CHAPTER I

SURVEY OF LITERATURE ON MASS TRANSFER EQUIPMENT
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The use of absorber in the chemical industry started so long ago that the use of certain types of conventional equipment has become an established practice in the industry. Also for special reasons, some less conventional types of absorbers have been studied either to obtain basic information on the absorption process or to develop a more efficient type of equipment. The most popular and conventional gas-liquid contacting devices for absorption of gases can be classified as packed towers, bubble cap towers, spray towers and agitated vessels and film absorbers. Less conventional devices include simple stopcock type absorbers, vertical wire type absorbers, venturi scrubbers, pulsating columns, gas lift absorption columns, rotary membrane columns, spiralling film columns, and cyclonic type diffusion apparatus. All of the above types have not been accepted commercially. Commercial acceptability of an absorber depends upon the achievement of high rate of mass transfer at acceptable and optimum cost. This objective is, in general, attained by increasing the area available for mass transfer and by suitable adjustments in hydrodynamic variables which control the resistance offered by the gas film or the liquid film or both.
1.1 Conventional Equipment

(A) Packed Towers:

The first major type of conventional equipment is packed tower which consists of a hollow shell, usually cylindrical, filled with any one of a number of industrial packing materials such as raschig rings, double spiral rings, multibeam support plates, spray pack, panapak, helipak, heligrid, berl saddles, etc. The operation is in almost all cases countercurrent: the gas moving upwards through the tower and the liquid descending in cascade fashion over the packing. In its downward course the liquid spreads around the packing in the form of a thin film, thus providing an increased interfacial area. The time of contact also increases because of the tortuous path taken by the liquid as well as the gas. Best results are obtained when the pressure drop in the tower is low enough to preclude any possibility of flooding. The tendency of the liquid to form preferential channels is checked, if necessary, by redistribution over the entire cross sectional area; the corrosion and thermal resistances of the packing material are high; the ratio of the effective wetted surface to the total specific surface of the packing is large; and the pressure drop due to skin friction on the surface of the packing is high compared to that due to turbulence. Cost optimisation is obtained by balancing the contradictory
requirements of height/diameter ratio and packing size/diameter ratio necessary for these, and other desirable features.

In the absence of actual operational data, the required height of a packed tower may be approximately calculated from the number of transfer units required for a given operation and the average height of a transfer unit. The number of transfer units required for a given absorption duty is a measure of the difficulty of the process. The height of a transfer unit in a packed tower may be estimated from empirical equations available in published literature. The general form of the equation which is applicable when the rate of absorption is not limited by a slow chemical reaction, and when the resistance to mass transfer is due to the liquid film as well as the gas film, is:

\[
(HU)_{OG} = H_G + \frac{m Q_m}{L_m} H_L \\
(HU)_{OL} = H_L + \frac{L_m}{m Q_m} H_G
\]  

The use of this general form of equation is not limited to packed towers only but also is common to most types of absorption equipment. Experimental methods have been devised for determination of \( H_G \) in gas-phase resistance controlled processes and \( H_L \) in liquid-phase resistance controlled processes. In the
former case, these include study of absorption of very soluble gases, vaporisation of pure liquid into inert gas streams, absorption of a gas in a liquid with which it reacts chemically in a rapid irreversible fashion with little or no equilibrium back pressure etc. For packed towers in particular, data from the books by Perry and Sherwood and Pigford indicate that in a situation in which diffusion is controlled by the gas-phase resistance, the height of a transfer unit is dependent on both gas and liquid flow rates. At a constant liquid rate, this height increases as the gas rate is increased until the loading point is reached. Below the loading point the height is proportional to $G^n$, the value of 'n' being usually between 1/4 and 1/2. At a constant gas rate the height decreases as the liquid rate is increased owing to the improved wetting of the packing, and is nearly proportional to $L^{-1/2}$. To take a specific instance out of many which are available in published literature, it has been reported that data covering a range of liquid rates from 168 to 6100 lb/(hr)(ft²) can be represented by the equations:

$$H_0 = \frac{1.01 G^{0.31}}{L^{0.33}} \quad \ldots \ldots (1.3)$$

Reported data for the situation in which diffusion is controlled by the liquid-phase resistance in packed towers may be illustrated by the results obtained by Sherwood and Holloway; these are represented for a wide range of liquid flow rates by the equations:
\[ H_L = \frac{1}{\alpha} \cdot (L/m_L)^{2} \cdot \beta \cdot \gamma \cdot \delta \cdot \varepsilon \]  

(1.4)

For different types of packings, values of \( \alpha \) vary between 80 and 550, and of 'n' between 0.22 and 0.46.

