Chapter II

LITERATURE REVIEWS AND THEORECTICAL CONSIDERATION

2.1 Gas Absorption Equipment

Gas absorption operations are commonly conducted in towers which are designed to provide intimate contact of the two phases. Welty(1) classified this equipment into four general types according to the method used to produce the interphase contact as shown in Figure (2.1)

1. Spray towers consisted of large open chambers through which the gas phase flowed and into which the liquid was introduced by spray nozzles or other atomizing devices. Figure (2.1a) illustrates the direction of phase flow in a spray tower. The spray nozzle was designed to subdivide the liquid into a large number of small drops; for a given liquid flow rate smaller drop provide a greater interphase contact area across which mass was transferred. But if the droplets so fine they become entrained in the exiting gas stream, that made it could not use high gas velocity. Spray towers are used for mass transfer of highly soluble gases where the gas-phase resistance normally controls the rate of mass transfer.

2. Bubble tower exactly opposite in principle to the spray tower, the gas was dispersed into the liquid phase in the
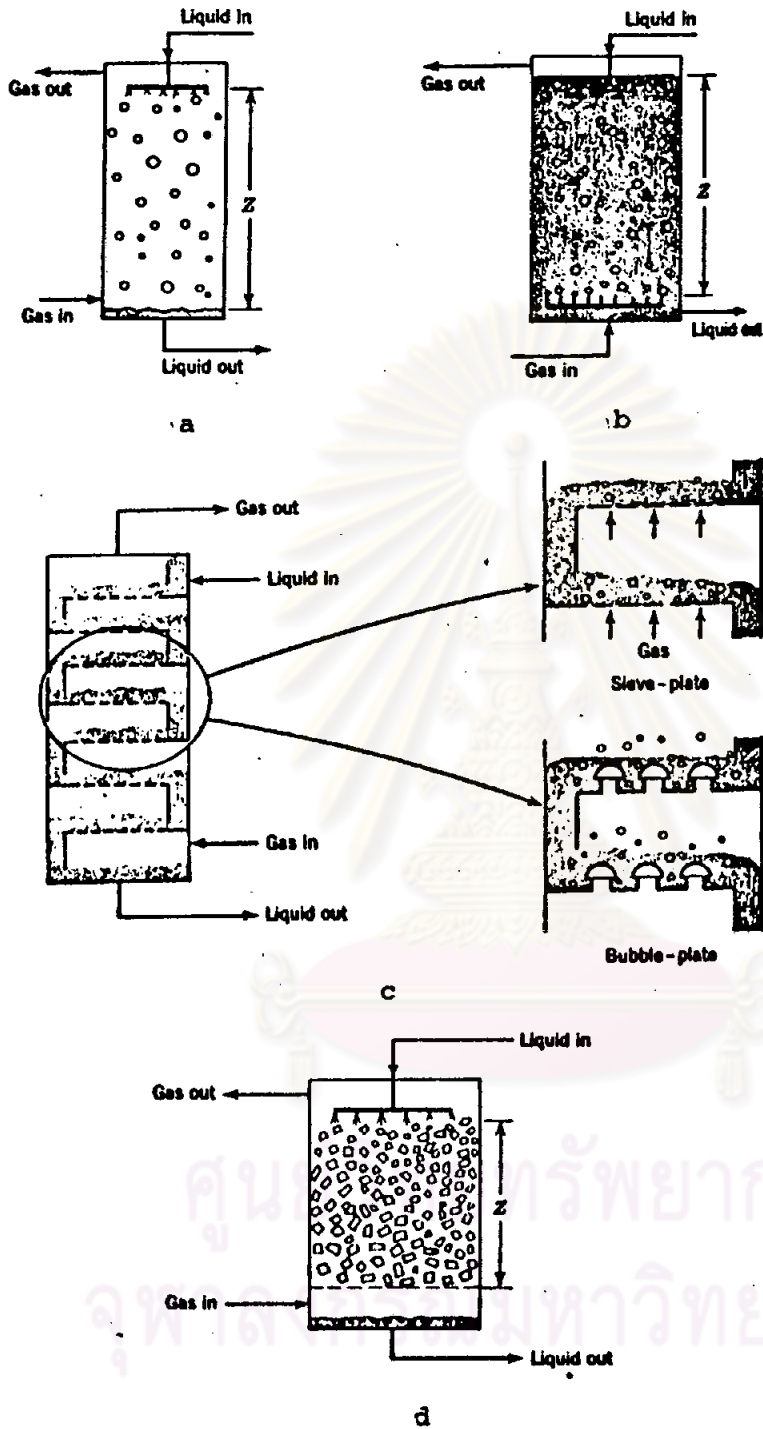
Fig.2.1 Gas absorption tower. (a) Spray tower.
(b) Bubble tower.(c) Plate tower.(d) Packed tower.

form of fine bubbles. The small gas bubbles provide the desired large interphase contact area. Mass transfer takes place both during the bubble formation and as the bubbles rise up through the liquid. Bubble towers are used with systems in which the liquid phase controls the rate of mass transfer; i.e., the absorption of relatively insoluble gases.

3. Bubble-plate and sieve-plate towers were commonly used in industry. They represent the combined transfer mechanisms observed in the spray and the bubble towers. At each plate, bubble of gas were formed at the bottom of a liquid pool by forcing the gas through small holes drilled in the plate or under slotted caps immersed in the liquid. Interface mass transfer occurs during the bubble formation, and as the bubbles rise up through the agitated liquid pool.

4. Packed towers used for continuous countercurrent contact of two immiscible phases, were the fourth general type of mass-transfer equipment. These towers were vertical column which had been filled with a packing materials. The liquid flowed down the packing surface as thin film or subdiveded streams. The gas generally flowed upwards. Both phases were well agitated. Thus this type of equipment may be used for gas-liquid systems in which either of the phase resistance controls or in which both resistances were important.

