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FLUE GAS DESULFURIZATION BY ROTATING BEDS

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PART I

**OPERATING CHARACTERISTICS OF ROTATING
FOAM METAL BEDS FOR FLUE GAS
DESULFURIZATION**

BY

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OPERATING CHARACTERISTICS OF ROTATING FOAM METAL BEDS FOR FLUE GAS DESULFURIZATION

1. INTRODUCTION

Flue gas desulfurization (FGD) is an important problem, particularly considering the utilization of the high-sulfur coals for power generation. The economical separation of sulfur dioxide from flue gas is difficult, as the volume of gas to be treated is large, about 3000 actual cubic feet per minute per megawatt generating capacity, and the concentration of sulfur dioxide in the flue gas is small, on the order of 0.5%.

Absorption processes, the scrubbing of sulfur dioxide by counter-current contacting of the flue gas and solvent in a packed tower, is one approach. Substantial improvements in scrubber performance have been achieved through superior packing design, but that technology is now mature, and further substantial gains in performance are unlikely. The perception has been that the major improvements in the economics of FGD would most likely be forthcoming in novel process ideas, careful selection of solvents, and care in the conservation of energy.

Process intensification, the use of high gravitational, electric acoustic, or magnetic fields to enhance reaction and separation processes, is a very interesting and promising research and development area. In particular, heat and mass transfer can be intensified by carrying out processes in high gravitational fields. Imperial Chemical Industries (ICI) pioneered the development of rotating packed beds, which are one to two orders of magnitude smaller in volume than conventional packed columns. Heights of theoretical plates are much reduced, typically from about two feet to four feet in conventional column, operating at 1 g. to one-half to one inch in rotating packed beds, operating at 300 g's. Benefits include:

- much smaller equipment and supporting structures
- much reduced solvent inventory

- shorter start-up and shut-down times
- faster response with improved process control
- reduced pressure drop

The benefit of much reduced solvent inventory is particularly important, as there is the possibility of using solvents that hitherto have been considered too expensive for use in FGD scrubbers. With rotating packed beds, there is also the possibility of substantial gains in scrubber performance and a possible breakthrough in the economics of flue gas desulfurization. To date, there is little in the literature about the operating characteristics of rotating beds, particularly high performance foam metal beds.

This work is concerned broadly with the operating characteristics of rotating beds. The objectives of this work are to 1) characterize the pressure drop in rotating bed, 2) investigate the effect of gravitational acceleration on volumetric mass transfer coefficients of SO₂ absorption, 3) determine the residence time distribution of the liquid phase in rotating beds, 4) investigate the required power consumption of rotating beds.

2. PREVIOUS WORK:

2.1. Centrifugal Tray Absorbers:

The concept of contacting liquids and vapors in a centrifugal field was reported in early studies of Podbielniak [1,2]. The centrifugal absorber consisted of perforated cylindrical sieve trays for gas-liquid contacting. The perforations were dimpled inward to provide weirs for retaining some of the liquid. The perforations were off-set axially from tray to tray to prevent any direct short circuiting. This device was later improved by Todd [3,4] to provide more effective contacting for the phases, and to increase the stage capacity by utilizing a number of units operating on a vertical shaft. Similar work was reported by Chambers and Walls [5]: their centrifugal absorber consisted of a set of rings attached to the rotating plate to generate climbing liquid films and sprays repeatedly.

2.2. Rotating Packed Beds:

Vivian et al [6] studied the effect of gravitational field on gas absorption in a packed column by mounting a cylindrical packed bed on the horizontal arm of a large centrifuge. The gravitational force was varied from 1 to 6.4 times the gravity force. For their system configuration, the liquid phase volumetric mass transfer coefficient varies with the gravitational force to the 0.41 to 0.48 power.

In the late 1970's, there appeared in the European and U.S. patents literatures a series of patents assigned to Imperial Chemical Industries (ICI London) [7-10], introducing the HIGEE packed bed rotor for absorption and distillation applications. The HIGEE packed bed was operated at mean acceleration fields of 200 to 500g, and the reported volumetric mass transfer

coefficients were up to 100 times larger than the values obtained in conventional packed columns. The reported volumetric mass transfer coefficients varied with the gravitational force to the 0.14 to 0.54 powers. ICI, in spite of extensive tests, published very little information. Glitch Inc. (Dallas), which has acquired worldwide marketing rights, also reported a test history of HIGEE with various systems at a wide range of process conditions and physical properties [11]. Glitch Inc. already installed a unit two years ago for the U.S. Coast Guard at a facility near Traverse City, Michigan, to air-strip aviation fuel from contaminated ground water, and field tests are underway for an undisclosed natural-gas producer in the U.S. [12]. However, no publications are cited in the open literature about the results of the tests.

Recently, Dudukovic et al reported theoretical and experimental work on rotating packed beds [13,14,15]. They reported separately the effect of rotational speed on gas-liquid interfacial area and liquid side mass transfer coefficient to explain the overall enhancement of volumetric mass transfer coefficients in rotating packed beds. Their findings show that the volumetric mass transfer coefficient increases with the rotational speed as a fractional power of 0.61 to 0.89 which is in agreement with the data of Vivian et al and within the range estimated by ICI. They showed that the effective surface area increases with the body force with 0.14 to 0.28 power (contribution of effective surface area to the enhanced volumetric mass transfer coefficient). Their theoretical framework was based on film flow on rotating blades and rotating disks. Their flooding tests with air-water system showed that the Sherwood correlation for the dumped rings is

conservative by 40–70%. The bed in this study was 3 mm glass beads which had a specific surface area of $120 \text{ m}^2/\text{m}^3$ and a porosity of 0.4.

