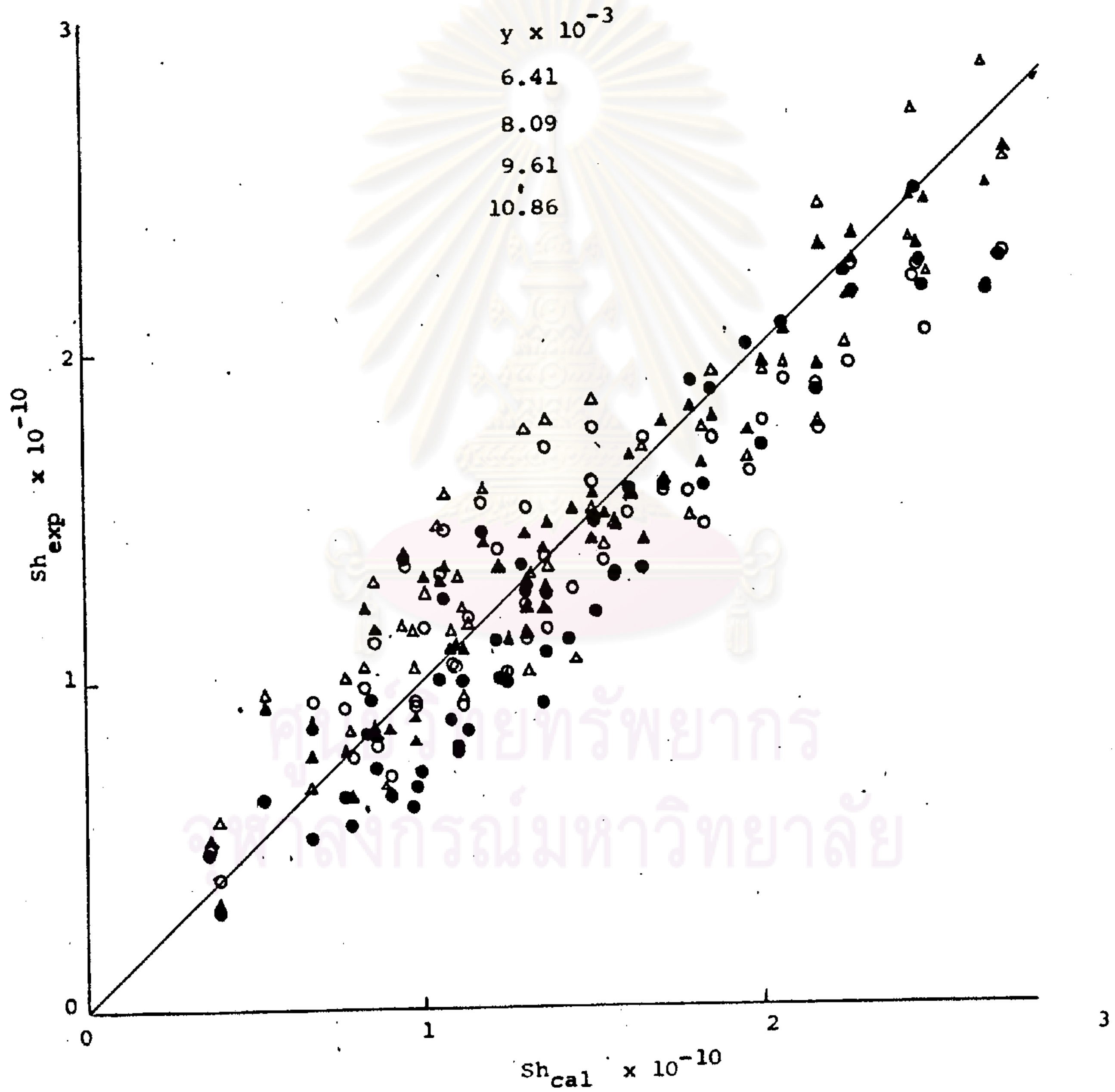

Fig.5.4 Comparison between Sh_{exp} and Sh_{cal}

occured. The range of gas and liquid mass velocities used by Fellinger and most other investigators of fixed-bed absorbers did not, however, overlap the range of these variables used in the present study. The gas mass velocity in most of these studies was limited to values below 0.1356 $gm/cm^2 sec$ because of loading and flooding of the columns. For fixed bed absorbers, one exception to the normal low capacity limitation was the study of Williams Akell and Talbot (27), in which they used a special, vertically stacked Fiberglas packing in a 6 in diameter column to get liquid mass velocities up to 4.068 $gm/cm^2 sec$ at gass mass velocities up to .202 $gm/cm^2 sec$. Lines representing their data had been added to Figure 5.5 and 5.6 for each of comparison. It is seen that the fluidized-bed absorber is considerably more efficient than this packed bed, as indicated by the considerably higher values of height of transfer unit obtained by Williams, et al.