2.2 Fluidized-bed Column

Packed and plate tower are the most important equipment

in absorption processes. Plate tower may be preferred when there are deposite of solid material that must be removed periodically because it can be fitted with man holes. Also total weight is ' less than packed tower for the same duty. But construction of packed towers is usually simpler and cheaper when compare with plate tower. Other things being equal, economic consideration usually show that packed tower are favored over plate tower for absorption when the column diameter is less than about 60 cm. Because of the plugging problem in packed tower and large size in plate tower, a new type of absorber for the efficient removal of particulate matter from gas was developed by A. Kielback of the Aluminium Co. of Canada and had been described by Aerotic Industries, Inc. (2). This unit, shown in Figure (2.2), used low density spheres for packing which were retained between two grids with relatively small free board. These spheres formed a moving or floating bed when adequate gas and liquid flows were used and this virtually eliminates plugging or channeling. This process was currently used by the Aluminum Co. of Canada for absorption of hydrogen fluoride from hot process gases. Emanating from Soderberg pots used for electrolytic production of aluminum from alumina, these gas streams also contained a considerable amount of tar derived from electrode binding materials, as well as fluoride particulates. In all Aluminum Co. of Canada has installed six fluidized bed unit, each 47.2 m^3/sec feed capacity. Absorption efficiencies could range upwards of 90% at a pressure drop of 10 cm

Clean gas

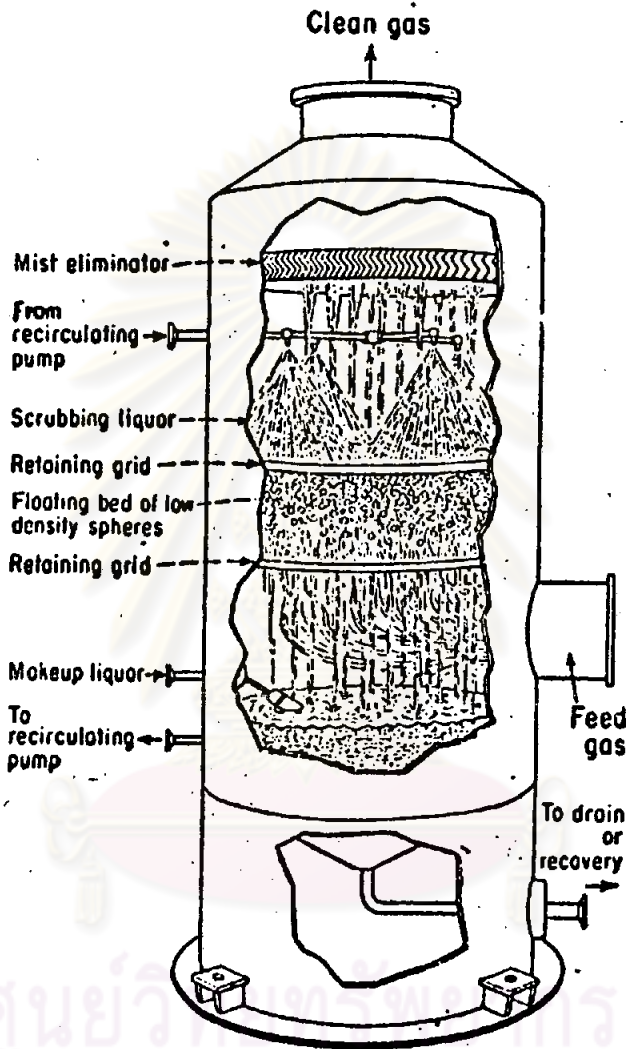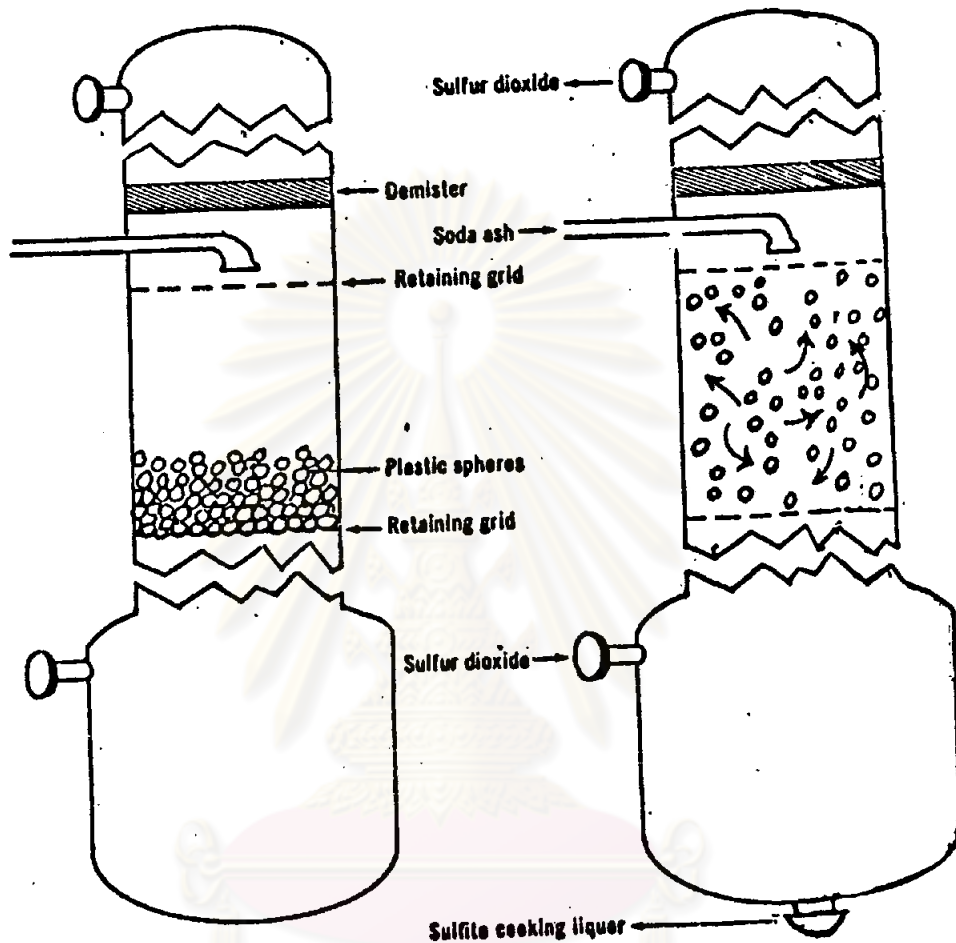Mist eliminator

From
recirculating
pump

Scrubbing liquor

Retaining grid

Floating bed of low
density spheres

Retaining grid

Makeup liquor

To
recirculating
pump

Feed
gas

To drain
or
recovery

Fig.2.2   Floating-bed scrubber

H_2O according to this firm solids-collection efficiency varies between 95% and 100%. Out standing was that the processes did not become plugged, even when handling sticky agglomerates. So gas absorption in fluidized-bed column process had its greatest initial use as the scrubber. Liquor makeup rates varied to some extent with the kinds of gases being absorbed, or with the incoming solids concentration. In the Canadian applications, about 946 cm^3 liquor would treat 28.3 m^3 incoming gas. Aerotec offered their floating-bed scrubbers with body and gridwork fabricated of virtually any construction material-from wood and plastic to lead-lined steel. Standard capacities available range from 0.71 m^3/sec (60 cm dia. by 300 cm height) to 18.90 m^3/sec (300 x 975 cm).