In all the above work, no attempt was made to model the pressure drop characteristics of the rotating packed bed, nor is there any information available about residence time distribution on rotating packed beds. No work has been reported in the literature on power requirement of rotating packed beds.

3. EQUIPMENT and PROCEDURE:

3.1. Schematic of the Rotor:

Figure 1 shows a simplified schematic of the rotating bed. The liquid enters from the inner pipe of the stationary double pipe (2), is sprayed by the distributor (1) on the inside edge of the bed, and then is thrown out into the housing (3) by the centrifugal force. The gas enters the stationary housing and, due to the imposed pressure gradient, flows countercurrent to the liquid through the rotating bed. After leaving the bed, the gas exits through the double pipe (4). The liquid collects in the bottom of the housing wall and forms a liquid seal (5) in a trap.

3.2. Packing:

The packing for the rotating bed is a one piece torus-shaped unit (6), made of an aluminum foam metal. It is dimensionally stable in the high gravitational fields. The outside diameter is 45.7 cm, the inside diameter is 25.4 cm, and the thickness is 4.4 cm. Three different beds were studied. The beds' specifications are presented in Table 1. Our stress analysis indicates the beds can be operated from 600 to 2000 rpm, giving a gravitational field of 100 to 1000 times gravity (see the Appendix).

3.3. Liquid Distributor:

The liquid distributor is a manifold consisting of twelve distributing tubes. The division of the liquid stream into parts by means of this type of manifold is uneven because it is not possible to keep the liquid pressure perfectly constant inside the main channel [16]. However, for the above specifications, the difference

Figure 1. Schematic of Rotating Bed

TABLE 1. PACKING SPECIFICATION

Specific Surface Area (m^2/m^3)	Porosity	# of Pores/cm	Pore Size
656	0.92	4	0.4
1476	0.92	8	0.2
2952	0.92	16	0.1

in the rate of the flow between two extreme ports is less than 5%. This manifold provided adequate liquid distribution and reduced the chance of the liquid entrainment in the inner edge of the rotor where the flooding is most probable, since the centrifugal acceleration is lowest and the gas velocity highest at this point.

3.4. Rotor Assembly and Seals:

The rotor assembly is mounted on a 5.1 cm diameter shaft, supported by two bearings, in a cantilevered design. A variable speed 5 hp AC motor with a belt drives the rotor assembly.

Seals are required in two locations in the rotor assembly: the shaft (7), and the double pipe assembly for the gas exit (8). Inexpensive silicone rubber radial lip seals were used in both locations.

3.5. Flow Diagram:

Figure 2 shows the flow diagram of the general experimental apparatus. The feed and recycle pumps can supply up to 45 kg/sec-m² for this packing (30 gpm). The compressor can deliver up to 2.5 kg/sec-m² air for this packing (120 acfm). The air can be heated or it can enter the rotor at room temperature.

3.6. Measurements of Operating Parameters:

Both the liquid and vapor flowrates were determined by measuring the pressure drop across restriction orifices (RO). Differential pressure transducers were used to measure the pressure drop across the orifices. The pressure drops across the bed were measured by placing the probes for the transducers at the inlet and outlet of the rotor as shown in Figure 1.

Figure 2. Experimental Apparatus

The torque and rotational speed measurements were made to determine the power required to rotate the beds. A sprocket/pulley torque transducer was connected to the end of the drive shaft to measure the required torque for the rotating bed. A magnetic pick-up tachometer was used to measure the speed of rotation.

For residence time distribution experiments (RTD), concentrated NaNO_3 solutions were prepared and charged to a tracer inlet loop. By a simultaneous half turn of two shut-off valves, the tracer was introduced into the liquid stream inlet. The liquid passes the RTD sensors at the inlet (6) and outlet (7) of the rotor. The electrical conductivity inlet sensor consisted of two 0.158 cm (1/16") stainless steel rods mounted perpendicularly and 180 degrees apart on the body of a teflon tube. The teflon tube serves as one of the distributing tubes for the liquid distributor. The outlet sensor consisted of two 0.158 cm stainless steel rods (1/16" rods) mounted on a teflon trough which is placed very close to the rotor outlet. The purpose of the trough design is to catch the spray from the rotor.

Carbon dioxide concentrations in the liquid were determined by a specific ion electrode. Carbon dioxide concentration in the gas phase were determined by gas chromatography.

3.7. Data Acquisition:

All the flow, pressure, temperature, torque, and rotational speed transducers are interfaced to a computer. The data for the RTD experiments was taken by a digital oscilloscope.

4. PRESSURE DROP CHARACTERISTICS OF ROTATING BEDS

4.1 THEORY:

4.1.1. Dry Packing:

The total pressure drop in rotating packed beds is the sum of pressure drops from two regions: the pressure drop between the stationary housing and the spinning rotor, ΔP_o , and the pressure drop inside of the rotor (across the bed), ΔP_i . Each region will be analyzed.

4.1.1.a. Pressure Drop Across the Bed:

If the pressure drop through the bed is not large, the gas phase can be assumed incompressible. Assuming that the velocity in the axial direction is zero, and the angular velocity does not vary in the tangential direction, then the continuity equation in a cylindrical coordinate for the gas phase, at steady state, may be written as:

$$\frac{1}{r} \frac{\partial}{\partial r} (\epsilon r V) = 0 \quad (1)$$

Here V is the velocity in radial direction, r is the radial distance, and ϵ is the packing porosity. See the Appendix for the selected coordinate system for a control volume.