Cases are, however, not infrequent when the rate of diffusion is significantly controlled by both films. In such situations a general equation is available provided that the equilibrium relationship can be represented by a straight line which need not necessarily pass through the origin. If \( m' \) is the slope of this line in the required concentration range, and \( \beta', \gamma, \delta, \varepsilon \) and \( n \) are constants, the equation is:

\[ (HTU)_{OL} = \gamma L^{n} + \frac{29 \beta L^{1-q'}}{18 m' G \cdot L \cdot p} \]  

(1.5)

A system of this type which is of considerable interest is the sulphur dioxide - water system. The reported values of various constants are: \( m' = 32.4, \beta = 1.24, \gamma = 0.37, \delta = 0.30, \) \( q' = 0.25, \) and \( n = 0.18. \) Such equations are of limited utility, and actual operating data for design are obtained from pilot plants whenever feasible.

(B) **Plate Towers:**

The second major type of conventional equipment includes bubble cap towers, sieve plate towers and porous plate towers,
and is analogous to plate tower rectifying column in distillation. The usual bubble cap column is cylindrical in shape and is divided over its length by a number of horizontal plates. The liquid usually moves across each plate from one side to another and spills over an overflow weir to the plate below. The gas flows upwards through slotted bubble caps and forms a large number of bubbles which rise through the pool of the liquid on each plate. The slots are usually rectangular in shape, but may be trapezoidal, square, diamond shaped, triangular, oval or round. It was assumed that these promote better break-up of gas into jets or bubbles. It was however found that the type of teeth and slots has little influence on mass transfer. The main purpose of the teeth or slots is to eliminate a one sided gas escape from underneath the cap in case there is deviation of the lower periphery from the horizontal due to skewed assembly. The plate trays and bubble caps are designed to produce adequate mixing on each plate, curtail short circuiting of liquid directly from one overflow pipe to another, avoid clogging, reduce unproductive pressure drop, inhibit weeping or dumping of liquid, minimise foaming and entrainment, and ensure a stable operation. Different hydrodynamic conditions are observed, depending on the gas velocity on the bubble trays. The transition from one set of conditions to another takes place gradually. The conditions change in different ways with different types of trays. Various
authors have studied the phenomena under different names. Kasatkin et al.\textsuperscript{116} suggested that depending on reduced gas velocity three main conditions can be observed which can be termed as (i) non-uniform, (ii) uniform gas jet and (iii) splash conditions. Entrainment caused by excessive gas flow rates can result in back-mixing and reduction of efficiency. There is, however, evidence\textsuperscript{103,161,156,177} to show that entrainment plays no important role at any gas velocity below that at which the column ceases to function because of excessive foaming or blocking up of an overflow pipe due to gas pressure. Back-mixing is also encountered in sieve trays. On cross flow trays the efficiency decreases as a result of mixing. Danilychev\textsuperscript{44} proposed the following equation for mixing on sieve tray:

For vapour/gas velocity, \( U = 1 - 7 \text{ m/hr} \) and the clear liquid height \( h_0 < 25 \text{ mm} \)

\[
P_{\text{L}} = 0.22 \cdot \text{Re}_L^{0.6} \cdot \text{Re}_g^{-6.5}
\]

where gas Reynolds number is calculated from the reduced gas velocity and the decisive parameter for all three generalized variables is the height of the clear liquid \( h_0 \).

Planovskii\textsuperscript{159,160} used \( h_0 \) as a decisive dimension for Sherwood number in bubble cap trays.