A related technique, had higher gas velocity and higher free space than that described by aerotec, developed by Dominion Tar and Chemical, Montreal, Canada 2.3. As shown in Figure 2.3, this technique used absorber columns one tenth the size of conventional packed columns of the same capacity. Domtar's columns were designed to operate without flooding at several times the gas and liquid rate practicable in packed absorbers.

Each stage of the column, which was sometime called turbulent contact absorber, contains shallow layers of lightweight polyethylene or polypropylene spheres on widely spaced retaining grids. During operation, the spheres were in free, turbulent

Sulfur dioxide

Demister

Soda ash

Retaining grid

Plastic spheres

Retaining grid

Sulfur dioxide

Sulfite cooking liquor

STATIC CONDITION
About one third of the
volumn between grids
may be filled with
polyethylene
or polypropylene
spheres.

OPERATING CONDITION
Sphere act much like
particulate agitators.

Fig. 2.3 Turbulent contact absorber makes sulfite cooking liquor
for papermaking at Dominion Tar.

motion in the space between the grids. The resulting agitation brought gas and liquid into close contact. According to ·Domtar, the tendency of static packing to cause liquid hold-up and flooding was avoided. The threat of flooding limits flow rates in packed columns, holdling designed flow rates down to about 60%. The turbulent contact design was said to permit much higher rates-superficial gas velocities up to about 609 cm/sec, compared to a typical packed-column rate of about 101 cm/sec.

When feed rates fell below a critical rate the sphere settle on each retaining grid and act much like conventional packing, Domtar's says. The allowable range of feed rates, though, was a broad one. Liquid rates down to 0.0679 cm^3/sec cm^2 at a gas velocity of 254 cm/sec had been used successfully, according to the firm. The estimate top rate for economical operation was about 3.4 cm^3/sec cm^2 with a gas velocity of 762 cm/ sec. The absorber was developed at Domtar's central research laboratories, Cornwall, Ont., for use in the company's papermaking operations. A unit now was used at a Domtar pulping mill in Trenton, Ont., absorbs sulfur dioxide in a solution of soda ash to form sulfite cooking liquor. It handled as much as 1.70 cm^3/ sec cm^2 liquid at superficial gas velocities ranging from 254 to 508 cm/sec. Pressure drop through the column is said to be about 25.4 cm of water. This three-stage column was 35 cm in diameter and 480 cm tall. By comparison a column packed with Raschig rings

specified for this service would have to be 122 cm in diameter
and 915 cm tall. The moving spheres were not easely clogged
when the feed streams contained solids or when precipitated
form in the column it self. This was important when making
sulfite cooking liquor, since sulfur could precipitate in the
absorber under some operating conditions halting the operation.
This process also applicable to other chemical operations.
A rubber lined turbulent contact column was used as a scrubber
to remove hydrogen fluoride and other corrosive gases evolved
when sulfuric acid contacts crushed phosphate-bearing rock.

Pilot studied with this type of operation, first carried
out by Douglas (4). The experimental unit was constructed of
929 cm^2 by 305 cm long stainless steel. The packing used for
these studies was 3.8 cm diameter hollow-polyethylene spheres.
It was found that the superficial gas velocity could exceeded
914 cm/sec., at the pressure drop in column not more than 25.4 cm.
The rather startling magnitude of counter current gas and liquid
rates was in sharp contrast to those possible in conventional
packed towars which have limited gas and liquid rate due to the
flooding characteristics of static packing as shown in Figure 2.4.
These curve defined the maximum countercurrent gas and liquid
rate for a given packing; normal operation was below these rates.
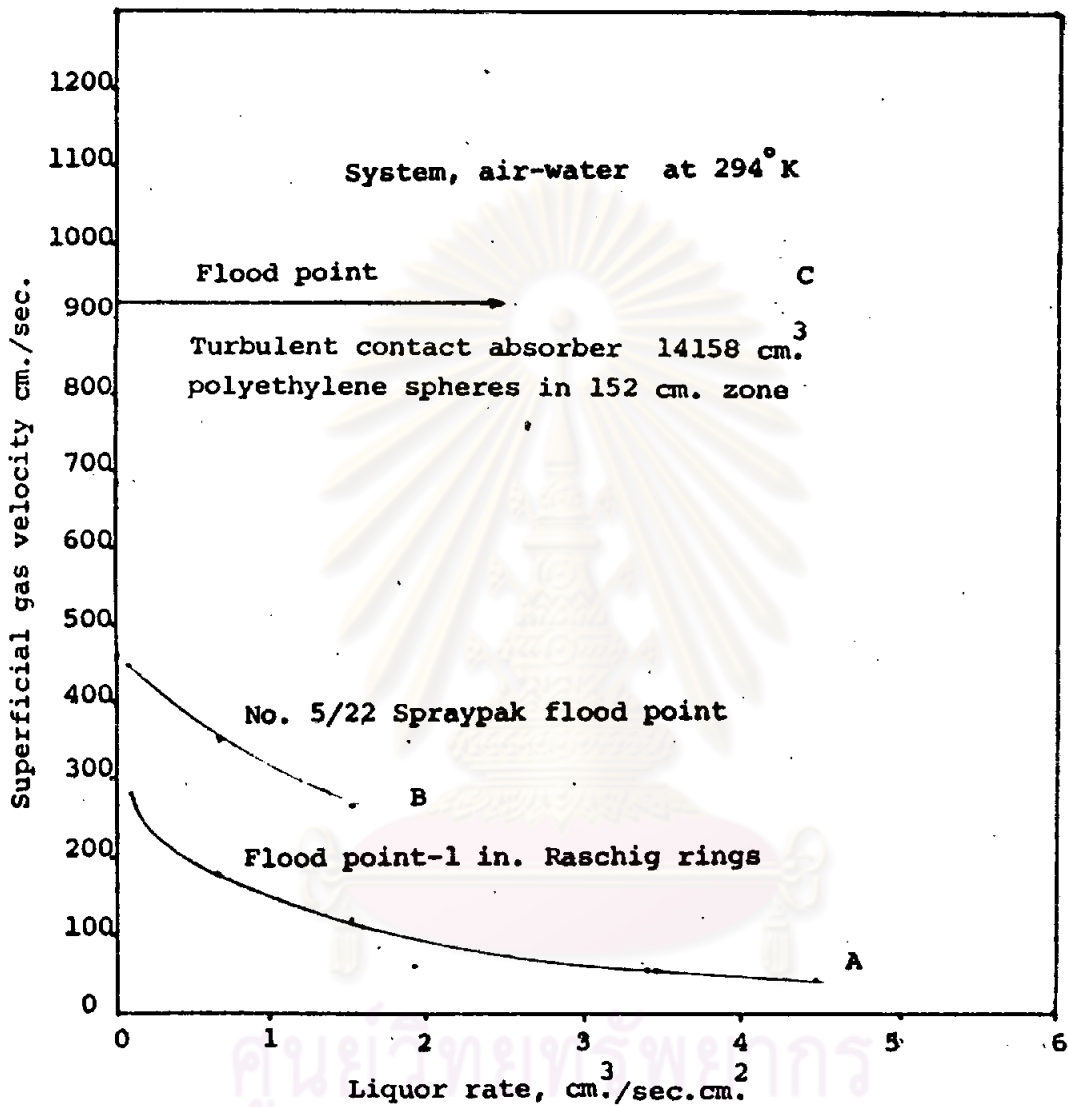Literature values for 2.54 cm × 2.54 cm. Rasching ring (5) were