A general momentum balance can be written as follows:

$$\{\text{rate of momentum accumulation}\} = \{\text{rate of momentum in}\} - \{\text{rate of momentum out}\} + \{\text{sum of forces acting on system}\}$$

The radial momentum balance, at steady state, may be expressed as:

{change in momentum} = {sum of pressure, drag, and centrifugal forces}, while in the tangential direction (θ direction) the momentum balance may be written as: {change in momentum} = {sum of drag and coriolis forces}.

The equations of motion have the following forms:

$$-\rho_g V \frac{dV}{dr} = \frac{dP}{dr} + \rho_g \frac{U^2}{r} + \frac{\text{Drag Force}}{\text{unit volume}} \quad r\text{-direction} \quad (2)$$

$$-\rho_g V \frac{dU}{dr} = \rho_g \frac{UV}{r} + \frac{\text{Drag Force}}{\text{unit volume}} \quad \theta\text{-direction} \quad (3)$$

Here U is the tangential velocity, P is the pressure, and ρ_g is the gas density.

To further simplify the model, a solid body rotation in the bed is considered and the coriolis force is ignored (an assumption that may be verified later).

Assuming that the drag force exerted on the unit volume of gas by viscous transfer in the bed has the form of Ergun correlation (or similar available models such as the Carman-Kozeny or Morton model), equations (2) and (3) reduce to:

$$-\rho_g \frac{\bar{V}}{\epsilon^2} \frac{d\bar{V}}{dr} = \frac{dP}{dr} + \rho_g \omega^2 r + \rho_g A \bar{V} + \rho_g B \bar{V}^2 \quad (4)$$

Here $\bar{V}$ is the local superficial gas velocity, related to interstitial gas velocity by $\bar{V} = V\epsilon$. ω is the rotational speed, and A and B are Ergun correlation coefficients which depend on gas physical properties and the packing specifications and defined as follows:

For Ergun's model: $A=150(1-\epsilon)\mu_g / \epsilon^3(\varphi_s d_p)^2$, $B=1.75(1-\epsilon)\rho_g / \epsilon^2\varphi_s d_p$

For Morton's model: $A=8.5\mu_g a_p / \epsilon^3$, $B=a_p \rho_g (Re)^{-1} / \epsilon^3$

Here a_p is the specific surface area of the packing, d_p is the packing diameter, φ_s is the sphericity of the packing particles, and Re is the gas Reynold's number based on the specific surface area. In general, these models are written in terms of the slip velocity of the two phases. However, since the superficial solid velocity is zero, only the gas velocity appears in the equation (we will see later that this is not the case when the bed is irrigated by liquid).

Equation (4) may be integrated across the entire bed to obtain the pressure drop inside of the rotor, ΔP_i , for a dry bed.

4.1.1.b. Pressure Drop Between the Housing and Rotor:

Outside of the rotor, we have a flow situation similar to flow between two disks, one spinning and one stationary, with the inflow. There have been numerous studies of this type of flow in the literature [17–21]. The basic flow configuration is as follows: the rotating disk imparts rotation to the fluid, leading to a distribution of angular velocity from the rotating disk value at the disk to zero at the stationary housing wall. The drag forces act on the fluid in the direction of rotation at the rotating disk and in the opposite direction at the wall surface. The pressure gradient is generated centrifugally, in the gap between the disk and the wall, and its value depends on the effective average angular velocity of the flow

field. The pressure tends to be a function of radius only and independent of axial position in the gap. The angular velocity of the fluid near the rotating disk is above the effective average, and hence the fluid near the rotating disk is thrown radially outward. On the other hand, fluid close to the stationary housing has smaller angular velocity than the effective average so that it flows radially inward. The effect of inflow due to leakage (in the rotating packed bed the leakage is the total gas flow rate) is to increase the radial pressure gradient. This increase in pressure gradient can be explained in terms of the rotation effect of inflow. The fluid tends to rotate faster due to inflow and so generates larger pressure gradient.

A momentum integral method is used to analyze the pressure gradient between the rotating disk and the stationary housing. The governing equations are the same as equations (1) to (3). The difference is that the simplifications made in evaluating ΔP_i , ignoring the coriolis force and the assumption of solid body rotation, are not valid here. A distribution of tangential velocity exists which can be obtained by solving equation (3). The drag forces are the radial and tangential friction stresses on the wall and the rotating disk. These forces are deduced from work performed by Von. F. Shultz-Grunow [20] and are defined as follows:

$$\tau_{rw} = \text{stress in radial direction at wall} = f_r \rho_g \frac{U^2}{2}$$

$$\tau_{rd} = \text{stress in radial direction at rotating disk} = f_r \rho_g \frac{(r\omega - U)^2}{2}$$

$$\tau_{tw} = \text{stress in tangential direction at wall} = f_t \rho_g \frac{U^2}{2}$$

$$\tau_{td} = \text{stress in tangential direction at rotating disk} = f_t \rho_g \frac{(r\omega - U)^2}{2}$$

where f_r and f_t are the friction coefficients that take the following expressions for the turbulent flow:

$$f_r = \frac{0.0391}{(\rho_g \omega r_o^2 / \mu_g)^{0.2}} \times \frac{1}{(1 - r/r_o)^{0.4}} \quad (5)$$

$$f_t = \frac{0.0996}{(\rho_g \omega r_o^2 / \mu_g)^{0.2}} \times \frac{1}{(1 - r/r_o)^{0.4}} \quad (6)$$

Here μ_g is the gas viscosity, and r_o is the disk outside radius. After substituting for f_r and f_t from equations (5) and (6), equations (2) and (3) can be solved simultaneously to evaluate the pressure gradient in the gap between the rotating disk and the stationary wall, ΔP_o . The equations were solved numerically and a 4th order Runge-Kutta method was employed to approximate the integrations.