As a bubble cap column is essentially a stage-wise contacting device, it is usual to express its performance data
in terms of local efficiency at a point within a plate, or efficiency of a plate (Murphree efficiency) or an overall efficiency which equals the ratio between the number of theoretical plates and the number of actual plates used in the column. While some empirical relationships between design dimensions and efficiency have been reported in the literature, the theoretical model for each plate is essentially that of a short spray column in which bubbles are formed at the lower end, travel through the liquid and finally burst into the gas space above the liquid. Primarily because of higher viscosities of liquids at lower temperatures, the overall efficiency of an absorber column may be as low as 8 to 10% in contrast to an efficiency of about 65% in distillation columns.

(C) Spray Towers:

The third major type of conventional equipment is the spray towers. Though the term spray tower is usually applied to a column in which a liquid is introduced in the form of droplets by means of spray nozzles or other atomizing devices into an atmosphere of circulating gas, it is also used frequently for the analogous operation in which bubbles are introduced into a large body of liquid in order to dissolve a soluble gas. Depending upon the degree of agitation produced in the continuous phase, a spray tower may be equivalent to one or more theoretical stages. The spray is formed by the
collapse of high velocity gas or liquid jets at the nozzles. The interfacial area depends on the size of droplets or bubbles, but there is an effective upper limit on the amount of the interfacial area in a given equipment due to coalescence of drops or bubbles. As shown by Johnston and Williams, this area may be surprisingly small: for a liquid rate of 500 lb/(hr)(ft²) and countercurrent gas velocity of 5 to 10 ft/sec it may be as little as 0.1 to 0.2 ft²/ft³ of tower volume. At high continuous phase velocities, the problem of entrainment may be serious, but may be solved by such devices as centrifugal sprayers.

Mass transfer equations applicable to spray towers are complex and of a considerable variety depending on whether the continuous phase is liquid or gaseous, whether the controlling resistance lies in the continuous phase or in the dispersed phase, whether the continuous phase is turbulent or not, and whether there is internal circulation or oscillation associated with the dispersed phase. There may be further differences due to the wall effect, and the effect of presence of other drops or bubbles. In addition, different equations apply during formation, travel and coalescence stages. During the formation stage, the best equation appears to be.
where $D$ is the diffusivity in the continuous phase, $t$ is the time of formation, $d$ is the final diameter, and $m$ is the distribution ratio between the continuous phase and the dispersed phase. The coefficients $a_{n,n'}$ are defined as:

$$
\bar{E} = \sum_{n=0}^{\infty} \sum_{n'=0}^{\infty} a_{n,n'} \left\{ 5.1709 \frac{D^{1/2}t^{1/2}}{d} \right\}^{n'} \left\{ 9.9795 \frac{D t}{d^2} \right\}^{n'} \quad \ldots \ldots (1.6)
$$

and equal zero when $n$ or $n'$ is negative. $E$ is the fraction of equilibrium achieved in the concentration in the dispersed phase. For the coalescence stage no reliable equations are available. For design purposes it is usual to assume arbitrarily that the degree of equilibrium achieved during this stage is of the same order as during formation. For the travel stage the popular form of equation is:

$$
Sh = a + b Re^x Sc^y \quad \ldots \ldots (1.8)
$$

with the reported values of constants $a$ and $b$ and of indices $x$ and $y$ varying widely for different systems and variable ranges. Proeselting found, for instance, that his extensive data could be correlated by taking $a = 2.0$, $b = 0.532$, $x = 0.5$, $y = 0.25$. 

$$
\bar{E} = \sum_{n=0}^{\infty} \sum_{n'=0}^{\infty} a_{n,n'} \left\{ 5.1709 \frac{D^{1/2}t^{1/2}}{d} \right\}^{n'} \left\{ 9.9795 \frac{D t}{d^2} \right\}^{n'} \quad \ldots \ldots (1.6)
$$
$y = 0.33$. Most of such data are, however, derived from study of single drops and are not applicable in this form to industrial spray towers in which the wall effect and the effect of the presence of other drops or bubbles may be considerable. Pigford\textsuperscript{158} isolated the wall effect (the liquid collected at the wall was 60 to 80\% of the total sprayed), and found that in its absence the value of $(NTU)_{OG}$ increases with gas velocity and $(NTU)_{OG}$ increases with liquid rate.