Fig.2.4 Comparison of flood point data for various absorber
packings. System: air water at 294° K.

shown in curve A as typical of conventional packing. Curve B illustrated the increased fluid rates possible with Spraypak, a new type of tower packing. Curve C illustrate the fluid rate possible in the turbulent contact absorber. They also study the absorption of CO_2 in alkaline process liquor in 929 cm^2 fluidized-bed tower with 0.0736 m^3 of packing and compare the result with carbonation in the process production of Vanillin in Cornwall Ontario, mill of Callery Chemical Co., that carbonation was carried out with flue gas using two coke-packed towers (122 cm diameter and have 13.19 cm^3 of packing). These towers were always subject to plugging problem. The absorption coefficients, both in high and low pH liquor, of pilot plant were over 70 times that obtained in the mill tower.

Dauglas (7) had studied heat and mass transfer in a turbulent bed contactor 929 cm^2 cross section, packed with hollow polyethylene spheres, the fluids being air and water. For absorption of Ammonia it was observed that the height of transfer unit decreased with increased liquid flow rate and increased with increasing gas flow rate. It was also observed that the height of a transfer unit decreased with decreasing static bed height. The gas mass velocity in most fixed bed absorber was limited to values below .1356 gm/sec cm^2 because of loading and flooding of the column. One exception to the normal low capacity limitation was the study of Williams, et al.

(8), in which they used a special, vertically stacked Fiberglas

packing to get liquid mass velocities of up to 4.068 gm/sec cm^2

and gas mass velocity up to 0.202 gm/sec cm^2.

Groeneveld (9) had measured the specific gas-liquid

interfacial area in a bed of ping-pong balls. The specific in-

terfacial area was proportional to liquid flow rate, and increased

with increasing gas flow rate, a slow increase being observed

below the flooding point and a rapid increase at the flooding

point, where the interfacial area reached a value of 200 m^2/m^3.

Wozniak & Ostergoard (10) had studied the system of tur-

bulent bed of hollow spheres contained in a column of circular

cross-section of 0.1 m.dia. mass transfer was studied by using

the reaction between carbon dioxide and sodium hydroxide. They

studied the gas-liquid interfacial area and the gas-side resis-

tance to mass transfer and concluded that turbulent bed contactors

made it possible to obtain high value of the effective interfacial

area, strongly increasing for an increase of the liquid superficial

velocity. Such contactor are not suitable, however, for gas film-

resistance controlled absorption processes, as indicated by the low

value of K_G.

Balabekov, et al (11) had study the operating condition of

columns with wetted moving spherical packing. They confirmed that

columns with wetted fluidized packing could be operated at high gas

velocities (up to 4-7 m./sec) and liquid rates (up to 125 m^3/m^3.hr.),

which were several times the rate used in ordinary packed and plate

towers; this ensured high efficiency per unit volume. They
also described the existance of four hydrodynamic states of
column operation in the range of gas velocities studied, as
shown in Figure 2.5 . In the first state (stationary packing)
the spheres were in close mutual contact and the bed volume
remains unchanged with some variations of the velocity; the
liquid flows down the center of the packing in film for. while
the gas passes predominantly along the column walls. This
state existed in a narrow range of flow velocities, the upper:
limit being the start of fluidization. The second state (ini-
tial fluidization) began immediately beyond the stability
limit of the compact bed and was characterized by fluidization
of a certain proportion of the packing with out intense motion
of the spheres. The range of existence of this state depends
on the liquid rate and on the characteristics of the spheres
and supporting grids. The constancy or slow increase of the
bed resistance with increase of the gas velocity in this regime
was due to a proportional increase of the number of moving
spheres in the bed, which leads to increase of the cross section
open to the passage of gas. Here only a part of the liquid was
present in the fluidized part of the bed; most of it flows down
the stationary packing at the walls without coming into contact
with the gas. The third state (developed fluidization) mass
characterized by intense motion of the phases and instability of
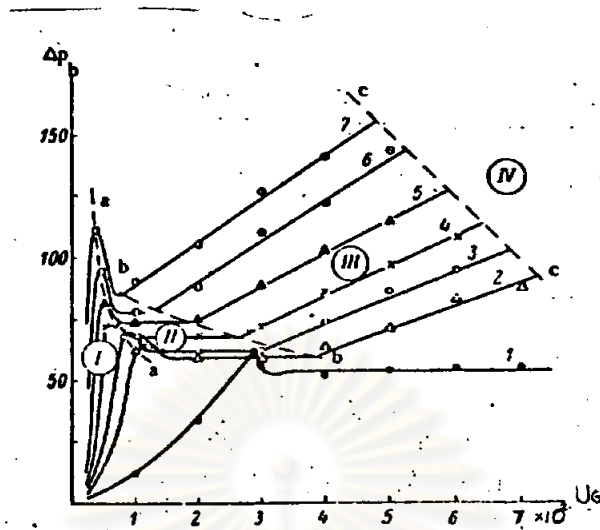the interface, with vortices of one phase penetrating rapidly