4.1.2. Irrigated Packing:

4.1.2.a Pressure Drop Across the Bed:

To extend the proposed model for dry packing to the irrigated packing equation (4) will be modified to account for the liquid presence. The velocity terms related to Ergun formula will be written in terms of the slip velocity of the gas and liquid phases:

$$\frac{dP}{dr} = A (\bar{V} - V_i) + B (\bar{V} - V_i)^2 + \rho_g \omega^2 r + \rho_g \frac{d\bar{V}}{dr} \quad (7)$$

Here V_i is the interfacial velocity.

The equation of motion for the liquid steam, assuming laminar and fully developed flow within the short distance that both phases are brought into contact, may be written as follows:

$$\mu_1 \frac{d^2 V_1}{dz^2} = -\frac{dP}{dr} + \rho_1 \omega^2 r \quad (8)$$

Here V_1 is the liquid phase velocity, μ_1 is the liquid viscosity, ρ_1 is the liquid density, and z is the distance in direction perpendicular to the radial direction. The above equation can be solved based on the boundary conditions:

$$\begin{aligned} V_1 &= V_i \quad @ \quad z = \delta \\ V_1 &= 0 \quad @ \quad z = 0 \end{aligned}$$

δ is the liquid film thickness. The solution, after integrating over the liquid film thickness, may be written in terms of the liquid flowrate as follows:

$$Q = \left(\rho_1 \omega^2 r - \frac{dP}{dr} \right) \frac{h \delta^3}{12 \mu_1} + \frac{h V_i \delta}{2} \quad (9)$$

where Q is the liquid flowrate, and h is the axial height of the packing.

Equations (7) and (9) are coupled through dP/dr , V_i , and δ terms. Therefore, one more equation is needed to solve for the above unknowns and that arises from the boundary condition indicating equal stresses at the gas and liquid interface:

$$\tau_g = \tau_l \quad \text{at } z = \delta$$

Here τ_g and τ_l are the gas and liquid shear stresses, respectively.

Solving the equations by using the above boundary condition is excessively involved and complicated. Instead, we try to estimate the liquid film thickness by using either the asymptotic solution for liquid film on a rotating disk [13] or liquid film flow on an inclined surface [22] after substituting centrifugal acceleration for the gravity (it is shown later that the film thickness can be estimated for a rotating bed by carrying out the Residence Time Distribution (RTD) experiments).

For given film thickness, gas and liquid flowrates, speed of the rotation, and other liquid and gas physical properties, equations (7) and (9) can be solved to evaluate the pressure drop across the rotating bed. A 4th order Runge Kutta method was employed to solve the equations numerically.

4.1.2.b. Pressure Drop Between the Housing and Rotor:

The above analysis for the irrigated packing gives the pressure drop across the bed. It is assumed that the pressure drop between the rotating disk and the stationary housing is not affected by the presence of the liquid; therefore, the same analysis as dry packing is applied for the irrigated packing for the outside of the rotor.

4.2. EXPERIMENTAL RESULTS & DISCUSSIONS:

4.2.1. Typical Pressure Drop Observation:

A peculiar phenomenon is observed in pressure drop measurements of rotating beds. Figure 3 shows a typical plot of pressure drop against time. When

Figure 3. Pressure Drop Across Bed

the rotor is stationary and the gas is turned on, the pressure drop is 11 Pa/cm of bed. When the rotor is turned on, the pressure drop increases to 50 Pa/cm of bed. When the liquid is turned on, countercurrent to the gas flow, the pressure drop decreases by more than a factor of 2, to 22 Pa/cm of bed. For this experiment, a gas flowrate of 1.0 kg/sec-m² and a liquid flowrate of 36 kg/sec-m² were used. Packing surface area was 1476 m²/m³, and the packed bed was rotating at 600 rpm.

To understand this phenomenon, various experiments were performed to isolate the influence of different operating parameters.

4.2.2. Gravitational Effects (Constant Liquid and Gas Flow):

Experiments were carried out to measure the pressure drop at various acceleration fields (varying speed of rotation). The liquid and gas flowrates were held constant at 36 kg/sec m² and 1 kg/sec m² respectively. The packing with the specific surface area of 1476 m²/m³ was used. The pressure drops were measured for the following situations: i) zero flow (gas off-liquid off), ii) liquid off-gas on, iii) liquid on-gas off, and iv) liquid on-gas on. Representative data are presented in Table 2. and are plotted in Figure 4.

At zero flow a radial pressure gradient develops because of the rotation of the air both inside the rotor and between the rotor and the stationary housing. This is associated with the pressure required to overcome the centrifugal head set due to the rotation. The pressure drop increases as the gas flow rate is turned on (gas on-liquid off). This increase results mainly from the drag (frictional pressure drop) and inertia effects. The measured pressure drops for the case of liquid-on

TABLE 2. RESULTS OF PRESSURE DROP AT VARYING ROTATIONAL SPEED

State of Experimental Conditions	Speed of Rotation (rpm)	Pressure Drop (Pa/cm)
LIQUID OFF GAS OFF	400	10.3
	600	17.0
	679	22.1
	797	29.4
	853	34.8
	944	42.3
	1025	49.2
	1182	64.6
	1253	72.0
LIQUID OFF GAS ON (1.0 kg/sec m ²)	400	50.9
	600	69.5
	681	86.2
	790	109.9
	855	127.3
	947	147.4
	1047	174.1
	1136	189.3
	1203	221.1
LIQUID ON (36 kg/sec m ²) GAS OFF	400	9.9
	600	16.7
	652	17.4
	740	19.6
	830	26.6
	937	35.8
	1039	45.8
	1159	54.1
	1269	65.5
LIQUID ON (36 kg/sec m ²) GAS ON (1.0 kg/sec m ²)	400	19.5
	600	30.0
	700	39.9
	840	51.0
	950	74.5
	1200	124.0

Figure 4. Pressure Drop at Varying

gas-off are almost the same as those of the liquid-off gas-off case, indicating that liquid has no effect on pressure drop for gas-off situations. However, a decrease in the pressure drop is observed in the case of liquid-on gas-on as compared with liquid-off gas-on case. This is not a typical trend observed in conventional packed towers. In conventional towers the irrigated bed shows a higher pressure drop than the dry bed.