The experimental results also compared with that obtained by Douglas (7). As shown in Figure 5.5 and 5.6, Dauglas's data show higher value of height of transfer unit than the experimental results. The experimental unit used by Douglas was a square tower instead of a cylindical tower that result in nonuniformity in the movement of the spheres which lowering the efficiency of the system.

5.3.7 Industrial Application

The very high gas and liquid rate are possible for this

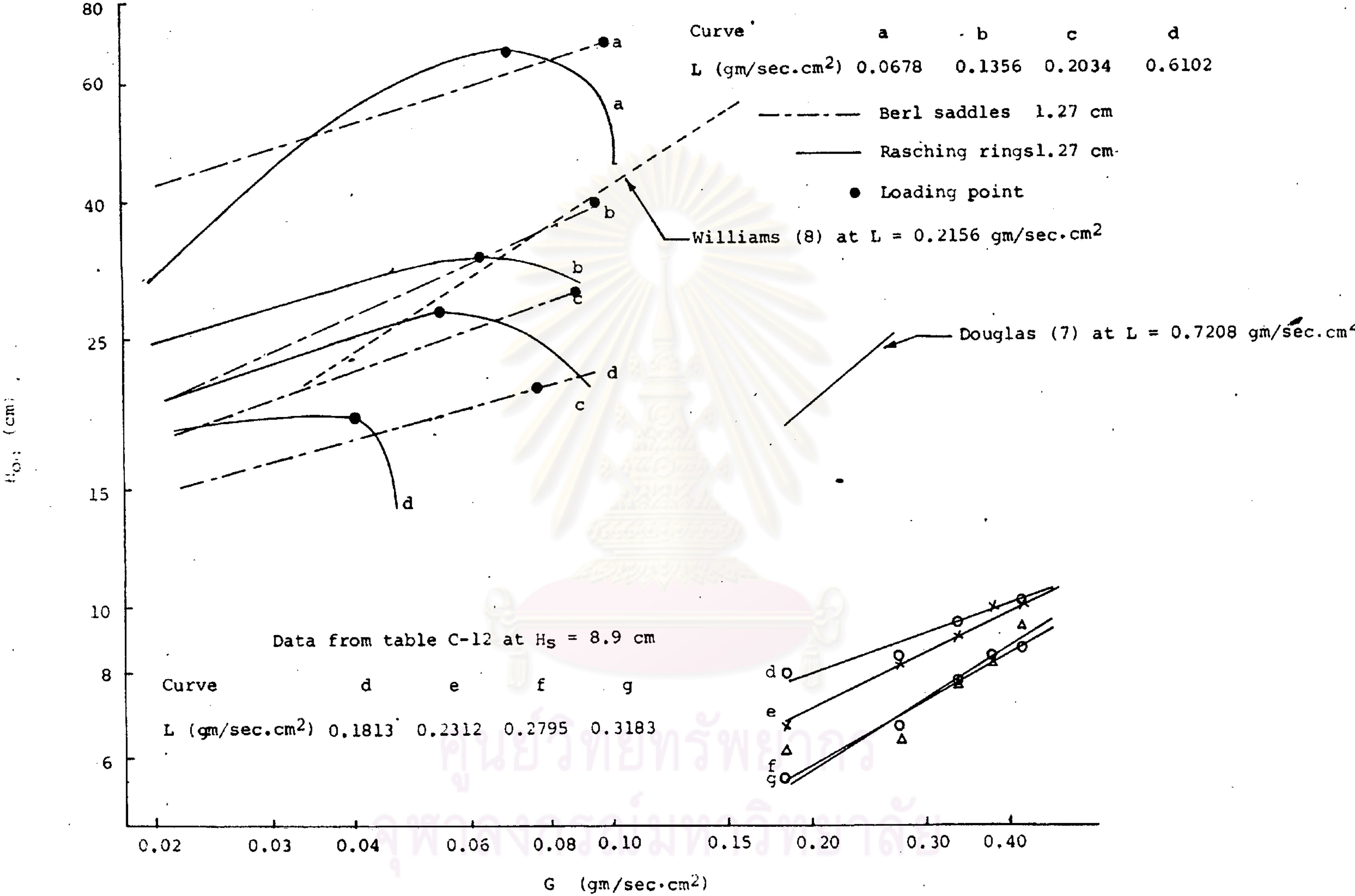

Fig.5.5 Comparison of H_{og} between fixed bed and fluidized-bed at various G

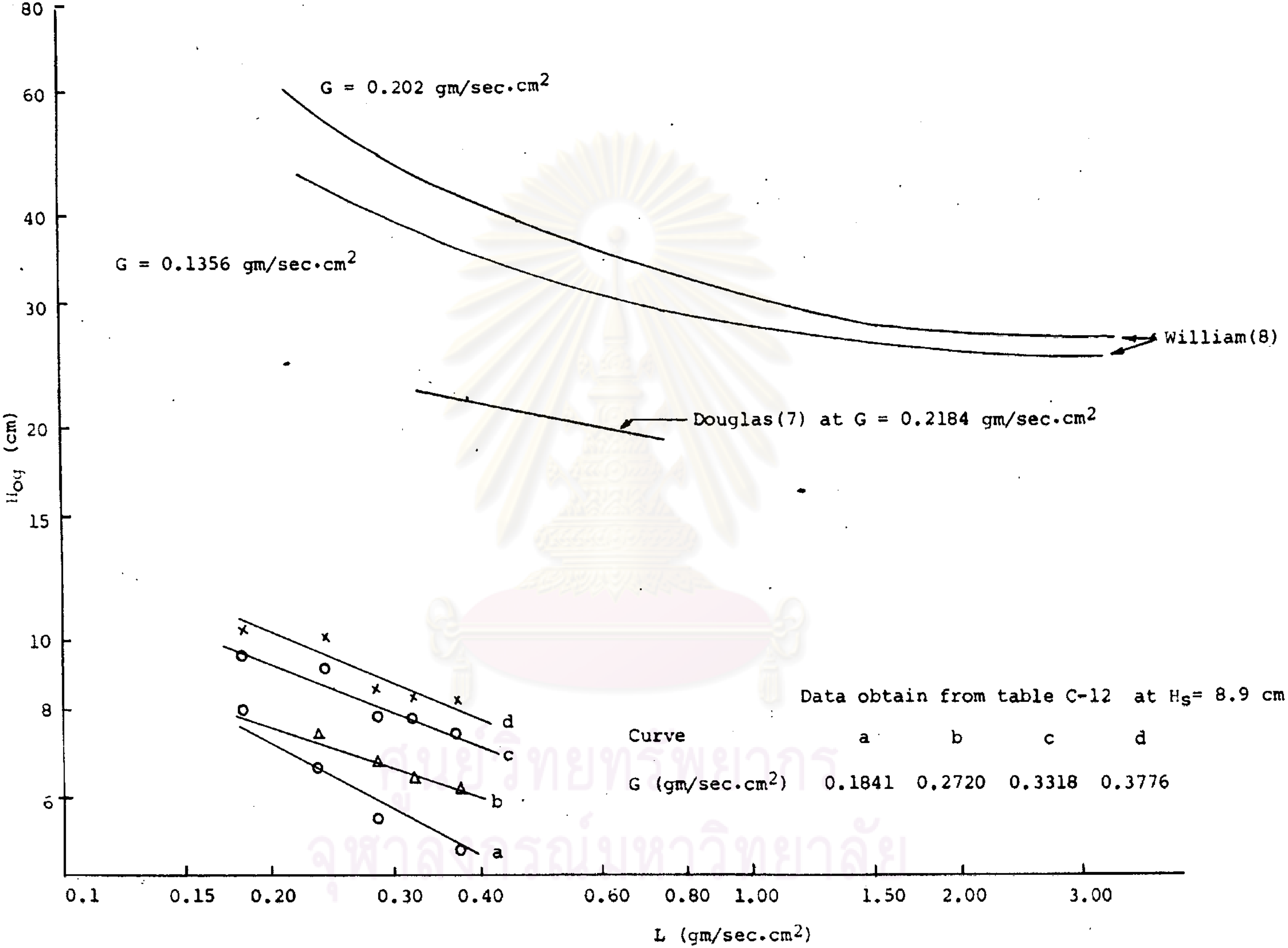

Fig.5.6 Comparison of H_{og} between fixed bed and fluidized-bed at various L

fluidized-bed absorber. The high gas velocities contribute to an extremely high turbulent. The spheres are rotated violently over the bed in which liquid in the absorption zone are very active in all directions. Absorption takes place not only on the wetted surface of the spheres but also throughout the whole active zone. The high rate of intimate mixing tends to minimize any effect of the relatively slow diffusion rate normally encountered in packed towered, thus maintaining large driving force with resultant high absorption rates. The high gas and liquid rates at low pressure drop across the bed result in high capacity and efficiency for a given tower volume when compare with packed-bed absorber. This system is also simplified to control and operate. It is very interesting in applied fluidized-bed process to air pollution control such as gas scrubbers in sulfuric acid plant for scrubbing sulfur dioxide from the waste gas. Since the motion of the packing prevents the plugging problem which often occur in conventional fixed-bed absorber, Fluidized-bed process also very useful in absorption of gas that a solid phase is presented or is formed by reaction of the contacting fluids, for example, absorption of sulfur dioxide in sodium hydroxide solution to produce sodium bisulfite liquor and dust collectors in wood working processes.