Spray chambers have been found to be reasonably efficient for operations in which the liquid phase resistance predominates, in contrast to what has been frequently assumed. The gas appears to be so thoroughly mixed in the continuous phase that it has almost the same concentration throughout, especially during air conditioning or deaerating operations. For such cases, the following equation\textsuperscript{111} is said to represent the data adequately:

$$(NTU)_{OG} = 62.8 \frac{L}{G_{e}^{0.9}} \frac{d_{e} P}{S d_{d} \text{Sc}^{2/3}} + 0.0064 \frac{A_{w} P}{G_{e}^{0.87} \text{Sc}^{2/3}}$$

(1.9)

In this equation, the first term on the right hand side accounts for the effect of the spray, and the second term accounts for the wall effect, $A_{w}$ is the wetted area of the wall, $P$ is the total pressure, $G_{e}$ is the mass velocity through tangential
entrance, $S$ is the cross sectional area of the gas entrance nozzle, $Sc$ is the Schmidt number for the gas, $d_t$ is the diameter of the spray chamber, $d_d$ is the mass median diameter of spray droplets, and $L$ is the length of the chamber, all values being expressed in the British Engineering System.

(D) **Agitated Vessels:**

The fourth major type of conventional equipment is the agitated vessel. Agitated vessels find their application in situations in which prolonged contact times are essential. These are mostly operated batchwise and are limited to the equivalent of one theoretical stage. The gas is dispersed in the liquid through porous gas spargers, and the stirrer itself may act as a suction device, though it is usual to supply the gas under pressure. The power consumption depends on the size and configuration of the vessel, number and properties of phases and the type of impeller used. The final bubble size produced (and also to some extent, the rate of mass transfer) depend largely upon turbulent conditions at the impeller periphery. The limiting and optimum size and speed of the impeller are determined by the need to increase the turbulence at the periphery without wasting power in producing unnecessarily excessive turbulence in the bulk of the liquid.
Calderbank and co-workers\textsuperscript{26,27} have shown that a first approximation to $Sh$ for agitated vessels may be obtained by either of the two following equations:

\begin{align*}
Sh &= 0.31 \ Re^{2/3} \ Sc^{1/3} \ Fr^{-1/3} (\Delta \xi / \xi)^{1/2} \quad \ldots \quad (1.10) \\
Sh &= 0.42 \ Re^{1/3} \ Sc^{1/2} \ Fr^{-1/3} \quad \ldots \quad (1.11)
\end{align*}

where $Sh$, $Re$, $Sc$ and $Fr$ denote respectively the Sherwood, Reynolds, Schmidt and Froude numbers.

The type of equation which is more directly applicable to gas liquid dispersions in agitated vessels is of the form\textsuperscript{26}:

\begin{equation}
Sh = a \ Sc^{1/2} \ Re^{2/3} \quad \ldots \quad (1.12)
\end{equation}

When the presence of surface active agents renders the bubbles rigid, this form of equation changes to\textsuperscript{27}:

\begin{equation}
Sh = a + b \ Sc^{1/3} \ Re^{-1/2} \quad \ldots \quad (1.13)
\end{equation}

in which the values of the constants appear to vary with the system, and may be of the order $a = 2.0$, $b = 0.79$. In general, increased power input does not have a commensurate effect on the value of the mass transfer coefficient; a three-fold increase in power input may, for instance, serve to increase the mass transfer coefficient by one-fourth.
(5) **Film Absorbers:**

In film absorbers the gas and liquid come into contact on the surface of the flowing liquid film. On the basis of vertical surface on which the liquid film is formed, film absorbers can be divided into three categories:

1. **Tubular absorber with sheet type packing (plane-о-parallel)** in which liquid flows along both surfaces of vertical plates;
2. **Tubular absorbers** in which the film flows over the internal surface of vertical tubes;
3. **Absorber with upward movement of film.**