The solid lines in Figure 4 indicate that the dependency of the pressure drop on rotational speed is $\Delta P \propto \omega^2$ at constant liquid and gas flowrates, as suggested in the proposed model.

4.2.3. Gas Flowrate Effects:

Pressure drop vs. gas flowrates curves were obtained at constant liquid flowrate and speed of rotation. Gas flowrates ranged from 0.1 kg/sec-m² to 1.95 kg/sec-m². Corresponding semi-log representation of the results (as suggested by Teutsch [23]) has two advantages: i) lines for constant pressure and liquid flowrate are practically straight, ii) flooding points can be clearly identified. Figure 5 shows a typical result for 12 kg/sec-m² and 36 kg/sec-m² liquid flowrates at 600 rpm speed of rotation with 1476 m²/m³ packing specific surface area. However, the flooding points, the breakpoint in the plots where the pressure drop increases rapidly, was not reached. Liquid entrainment in the region of the distributor was encountered at the points indicated. The fact that the distributing tubes of the distributor are not placed as close as possible to the inner surface of the packing may cause the excessive entrainment of the liquid before the flooding (the liquid is

Figure 5. Pressure Drop at Varying Gas Flowrate

not accelerated since it does not reach the bed). Note that the change in pressure drop is not significant as the liquid flowrate is tripled. To confirm that the same pressure drop characteristics are observed at different acceleration fields, a liquid flowrate was selected (36 kg/sec-m^2), and the total pressure drop was measured at varying gas flowrates for the rotational speeds of 600, 840, 1200 rpm to give a gravitational acceleration of 50, 100, and 200g, respectively (based on inner radius of the rotor). Figure 6 shows the results. Obviously, a higher pressure drop is observed, at constant liquid and gas flowrates, for higher speed of rotation (the centrifugal force contribution to the total pressure drop). It is very clear that the curve corresponding to 1200 rpm is not even near flooding (no entrainment was observed at full capacity liquid and gas flowrates).

4.2.4. Liquid Flowrate Effects:

Pressure drop versus liquid flowrates curves were obtained at constant rotational speed and gas flowrate. The results are shown in Figure 7. The pressure drop is not influenced as much by liquid flowrates as it is by gas flowrates.

4.2.5. Effect of Packing Specific Surface Area:

To investigate the effects of packing surface area, tests were made with different surface area packings. The packings have a specific surface area of 656, 1476, and $2952 \text{ m}^2/\text{m}^3$, all with 0.92 void fraction. The pressure drops were measured for dry and wet beds at rotational speeds of 600 and 900 rpm. The results are shown in Figure 8. The higher pressure drops are associated with the higher surface area packing for the dry and wet beds. Note also that the dry bed shows higher pressure drop than the wet bed.

Figure 6. Pressure Drop at Varying Gas Flowrate

Figure 7. Pressure Drop at Varying Liquid Flowrate

Figure 8. Pressure Drop vs Packing Specific Surface Areas

4.3. Comparison with Model:

4.3.1. Zero Flow:

With no gas and liquid flowing, the pressure drops were measured at varying rotational speeds. The experimental results for dry bed (gas flowrate of 1.0 kg/sec-m², and packing surface area of 1476 m²/m³) are compared with the values estimated by the proposed model. Figure 9 shows this comparison.

4.3.2. Dry Bed:

The experimentally obtained values of the pressure drop, for dry packing, are compared with the values estimated by the proposed model in Figure 10. All of the predictions are within $\pm 20\%$ error shown by the two broken lines in Figure 10. The experimental data was taken as the gas flowrates were varied from 0.163 to 1.31 kg/sec-m², while the speed of rotation was ranging from 600 to 1200 rpm. Packing surface areas of 1476 and 2952 m²/m³ were used. Table 3 presents these results.

4.3.3. Irrigated Bed:

The experimentally obtained values of the total pressure drop, for irrigated packing, are compared with the values estimated by the model proposed for the wet packing in Figure 11. The operating conditions were similar to those of the dry packing experiments; however, the liquid flowrates were varied from 9 to 36 kg/sec-m². The agreement between the calculated values and the experimental data is satisfactory (within $\pm 20\%$ error). Although the agreement between the experimental data and the proposed model for the wet packing is promising, but