Fig. 2.5 Dependence of hydraulic resistance ΔP_b (mm water column) of the packing on the gas velocity (cm./sec.). $H_s = 110$ mm, $D_p = 16$ mm $\rho_s = .85$ gm./cm.^{3} ; slotted grid, open cross section = 40%, slot width = 3 mm. 1) Dry packing; liquid rate (m^3/m^2·hr.); 2) 15; 3) 30; 4) 50; 5) 75; 6) 100; 7) 125. State: 1) stationary packing; 11) initial fluidization; 111) developed fluidization; 1V) flooding. aa) Start of fluidization; bb) start of developed fluidization cc) start of flooding.
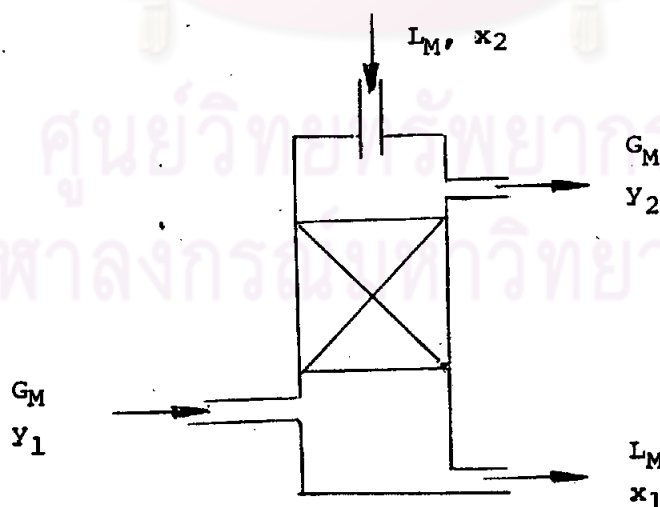

Fig. 2.6 Flow diagram for counter-current absorption in a packed tower.

into vortices of the other. This was the most important state in absorption because of high interfacial area and high phase contact. The fourth state (flooding of the column) was characterized by sharp increase of liquid entrainment by the gas; the packing was pressed against the upper retaining grid, and the hydraulic resistance of the packing therefore increased sharply.

Balabekov, et al., (12) also studied the hydrodynamic characteristic of column and reported the results of studies of the dependence of the main hydraulic characteristic of column operation on the characteristic of the packing and grid at various liquid rate. They also state that three zones could be observed in a column packed with spheres of density 356 kg/m^3.

1. The zone over the grid, with a large amount of turbulent liquid penetrated by bubble or streams of gas and a low concentration of spheres. In this zone the movement of the phases was the most intensive, owing to the high kinetic energy of the gas flowing through the section of the grid not occupied by descending liquid.

2. The middle zone, consisting of fluidized packing together with liquid in the form of film and streams, intimately singled with bubbles and streams of gas. The concentration of the spheres was higher in this zone than in the others.

3. The upper zone, consisting of liquid spray and individual spheres.

These three zones were always present in the fluidization

range of wetting packing. Only their relative heights vary.
A very law-gas velocities the heights of the zone over the grid
and the upper zone were negligible; conversely, at very high
gas velocities they were very large while the middle zone was
small, The heights of the zone over the grid and of the upper
zone decreased with increasing density of the spheres. They
also conclude that, in a column packed with spheres having the
same density as the wetting liquid, the bed consisted mainly of
suspended liquid (foam) and spheres distributed uniformly over
the entire bed height. In a column packed with spheres of
density 2600 kg/m^3 the space over the grid could also be sub-
divided into three zones,corresponding to three different gas
liquid-solid systems; 1) the zone of fluidized packing and
bubbling liquid, located directly above the grid; nearly all
the packing was present in the zone; 2) the zone of stable foam;
3) the liquid spray zone. The packing consisting of spheres of
much higher density than the wetting liquid, operates like the
grid in a foam column.

Chen and Douglas (13) have defined the minimum fluidi-
zation velocity in a turbulent bed contactor as the maximum gas
velocity at which the packed bed maintains its static height.
They have determined the minumum fluidization velocity of poly-
styrene foam spheres in a bed of 30.48 cm diameter and 30.48 cm.
static height, the fluids being air and tap water. Special
packing of three size,1.27 cm, 2.54 cm and 3.81 cm were used.

The results demonstrate that the minimum fluidization velocity decreased with increasing liquid velocity and increased with increasing particle diameter.

Levsh et al (14) have presented a theorectical analysis of pressure drop in a turbulent bed contactor. They regarded the operation as a combination of of fluidization of the packing by the gas stream and bubbling of the gas through the layer of liquid hold up by the supporting screen. They proposed equation for the determination of the pressure drop of screens without packing as a function of the gas and liquid velocities. The theorectical pressure drop values deviced from these equations compare well with experimental determination.

Chen and D. uglas (13) have suggested that the pressure drop across the packing may be determined by use of a correlation for irrigated fixed packings.

Ticky, et al., (15) had measured the vertical static pressure profile in a mobile bed contactor for spherical packing of two sizes, 1.27, 1.90 cm, and three fixed-bed heights. Foam polystyrene spheres with a density of 155 kg/m^3 were used as packings. They concluded that for a fully-mobile bed the pressure drop is independent of the diameter of the packings used. The vertical static pressure profile within the bed was found to be linear. The dependence of the "drag parameter" on the liquid flow rate and the static bed height was given by simple linear relationships.

Chen and Dauglas (13) have determined liquid hold-up in three phase fluidized bed system by transient tracer method. The experiments were carried out below the flooding point and it was observed that liquid hold-up was independent of gas flow; this was also the case for fixed beds below the flooding point, as observed by several investigators. They also pointed out the important difference between irrigated fixed and turbulent bed contactors was that liquid hold-up in a fixed bed consists of the operational hold-up and the static hold-up, and the latter contributes only to a limited extent to interfacial transport processes; but in a turbulent bed contactor the static hold-up is practically absent due to the motion of the packing, and the total liquid hold-up thus contributes to interfacial transport. This may account partly for the higher rates of heat and mass transfer in turbulent bed contactors as compared with fixed bed absorbers under similar operating conditions. It was found that liquid hold-up increased with increasing liquid flow rate and with decreasing particle diameter.