The first two types of apparatus generally operate under *counter current* flow conditions of the gas and liquid. However, they may also operate under *descending* or *ascending parallel flow* conditions. The absorber with ascending film is usually used when heat is simultaneously removed in absorption process. In film type apparatus with descending and ascending parallel flows, the separate flows of the two phases may lead to splash entrainment conditions if the gas velocities are high. Zhivaikin and Volgin introduced *dimensionless parameter, p* for such conditions. The parameter is given by:

\[
p = \frac{V \mu_L}{\sigma} \quad \ldots \ldots \ (1.14)
\]

where \( V \) is the velocity of gas in meters/sec, \( \mu_L \) is the liquid viscosity and \( \sigma \) is surface tension. Lastovtsov et al. introduced
their studies on ascending parallel flow showed that
entrainment depends to a small extent only on the gas
velocity and liquid viscosity so that

\[
\frac{q}{\Gamma} = 0.039 \frac{Re}{u_D} p^{0.35}
\]

\( q \) = linear volume wetting density \( m^2/sec \)
\( \Gamma \) = linear wetting density \( Kg/m/sec \)
\( p \) = over all pressure in bar.

1.2 Less-Conventional Equipment

The development of less conventional equipment has
been based on the study of the effect of useful physical
variables on mass transfer rates. Such equipment have
usually tended to be confined to simple geometries for which
appropriate hydrodynamic models can be visualised and
interfacial areas can be calculated in a form suitable for
theoretical analysis. In some cases the main purpose has
been to obtain better overall efficiency of contacting
equipment by accentuating the effect of the useful variables
and decreasing the pressure drop through the equipment. These
have not usually been accepted by the industry for large scale
absorption processes.

(A) Stop-Cock Type Absorbers

A simple stopcock type absorber, Fig. 1.1, has been
reported by Takahashi and co-workers\(^{190}\) for measurement of
diffusivities of gases in small amounts of absorbents. It consists of a horizontal inner cylinder, A, and an outer sleeve, B, of about 8 cm. length. The inner cylinder has a slit, C, 4 cm. long and of 8 mm. diameter, and a slit, D, slanting to axis (x - x'). The outer sleeve is fitted with nozzles, E and E', for flow of liquid into slit, C, and nozzles F and F', for flow of gas into slit, D. The nozzles G and G', fitted at the ends of the inner cylinder, are for circulating a constant temperature fluid in the cylinder. During operation, this cylinder is fixed such that the axis of slit, C, is vertical. The outer sleeve, which is free to rotate, is turned from its initial position Fig. 1.1a to the operational position Fig. 1.1b. The liquid in this slit thus contacts the gas at the lower liquid surface. The liquid retains its position due to surface tension. Neither the liquid nor the gas is in motion. Results of experiments with chlorine absorption in water and solution of sodium chloride and hydrochloric acid in this apparatus indicate that absorption rates follow the penetration theory for long contact time provided the effect of free convection of liquid near the interface is negligible. Presence of a chemical reaction, as in the absorption of chlorine in sodium hydroxide solution, leads to more complicated phenomena. With a relatively longer time of contact, the stopcock type of absorber has an advantage over the laminar liquid jet absorber and the short wetted wall
1.1 Stop Cock Absorber
1.2 Vertical Wire Absorber
column for study of fundamental properties of gas-liquid systems.

(B) Vertical Wire Type Absorber:

Another experimental unit is a vertical wire type absorber, Fig. 1-2. It consists of an oil supply reservoir connected to a 13 gauge syringe needle. A needle valve controls the oil flow. The exit end of the syringe needle is cut perpendicular to the needle axis and is machined to a smooth surface. A rubber stopper closing the absorption chamber has entry holes for the needle, for a gas outlet tube and a thermometer. Another rubber stopper has holes for a take-off tube, a gas inlet tube and a manometer tube. A piano wire is stretched taut through the syringe exit to ensure angular symmetry of oil flow along the wire. High mass saturation rates are observed in this apparatus due to large interfacial area/liquid volume ratio, a high degree of turbulence caused by friction between oil drops and the wire, and a large contact time. It has been reported that the results obtained can be correlated according to the model suggested by Handlos et al.

(C) Venturi Scrubber:

An equipment of special use in desorption of liquids of low surface tension is the venturi scrubber, in which
The pressure losses are minimised. The major portion of absorption in the scrubber takes place between a point which is about two-thirds of the expansion section downstream from the point of injection and a point which is a few diameters further downstream. The venturi scrubber is not very useful for correlation of experimental and theoretical results as the interfacial area is difficult to estimate in this equipment. The area changes with pressure because there is an increase in pressure and a decrease in bubble size as the diameter of the venturi walls increases.