**Figure 9. Measured and Calculated
Pressure Drop of No Flow**

TABLE 3. COMPARISON OF MEASURED AND CALCULATED PRESSURE DROPS
AT VARYING EXPERIMENTAL CONDITIONS FOR THE DRY BED

a_p (m^2/m^3)	Speed of Rotation (rpm)	Gas Flow (kg/s- m^2)	$\Delta P_{\text{measured}}$ (Pa/cm)	$\Delta P_{\text{calculated}}$ (Pa/cm)
2952	600	1.0	68.4	73.2
2952	695	1.0	81.9	79.0
2952	766	1.0	91.2	81.9
2952	840	1.0	111.8	86.7
2952	917	1.0	127.2	95.4
2952	600	0.33	27.0	24.1
2952	600	0.48	32.8	32.8
2952	600	0.52	34.7	33.7
2952	600	0.58	45.3	40.5
2952	600	0.69	48.2	47.2
2952	600	0.86	49.2	55.9
2952	600	0.97	55.9	60.7
2952	600	1.03	61.7	73.3
1476	600	1.03	55.9	55.9
1476	681	1.03	65.5	59.8
1476	790	1.03	82.9	66.5
1476	855	1.03	96.4	68.4
1476	947	1.03	111.8	73.3
1476	1040	1.03	132.1	80.1
1476	1140	1.03	151.3	86.0
1476	1200	1.03	167.7	95.0
656	600	0.69	21.2	19.3
656	600	0.86	23.1	21.2
656	600	1.03	29.9	28.9

**Figure 10. Measured and Calculated
Pressure Drop of Dry Bed**

**Figure 11. Measured and Calculated
Pressure Drop of Wet Bed**

the model still does not explain the lower pressure drop for the wet bed (compared to the dry bed). One possible speculation is that the liquid film exposed to the centrifugal field gets thin and behaves like a lubricant film, lowering the drag against the gas phase. Another plausible explanation is a shorter path length for the gas in the wet bed compared to the dry bed. The gas may possess smaller rotational speed in the presence of the liquid, as it enters and leaves the rotor, and as a result a lower pressure drop is observed in the wet bed.

The experimental results of pressure drop indicate that the characteristics of a rotating bed are different from those of a conventional packed bed. Therefore, the conventional analysis and known hydrodynamic relations of packed towers may not be easily applied to rotating beds.

4.3.4. Comparison with Conventional Packed Beds:

To evaluate further the pressure drop characteristics of rotating beds, the generalized isobaric plot of pressure drop correlation, the modified Sherwood's flooding correlation, was investigated by comparing the measured values of the pressure drop with the values obtained from the correlation (the plot) at different gas and liquid flowrates, and rotational speeds (g , the gravitational field was replaced by $\omega^2 r$, the rotational acceleration in the correlation). The rough comparison suggests that the measured values are orders of magnitude higher than the corresponding values from the curves. Figure 12 shows how these magnitudes vary at different rotational speeds. It is important to keep in mind that the values of the pressure drop in the rotating bed resulted from a variety of sources (in

**Figure 12. Comparison with Conventional
Pressure Drop Correlation**

conventional columns the pressure drop is only a combination of drag forces). The air-water system was used for these experiments, and liquid flowrates ranged from 8.93 to 36 kg/sec-m², and the vapor flowrate varied from 0.1 to 1.22 kg/sec-m². The rotational speed varied from 600 to 1200 rpm to achieve a range of acceleration fields of 50 to 200g. The packing specific surface area was 1476 m²/m³ with 0.92 porosity.

5. MASS TRANSFER CHARACTERISTICS OF ROTATING BEDS

5.1 Theory:

The total height of a packed column is given by:

$$H = (NTU) \times (HTU) \quad (10)$$

where

H = tower height

NTU = number of transfer units

HTU = height of a transfer unit

The derivation of the above equation rests on simple material balance: the rate of change of a component within a phase must be equal to the rate of transfer to the phase:

$$d(\bar{L}a) = k_l(a - a_e) dA \quad (11)$$

Here, $\bar{L}$ is the liquid molar flowrate, a is the liquid concentration, a_e is the liquid concentration at equilibrium with the gas phase, k_l is the liquid side mass transfer coefficient, and dA is the interfacial transfer area, which may be expressed as:

$$dA = a_e S dH \quad (12)$$

where a_e is the interfacial area per unit volume of packing, S is the empty tower cross-sectional area, and dH is the differential tower height. Equation (11) may

be used to solve for the tower height required by integrating over the total change in concentration between the tower terminals:

$$\int_0^H dH = \int_{\alpha_0}^{\alpha_1} \frac{d(\bar{L} \alpha)}{k_1 a_e S (\alpha - \alpha_e)} \quad (13)$$

Simplifications may be made by assuming that the phase flowrate, $\bar{L}$, is constant. This is a reasonable assumption for dilute systems. Since the volumetric mass transfer coefficient is a function of the this flowrate, the coefficient must be constant, providing physical properties do not vary appreciably. Thus,

$$\int_0^H dH = \frac{\bar{L}}{k_1 a_e} \int_{\alpha_0}^{\alpha_1} \frac{d\alpha}{(\alpha - \alpha_e)} \quad (14)$$

The integral term is equal to the total change of composition divided by the available driving force. This is defined as number of transfer units, NTU. The quantity outside the integral is called the height of a transfer unit, HTU. The tower height, H , is the product of these two quantities as expressed in equation (10).

The above mathematical tools may be used to express the measured experimental results (for Air-CO₂-Water system) in terms of height of a transfer unit for rotating packed beds, as follows:

$$NTU = \int_{\alpha_0}^{\alpha_1} \frac{d\alpha}{\alpha - \alpha_e} \quad (15)$$

where,

α_1 = CO₂ concentration in the rotor inlet

α_0 = CO₂ concentration in the rotor outlet

α_e = CO₂ concentration in the liquid that would be at equilibrium
with CO₂ concentration at the bulk gas phase

A material balance on carbon dioxide gives:

$$\bar{L} (\alpha - \alpha_0) = \bar{G} (y - y_0) \quad (16)$$

Here $\bar{L}$ and $\bar{G}$ are the molar water and air flowrates respectively, and y_0 is the concentration of CO₂ in the rotor outlet. Assuming Henry's law to be valid for the dilute system:

$$y = \dot{H} \alpha \quad (17)$$

where $\dot{H}$ is the Henry's law constant. Using equations (16) and (17) to solve for α_e :

$$\alpha_e = \frac{1}{\dot{H}} \left\{ \frac{\bar{L}}{\bar{G}} (\alpha - \alpha_0) + y_0 \right\} \quad (18)$$