Groeneveld (9) measured liquid hold-up in a bed of ping-pong balls. A slow increase of hold-up with increasing gas velocity was observed below the flooding point. At the flooding point (at a superficial gas velocity of 6 m/sec) a rapid increase of liquid hold-up was observed. Liquid hold-up was found to be nearly proportional to the liquid flow rate.

Barile and Mayer (16) presented empirical correlations

for the liquid hold-up but it was applied only at the minimum fluidization velocity.

Kito, et al., (17) stated that, the amount of liquid retained per unit cross sectional area of the bed, H_L, in the range from minimum fluidization velocity to flooding velocity was not affected by gas velocity but affected by liquid velocity and physical properties of liquid. It increased with increasing liquid velocities, liquid viscosity and surface tehsion of liquid. Correlations were also presented for liquid holdups in terms of independent process variables.

The gas hold-up was studied by Balabekov, et al. (11), Kito, et al. (18) and Krainev, et al. (19). An emperical correlation for the gas hold-up was presented by Kito, et al.(17) which was obtained under the liquid stagnent flow system. The correlation, therefore, applied only to the systems investigated, and may not be extended for the design of apparatus operating under different process conditions. Kito, et al. (18), studied the effects of the packing density, physical properties of liquid and the static bed heights on the gas hold-up and concluded that the value of gas hold-up was nearly independent of packing density, liquid viscosity, static bed height, liquid velocity, packing diameter and free opening area of the supporting grid.

2.2.1 Design Consideration

Aerotic Industries Inc. (2) state that gas absorption in

fluidized-bed column must use low density spheres for packing
in order to allow the sphere in free, turbulent motion in the
space between the grids. Balabekov (12) added that if the
bed consist of spheres of much higher density than that the
wetting liquid. The bed acted like the grid in a foam column
and would cause increasing in hydraulic resistance. Also the
spheres must be produced from material that are not brittle
such as polystylene, foam, hollow polyethylene.

Douglas (7) pointed out that the movement or fluidiza-
tion of the spheres was the principle experimental difficulty
experience. In the deeper beds, the spheres tended to move
entirely in up flow on one side and entirely in packed down
flow on the other. This problem disappeared for small bed
depths and this behaviour was less in a cylindrical column.
It had subsequently been found possible to reduce this tendency
by using an upper grills of specially curve form so that the
spheres were in up flow in the center of the unit where the
hightest gas velocity exists, and were in downflow around the
edges.

Dauglas (4) stated that in three phase fluidized-bed
absorber it must be a relatively large distance between the
grids, say 150 cm. for 30 to 60 cm. of static depth of packing.
Under there conditions the sphere were in turbulent motion and
would not migrate as a bed to the top grid. Thus, an essen-
tially self-relieving packing is provided.

Because this process needed high gas velocity in order
to get turbulent and random motion of the sphere, so a high
volume blower must be used.

2.2.2 Theorectical Consideration

The transfer of material from a gas to a liquid takes
places in three stages, viz. diffusion from the gas to the
liquid surface, solution in the liquid and diffusion from the
surface to the body of the liquid. It was suggested by .
whitman (20) that the rate of absorption was controlled by the
rates of diffusion on each side of the interface, and the
equilibrium was reached at the surface almost instantaneously.

It is usually imposible to measure the partial pres-
sure and concentration at the interface, and it is therefore
convenient to employ overall mass transfer coefficients base
on the overall driving force between p_A and c_A. The overall
coefficients may be defined on the basis of the gas phase K_G,
or the liquid phase, K_L by the equation.

$$K_G = \frac{N_A}{p_A - p^*} \quad\quad\quad --------(2.1)$$

$$K_L = \frac{N_A}{c^* - c_A} \quad\quad\quad --------(2.2)$$

The design calculation for absorption tower are rendered
complicated by the fact that there are several different systems
of units in current use. Experimental data on the performance

of absorption towers are usually reported as gas and liquid film mass transfer coefficient or as height of transfer units. The concentration of the solute in the gas phase is expressed as its partial pressure or mole fraction, and in the liquid phase the concentration may be expressed as a mole fraction.

A considerable proportion of the industrial absorption processes are concerned with the scrubbing of a gas to remove a relatively small proportion of a soluble constituent. In this case there is little change in the total flow rates of the gas and liquid entering and leaving the tower and as a first approximation these flows can be assumed constant at all stages of the absorption. It can also be assumed that the transfer coefficients are constant at all stages.

The flow diagram for counter-current absorption in a packed tower is represented by Figure 2.6. The quantity absorbed may be calculated from the change in composition of either the gas or the liquid as it passes through the tower, since the materials balance equation for the process is.

$$G_M (y_1 - y_2) = L_M (x_1 - x_2) \quad -------(2.3)$$

Where G_M and L_M are the molar flow rates of gas and liquid, y and x are the mole fractions of the soluble component and the subscripts 1 and 2 denote the bottom and the top of the tower respectively. In the partial pressure and concentration units the materials balance equation is

$$G_M \frac{(p_1 - p_2)}{P} = \frac{L_M}{\rho_M}(C_1 - C_2) \qquad \text{------- (2.4)}$$

The rate of absorption on a unit area of the packing in any part of the tower is equal to the product of the over-all gas phase transfer coefficient K_G and the difference between the partial pressure of the solute gas at that point, p and the partial pressure in equilibrium with the liquid on the packing, p^* from equation (2.1)

$$N_A = K_G(p - p^*)$$
$$= K_G P(y - y^*) \qquad \text{-------- (2.5)}$$

The area of packing in a small element of a tower of unit cross-section and height dH is a.dH and the rate of absorption in this element is given by the equations

$$N_A \, a \, dH = K_G \, a \, P(y - y^*) \, dH$$
$$= G_M \, dy$$
$$= L_M \, dx \qquad \text{------- (2.6)}$$

$$\text{or} \quad N_A \, a \, dH = K_G \, a \, (p - p^*) \, dH$$
$$= G_M \frac{dp}{P}$$
$$= \frac{L_M}{\rho_M} \, d \, C \qquad \text{------- (2.7)}$$

In order to integrate these equations over the height of the tower it is necessary to determine the relation between the driving force $(y - y^*)$ or $(p - p^*)$ and the gas or liquid composition. In many cases involving dilute solutions this

problem may be simplified by assuming that the solubility obeys

Henry's law and is represented by the equations.