(D) Pulsating Column Absorber:

The principle that efficiency of a mass transfer equipment may be increased by pulsations in the vapour phase is utilised in the pulsating column by Zidkowski\textsuperscript{239}. It consists of a horizontal steel cylinder within which there are two discs moving with a reciprocating action obtained by means of a cam shaft of an adjustable stroke drive connected to the disc shaft. A clearance is provided between the discs and the casing. Harbeum and Iboughton\textsuperscript{29} studied the effect of pulsations on the process of liquid phase absorption by sparging carbondioxide into water. They found that the maximum increase in the rate to the order of 70\% can be obtained. Breuszaider and Pasiuk\textsuperscript{20} studied the effects of pulsations in the liquid phase on the absorption process. They obtained rate
increase between 20 to 99%. Harbaum et al. showed that efficiency increases with frequency of pulsation until the frequency of 9.3 hertz is reached. The maximum value appears to be independent of flow conditions of the two phases. A detailed study of the effect of pulsations under various flow conditions of the liquid film showed that in the case of laminar flow the intensification is almost independent of flow conditions. On the other hand in turbulent flow with ReL exceeding 1,000, the intensification is influenced by flow conditions to a certain extent. The flow conditions in the gaseous phase influence the process to a greater extent. When the latter increases, the intensification of the process decreases as a consequence of an increase of the overall coefficient of absorption under new pulsating condition.

(5) Gas Lift Absorbers

Gas lift absorption columns have been used by Von Heuven and Co-workers for absorption studies. The flow around a bubble in a narrow, vertical tube can generally be approximated by considering two sections. In the upper part, a potential flow may be assumed to exist as against the lower part where there is a laminar, viscous flow. Davies and Taylor have investigated the rate of rise of the gas-liquid interface in a tube which was closed at the top and from which liquid was drained by gravity, and have shown that the flow around
Diagram of rotary column

1. Inner cylinder
2. Outer cylinder
3. Flat disc
4. Liquid outlet
5. Support
6. Water feed
7. Solid ring
8. Outer ring
9. Liquid outlet
10. Gas inlet
11. Liquid outlet
12. Contact ball
13. Bracket
14. Support
15. Flat disc
16. Gas outlet

1.3 Rotary Membrane Absorber
the top of the rising interface is approximately potential over a height of about four times the tube diameter. The mass transfer between bubbles and the surrounding liquid is complicated by the fact that the surface elements which are formed at the bubble top are stretched during their movement. This results in a flow in the direction of the surface. This flow carrying fresh liquid from the bulk increases the absorption rate. The acceleration at the rear is zero or even negative. In this region, viscous forces or a negative gradient of surface tension in the direction of the flow plays its role. Thus the liquid is decelerated when approaching the end of the bubble. The absorption at the top of the bubble is large due to stretching of the surface.

(F) Rotary Membrane Absorber:

Rotary membrane columns by Pribyl and Trawinski shown in Fig. 1.3 offer an interesting possibility in the study of absorption processes. The working space in their case is the space between two coaxial cylinders. For experimental investigations both may be stationary, one may be rotating, or both may be rotating in the same or in opposite directions. Adjustment of speed of each cylinder provides another experimental variable. The centrifugal force resulting from rotation stabilizes the liquid film flowing down the inner wall of the outer cylinder. It can be expected that this will dampen the
Diagram of flow in a countercurrent apparatus of the
spiral type. 1 - Liquid inlet, 2 - gas inlet.

Diagram of model apparatus. 1 - Cyclone column, 2 -
drawoff receiver, 3 - pump, 4 - solenoid valve, 5 - regulation
valve for water circulation, 6 - regulator for outlet of liquid
inlet, 7 - liquid outlet from the column, 8 - box, 9 - outlet
valve, 10 - regulation valve for air circulation, 11 - air
receiver, 12 - CO₂, air, receiver, 13 - pressure-reducing
valve, 14 - rotameter, 15 - sampling valve, 16 - flow orifices.
17 - pressure measurement, 18 - temperature measurement.

Air distributor.