Substituting for α_e from equation (18) in to equation (15), then equation (15) may be integrated to give:

$$NTU = \frac{\alpha_i - \alpha_o}{(\alpha - \alpha_e)_i - (\alpha - \alpha_e)_o} \ln \frac{(\alpha - \alpha_e)_i}{(\alpha - \alpha_e)_o} \quad (19)$$

and finally to use the following expression to evaluate HTU:

$$HTU = (r_o - r_i) / NTU \quad (20)$$

Here, r_o , and r_i are the outside and inside radii of the rotating bed.

The number of transfer units is a function of the concentration of various components in the entering and exiting streams, and the phase equilibria properties of the components. The height of a transfer unit depends on liquid and gas flowrates, the nature of the packing, and the liquid and gas phase volumetric mass transfer coefficients. The number of transfer units for a given system is not changed by carrying out the separation in high gravitational fields, but the height of a transfer unit is drastically reduced because of its dependency on volumetric mass transfer coefficients, $k_1 a_e$, k_1 is the liquid side mass transfer coefficients and a_e is the effective gas-liquid interfacial area. The volumetric mass transfer coefficients increase because of the use of high surface area packing and the increase in the liquid side mass transfer coefficient. Dudukovic, et al showed [14] that the other contributing factor to enhance the volumetric mass transfer

coefficients in a rotating packed bed is an increase in the effective gas-liquid contacting efficiency, a_e/a_p , where a_p is the specific surface area of the packing.

Among theories proposed to describe the liquid mass transfer behavior in a packed column under conventional operation, the penetration theory appears most consistent with the picture existing in a packed tower. The penetration theory suggests that as the liquid flows over each piece of packing in laminar flow, it is completely mixed in passing from one packing piece to the next. Absorption is considered to take place during a series of brief contacts between the liquid and gas in which the dissolved gas diffuses only a short distance into the liquid film. The governing equation is:

$$V_i \frac{\partial C}{\partial r} = \mathcal{D} \frac{\partial^2 C}{\partial x^2} \quad (21)$$

and the boundary conditions are:

$$C = 0 \quad @ \quad r = 0 \quad (22a)$$

$$C = 0 \quad @ \quad x = \infty \quad (22b)$$

$$C = C_o \quad @ \quad x = 0 \quad (22c)$$

Here, C is the concentration of the diffusing component, V_i is the interfacial velocity of the liquid film flowing on the wall, $\mathcal{D}$ is the diffusion coefficient, C_o is the interfacial concentration, r is the radial direction, and x is the axial direction (perpendicular to r direction)..

Solving equation (21) under boundary conditions of (22) will result in the following expression for k_1 , the liquid side mass transfer coefficient:

$$k_1 = \sqrt{4 \mathcal{D} / \pi t} \quad (23)$$

where t is the time during which the surface of the liquid is exposed to the gas and can be estimated by the following equation:

$$t = \pi Z / 2 V_i \quad (24)$$

where Z is the packing length characteristic. The interfacial velocity, V_i , may be estimated by stream-line flow theory as

$$V_i = (9g\Gamma^2 / 8\mu_1\rho)^{1/3} \quad (25)$$

where Γ is the liquid flowrate per unit width, μ_1 is the liquid viscosity, and ρ is the liquid density. Equation (25) is obtained under assumptions of negligible interfacial shear and the complete mixing of the liquid at the junctions of packing pieces.

Substituting equation (25) into equation (24), and equation (24) into equation (23), results in the following expression for liquid side mass transfer coefficient:

$$\frac{k_1 Z}{\mathcal{D}} = 0.918 (a_p/a_e)^{1/3} Sc^{1/2} Re^{1/3} Gr^{1/6} \quad (26)$$

where,

$$Sc = \text{Schmidt number} = \mu_1 / \rho \mathcal{D}$$

$$Re = \text{Reynolds number} = L / a_p \mu_1$$

$$Gr = \text{Grashof number} = Z^3 \rho^2 g / \mu_1^2$$

Here, L is the superficial liquid mass flowrate.

The liquid side mass transfer coefficient, k_1 , may be evaluated using equation (26), after the effective gas-liquid contacting efficiency, a_e/a_p , is estimated. In estimating the effective gas-liquid contacting efficiency, a_e/a_p , it is important to keep in mind that the gas-liquid interfacial surface area is influenced by the body force. The fact that the interfacial area is sensitive to liquid flowrates, as most of the correlations in the literature predict, suggests it is likely that the interfacial area would be dependent on the body force. Dudukovic et al, showed that the effective surface area increases with the body force with 0.14 to 0.28 power [14]. Therefore, among the suggested correlations the one that takes into account the influence of the body force on the effective interfacial area should be used.

Among several formulae proposed to correlate interfacial areas, the most general one seems to be the equation by Onda et al [24] that predicts the wetted surface area and not only takes into account surface tension and the surface energy of the packing, but also has a gravity term built in to the correlation:

$$a_e/a_p = a_w/a_p = 1 - \exp\{-1.45(\sigma/\bar{\sigma})^{0.75} Re^{0.1} Fr^{-0.05} We^{0.2}\} \quad (27)$$

where,

a_w = wetted surface area of the packing

$\bar{\sigma}$ = surface tension of liquid

σ_c = critical surface tension of packing material

Fr = Froude number = $L^2 a_p / \rho^2 g$

We = Weber number = $L^2 / \rho \bar{\sigma} a_p$

Equation (17) combined with equation (18) may be used to evaluate liquid side volumetric mass transfer coefficient in rotating packed bed by replacing gravitational acceleration, g , with rotational acceleration, $a_r = \omega^2 r_i$, and liquid flowrate per unit width with liquid flowrate per perimeter. The major assumption is that the interfacial surface area of the gas-liquid, a_e , is the same as the packing wetted area, a_w . The reported experimental values of a_e by Dudukovic et al [14] may be estimated, within 7% to 10%, using this correlation.