$$y^* \quad = \quad mx \quad\quad \text{------- (2.8)}$$

$$p^* \quad =. \quad \mathcal{H}\, C \quad\quad \text{------- (2.9)}$$

With this simplifying assumption it can be shown that

the mean driving force for the whole of the tower is equal to

the logarithmic mean of the concentration or partial pressure

differences at the top and bottom which may be calculated from

the terminal conditions, the material balance equation and the

solubility. To prove this theorem it is first necessary to

derive relations between the liquid and gas composition and

the driving force at every point in the tower. The relation

between the two composition x and y at any level is obtained

by applying the material balance equation to the section of the

tower below that level; from equation (2.3)

$$y_1 - y \quad = \quad \frac{L_M}{G_M}\,(x_1 - x) \quad\quad \text{------- (2.10)}$$

From equation (2.8)

$$y_1^* - y^* \quad = \quad m\,(x_1 - x) \quad\quad \text{------- (2.11)}$$

and the driving froce is given by

$$\triangle y \quad = \quad y - y^*$$

$$= \quad y_1 - y_1^* + \left(m - \frac{L_M}{G_M} \right)(x_1 - x) \quad (2.12)$$

By differentiation of these equations

$$d\,\triangle y \quad = \quad \left(\frac{L_M}{G_M} - m \right) (dx) \quad\quad \text{------- (2.13)}$$

$$dy = \frac{L_M}{G_M} dx$$

$$= \frac{L_M/G_M}{\overline{(L_M/G_M)} - m} \, d\Delta y \qquad \text{------- (2.14)}$$

Also from equation (2.3) and (2.8)

$$\frac{L_M/G_M}{\overline{(L_M/G_M)} - m} = \frac{(y_1 - y_2)/(x_1 - x_2)}{(y_1 - y_2)/(x_1 - x_2) - (y_1^* - y_2^*)/(x_1 - x_2)}$$

$$= \frac{y_1 - y_2}{\Delta y_1 - \Delta y_2} \qquad \text{------- (2.15)}$$

Where

$$\Delta y_1 = y_1 - y_1^*$$

$$\Delta y_2 = y_2 - y_2^*$$

Hence

$$dy = \frac{y_1 - y_2}{\Delta y_1 - \Delta y_2} \, d\Delta y \qquad \text{------- (2.16)}$$

From equation (2.6)

$$\frac{K_G \, a \, P \, dH}{G_M} = \frac{dy}{\Delta y}$$

$$= \left(\frac{y_1 - y_2}{\Delta y_1 - \Delta y_2}\right)\left(\frac{d\Delta y}{\Delta y}\right) \qquad \text{----- (2.17)}$$

Equation (2.17) may be integrated between the terminal compositions:

$$\frac{K_G \, a \, P \, H}{G_M} = \frac{y_1 - y_2}{\Delta y_1 - \Delta y_2} \int_{y_2}^{y_1} \frac{d\Delta y}{\Delta y}$$

$$= \frac{y_1 - y_2}{\Delta y_1 - \Delta y_2} \ln \frac{\Delta y_1}{\Delta y_2}$$

$$= \frac{y_1 - y_2}{\Delta y_{LM}} \qquad \text{-------- (2.18)}$$

Where Δy_{LM} is the logarithmic mean of Δy_1 and Δy_2, given by

$$\Delta y_{LM} = \frac{\Delta y_1 - \Delta y_2}{\ln(\Delta y_1 / \Delta y_2)}$$

From equation (2.18)

$$K_G a = \frac{G_M(y_1 - y_2)}{H P \Delta y_{LM}} \qquad \text{---------(2.19)}$$

The method given above is limited in application to cases where only a small proportion of the total gas flow is absorbed. If the solute gas constitutes more than 5 to 10% of the gas entering the tower it is necessary to take account of the reduction in quantity of the gas as absorption proceeds which affects both the materials balance and the mass transfer coefficient. It is also necessary to take account of the heat liberated by the solution of the gas since this raises the liquid temperature and reduces the solubility of the gas.

2.2.3 <u>Mass Transfer Driving Force in Turbulent-Bed Column</u>

In the performance calculations, the important factor is the mass transfer driving force. The mass transfer differential equation must be integrated over the entire bed in order that the terminal concentrations may be used. The over all gas

phase driving force which enters the differential equation is the difference between the solute gas partial pressure in the bulk gas and the pressure which would be in equilibrium with the bulk liquid composition at the same point. Since this driving force varies in the bed, the correct mean value must be chosen to use with the integrated mass transfer equation. There are two, commonly used, limiting values for the mean driving force. If complete mixing exists between the inlet and outlet of the bed, then the correct mean driving force is equal to the driving force corresponding to the outlet gas and liquid bulk composition. On the other extreme, if true countercurrent flow of the gas and liquid prevails, then the correct mean driving force is the logarithmic mean between the values at the top and bottom of the bed as derive before. For any given problem, the mixed mean driving force is the smallest value and, hence, gives the highest mass transfer coefficient and the lowest height of transfer unit. Conversely, the logarithmic mean is the highest possible driving force, and hence gives much lower transfer coefficients. As very often is the case, the actual flow patterns in this study are intermediate between the extremes of no backmixing and complete backmixing. In such cases it has been more common to calculate the performance using the logarithmic mean driving force, and hence all results of this investigation are base on this choice.

2.2.4 Minimum Fluidization Velocity

Measurements of minimum fluidization velocity for con-
ventional two-phase fluidization velocity are facilitated by the
the existence of a well defined relationship between the pressure
drop across the bed and the flow rate of gas on liquid fluidizing
stream. Such a relation is possible only when the solid particles
exhibit good fluidization characteristics. In turbulent contac-
tors, the packings used are frequently 100 times larger than those
normally found in conventional fluidized beds, and hence no smooth
fluidization can be expected. For this reason, the conventional
method of determining G_{mf} is not suitable.

Chen and Douglas (13) had defined the minimum fluidization
velocity G_{mf} as the maximum gas velocity at which the packed bed
maintains its static height. This definition was consistent with
that commonly accepted for conventional fluidization, because,
the bed height at minimum fluidization may approach the static bed
height for large packing.