Material. 2 - liquid outlet, 3 - pressure tube.
4 - Sampling valve, 5 - cylindrical glass column.
surface waves and will allow a determined flow even at high relative velocities of the two phases. It has been found that when the number of revolutions relative to each other changes, the film thickness also varies, the results being less reproducible when the direction of rotation is the same for both cylinders. Absorption studies show that along with the character of the film, which is dependent on the relative rate of revolution, the ratio of resistance of the liquid phase to that of the gaseous phase undergoes a substantial change. At high relative rates of revolution, it has been reported that gas phase resistance becomes practically negligible.

(3) Cyclonic Absorber:

A cyclonic type diffusion apparatus has been reported by Koscky\textsuperscript{123} to produce very high rates of wetting and coefficient of mass transfer. In this apparatus the liquid is introduced through a tangential inlet of a rectangular cross section attached to the inner wall of a hollow vertical cylinder. The inlet is inclined with respect to the cylinder surface. The velocity of the entering liquid is adjusted to form a turbulent, downward flowing rotating film on the inside wall of the cylinder. The gas is introduced through a tangential inlet at the bottom side wall of the cylinder. The gas and the liquid flow countercurrently, and the gas rotates
in the same sense as the liquid film. In a modification of this apparatus, Kosseck⁷ has used two cyclonic units concentrically. The liquid is introduced tangentially on the inside surfaces of the two cylinders. The gas which contacts the liquid rotating on the walls of the outer cylinder flows through the annulus between the two cylinders. Thus, the arrangement results in using also the inner wall of the inner cylinder for contact between the two phases.

(H) **High Surface Rotary Film Reactor:**

A high surface rotary film reactor has been designed by Pribyl and Trawinski ¹⁶⁰ to obtain greatly increased interfacial areas, though at the expense of decreased volumetric throughput. This apparatus consists of rolling cylinders supported by special bearings at the outer ends of spokes radiating horizontally from the hub of the vertical rotor axle. The roller cylinder bearings support the roller cylinder axles in vertical and horizontal directions, while the bearings themselves can move radially with respect to the hub. The whole assembly is placed inside an outer cylinder. Thus, when the rotor is spinning, the rolling cylinders are pushed against the inside surface of the main cylinder by centrifugal force. The fluid guided against the inside surface of the outer cylinder is dispersed between this surface and the surfaces of the rolling cylinders. Compared to conventional
rotary film reactors, this arrangement provides more wetted surface for the same volume.

(I) Spiralling Film Absorber:

Alimov has shown that spiralling of gas and liquid streams can result in substantial improvements in mass transfer coefficients. In an apparatus used by him, a thin liquid film drawn along by a spiral gas stream was stabilised along the inner wall of a vertical cylindrical tube. The gas stream was rotated by allowing it to flow into a cylindrical vortex tube, closed at one end, through tangential slits. The liquid was fed into this spiralling gas stream through a nozzle to form a film on the tube wall. In the operation of this equipment, liquid drops entrained by the gas stream are thrown into the tube wall by centrifugal force and are deposited in the form of a thin layer. This liquid layer is carried by the spiral gas stream along a helical path to the outlet. A continuous spiral film over the entire surface of tube is obtained by feeding the gas and the liquid over a length comparable to the pitch of the spiral stream.

Cane and Bugared have studied the increase of surface offered to mass transfer by a spiral film with respect to a cylindrical falling film, and have found that
the surface increase may be as much as three times under
the action of waves or ripples. By tangential introduction
of the liquid, one may increase the surface to double the
value at low entering velocities, and to five times the
value at higher velocities. In the case of a falling film,
particularly in the transition zone (250 < Re < 1100), the
speed of surface renewal does not grow much with an increase
in the Reynolds number. With tangential introduction of
the fluid, the turbulent regime is established more quickly,
and the speed of renewal grows rapidly with the flow rate.
The coefficients of mass transfer are consequently higher.
Hydrodynamic studies indicate that tangential introduction
gives rise to new velocity components (tangential component
and, probably, radial component). Also, the decay of the
tangential impulse along the column length involves a
continuous variation in the characteristics of the film. The
mean thickness of the spiral film is slightly less than
that of the falling film. The radial velocity profile is
practically uniform.