With the definition of G_{mf} given, the determination of the
minimum fluidization velocity may be carried out through the
measurement of bed height. Chen and Douglas also showed that the
bed height of turbulent bed contactor varied linearly with the gas
mass velocity for any particular set of packing diameter and liquid
mass velocity. Thus, the linear plot of bed height vs. gas mass
velocity could be extrapolated to the point of bed height equal to
the static bed height, the abscissa of this point, according to

the definition of G_{mf}, was the minimum fluidization velocity. For gas flow through a bed of irrigated packings, the pressure drop equation of Leva's type (27) was applicable

$$\frac{\Delta P}{H} = b \frac{G^2}{\rho_f} (10)^{\beta L/\rho_L} \qquad \text{------ (2.20)}$$

For a bed of irrigated packings, the pressure drop for the maximum gas flow which would not cause bed motion may be written as

$$\frac{\Delta P}{H} = (1 - \epsilon)(\rho_s - \rho_f) + \epsilon_{SL} \rho_L \qquad \text{----(2.21)}$$

Expressing the hold-up in terms of w, the fraction of the bed weight, and setting $\lambda = (1 + w)$, Equation (2.21) reduces to

$$\frac{\Delta P}{H} = \lambda (1 - \epsilon)(\rho_s - \rho_f) \qquad \text{----- (2.22)}$$

At incipient fluidization Equation (2.20) and (2.22) may be equated to give

$$\log G_{mf} = k L + n \qquad \text{----- (2.23)}$$

with k and n given by Equation (2.24) and (2.25) respectively

$$k = \frac{\beta}{2 \rho_L} \qquad \text{----- (2.24)}$$

$$n = \tfrac{1}{2} \log \frac{\lambda}{a} (1 - E_{mf})(\rho_s - \rho_f) \qquad \text{-- (2.25)}$$

Equations (2.24) and (2.25) indicated that for fixed fluid properties, k was a function of d_p, and n was a function of d_p and L. But Chen and Douglas had studied and indicated that k

was approximately constant and n depends on d_p only and they presented the correlations for the minimum fluidization velocity as

$$G_{mf} \quad = \quad 0.0599 \; D_p^{1.15} \; 10^{rL} \qquad \text{------ (2.26)}$$

with $r = -0.38115$. When $L = 0$ equation (2.26) reduces to $G_{mf} = a \; d_p^{1.15}$. The latter relation may be compared with $G_{mf} = a \; d_p^{\eta}$ with η varying between 1.2 and 2.0 for conventional gas solid fluidization. The similality between these two relations tends to indicate that despite the presence of the additional liquid phase, the minimum fluidization velocity in a turbulent bed was still affected by packing diameter in much the same way as in the case for gas-solid fluidization.

2.2.5 Gas Hold-up (ϵ_G)

The gas hold-up was determined by directly measuring the height of the aerated bed and that of bed without aeration. The average fraction gas hold-up can be given as

$$\epsilon_G \quad = \quad H_G / (H_G + H_L + H_p) \qquad \text{------ (2.27)}$$

Kito, et al., (18) had studied the effects of packing density, liquid viscosity, static bed height and liquid velocity and derived an empirical equation for the fully fluidized mobile bed. This equation was expressed as below

$$\epsilon_G \quad = \quad 0.19 \left(\frac{D_p U_G^2 \rho_L}{\sigma} \right)^{0.11} \left(\frac{U_G}{(gD_p)^{0.5}} \right)^{0.22} \text{------ (2.28)}$$

2.2.6 Liquid Hold-up (ϵ_L)

Chen and Douglas (13) formulated an empirical correlation for ϵ_{SL} with D_p and L as

$$\epsilon_{SL} = 9.5 \times 10^{-2} (L)^{0.6} D_p^{-0.5} + 0.02 \quad \text{---} \quad (2.29)$$

Kito, et al., (18) had showed that in the range from the minimum fluidization velocity to the flooding velocity H_L, the amount of liquid retained per unit cross-sectional area of the bed, was not affected by the gas velocity but was proportional to $U_L^{0.64}$, $L^{0.16}$, $\sigma^{0.34}$, $H_s^{0.6}$, $\rho_p^{0.18}$, $f(d/D)^{-0.58}$ $D_p^{-0.84}$

The value of H_L was represented as follows

$$H_L = (H_s)^{0.6}(D_p)^{-0.84}(\rho_p)^{0.18}(f\frac{d}{D})^{-0.58}(U_L)^{0.64}$$
$$(\mu_L)^{0.16}(\sigma)^{0.34} \quad \text{--}(2.30)$$

The liquid hold-up defined as $H_L/(H_G + H_p + H_L)$ will be a complex function of process variable such as the gas and liquid velocities, the properties of packing and the characteristics of the supporting grid. Because the value of H_L is independent of the gas velocity. Kito et al. had defined the liquid hold up as H_L/H_s and proposed an experimental correlation for the liquid hold-up in dimensional analysis as follows

$$\epsilon_{SL} = \frac{H_L}{H_S}$$
$$= 12.8 \left(\frac{H_s}{D_p}\right)^{-0.4} \left(f\frac{d}{D}\right)^{-0.58} \left(\frac{gD_p^3\rho_p^2}{\mu_L^2}\right)^{0.09} \left(\frac{U_L^2}{gD_p}\right)^{0.83}$$
$$\left(\frac{D_pU_L\rho_L}{\mu_L}\right)^{-0.34} \left(\frac{D_pU_L^2\rho_L}{\sigma}\right)^{-0.34} \quad \text{------}(2.31)$$

where H_S/D_p was a measure of the number of expansion contraction cycles which the gas must undergo in passing through the bed giving rise to substantial form drag on both liquid and packing.

The liquid hold-up, ϵ_L was correlated with ϵ_{SL} as follows.

$$\epsilon_L \quad = \quad \frac{H_L}{H_S} \cdot \frac{H_S}{H}$$

$$= \quad \epsilon_{SL}/(H/H_S) \qquad \text{------- (2.32)}$$

The term H/H_S in the denominator means the expanded bed height Kito et al.(18) proposed the following correlation for the expanded bed height.

$$H/H_S \quad = \quad (1 + \epsilon_{SL} - \epsilon_{SP})/(1 - \epsilon_G) \qquad \text{------ (2.33)}$$

From equation (2.32) and (2.33), ϵ_L could be respresented as

$$\epsilon_L \quad = \quad \epsilon_{SL}(1 - \epsilon_G)/(1 + \epsilon_{SL} - \epsilon_{SP}) \qquad \text{------ (2.34)}$$