Then, the following equation will be used to estimate the value of height of a transfer unit, HTU:

$$HTU = L / \rho k_l a_e \quad (28)$$

Another conventional analysis that may be applied to rotating beds is the theoretical work made by Davidson [25]. Davidson was the first to consider a

theoretical approach to evaluate liquid side mass transfer coefficient by developing three statistical models to represent random packings. He used the penetration theory to make quantitative calculations for packed towers. In the vertical surface model, the packing was assumed to be made up of a large number of vertical surfaces of height Z which were completely and equally wetted. The model provides an expression for HTU:

$$\text{HTU} / Z = 0.345 \text{ Sc}^{1/3} \text{ Re}_1^{2/3} \text{ Gr}^{-1/6} \quad (29)$$

Here, $\text{Re}_1 = 4 L / \mu a_e$. He then extended this model to cases of random angle, and random angle and length, and he developed similar equations for all the cases, with different constants. This model may be used for rotating beds by replacing the appropriate terms as was suggested for the earlier model. Note that to evaluate Re_1 , a_e , the effective gas-liquid interfacial area should be estimated. Equation (27) may be used to estimate a_e .

Both of these models give a very simple picture of what could be happening in a rotating bed; complications such as actual flow pattern of the fluids, Coriolis force, fluid distribution, and packing shape are all neglected. On the other hand, this approach may provide some tools to reveal some of the significant differences between the rotating beds and conventional packed towers.

5.2. EXPERIMENTAL RESULTS AND DISCUSSION:

- CO_2 in water

Experiments are performed to investigate the effects of operating parameters, particularly the acceleration field, the packing specific surface area,

and the liquid and gas loading, on HTU. About 75 experimental runs were made at varying operating conditions. The results of the mean values of HTU with the standard error of the mean are summarized in Table 4. The results of all the experiments are presented in the Appendix.

5.2.1. Gravitational Effects:

Experimental runs were made as the liquid and gas flowrates were held constant at 36 and 1.0 kg/sec-m², while the rotor speed of rotation was set at 600, 850, 1130, and 1385 rpm to achieve a centrifugal acceleration of 50, 100, 182, and 272g, respectively (based on the inner radius of the rotor). All three beds were used. The influence of the acceleration field on HTU is shown in Figure 13 which indicates the dependency of HTU on g varies from -0.3 to -0.4, within the range estimated by Dudukovic et al [14,15] and Vivian et al [6]. Comparing the slopes of the lines for different packing surface areas suggests that the dependency of HTU on the acceleration field is more pronounced at higher surface area packings.

5.2..2. Effect of Packing Specific Surface Area:

The contribution of packing specific surface area on enhanced volumetric mass transfer coefficients is shown in Figure 14. The liquid and gas flowrates for all the experimental runs were set at 36 kg/sec-m² and 1.0 kg/sec-m², respectively, while the rotational speeds were set at 600, 850, 1130, and 1385 rpm to achieve a centrifugal acceleration of 50, 100, 182, 272g, respectively (based on the inner radius of the rotor). The measured volumetric mass transfer coefficients of 2952 m²/m³, in all different accelerations, are greater than the corresponding

TABLE 4. EXPERIMENTAL DATA AT VARYING OPERATING CONDITIONS

Liquid Flow (kg/sec-m ²)	Gas Flow (kg/sec-m ²)	# g	a _p (m ² /m ³)	HTU _{ave} (cm)	Standard Error of Mean
18	1.0	50	1476	3.0	0.052
36	1.0	50	1476	3.75	0.020
36	1.0	50	2952	2.42	0.076
36	1.0	50	656	2.74	0.019
36	0.66	50	1476	2.63	0.088
18	1.0	100	1476	2.24	0.051
36	1.0	100	1476	2.94	0.177
36	1.0	100	2952	1.97	0.038
36	1.0	100	656	2.20	0.000
36	0.66	100	1476	2.10	0.052
18	1.0	182	1476	1.97	0.012
36	1.0	182	1476	2.45	0.132
36	1.0	182	2952	1.67	0.014
36	1.0	182	656	2.10	0.050
36	0.66	182	1476	1.92	0.020
18	1.0	272	1476	1.79	0.003
36	1.0	272	1476	2.0	0.000
36	1.0	272	2952	1.30	0.095
36	1.0	272	656	1.92	0.085
36	0.66	272	1476	1.62	0.030

Figure 13. HTU at Varying Accelerations

Figure 14. Variation of Volumetric Mass Transfer Coefficient with Packing Area

packing will result in higher volumetric mass transfer coefficients. However, as can be seen in Figure 14, the results for 656 m²/m³ packing do not show the same trend. The volumetric mass transfer coefficients in the case of 656 m²/m³ packing are slightly greater than those of 1476 m²/m³ (or comparable within the experimental errors). This discrepancy may be attributed partly to end effects: the fact that the mass transfer cannot be confined to the packing alone. Some mass transfer may take place as the fine droplets leave the rotor and travel to the housing. Another factor which may contribute to the higher experimental volumetric mass transfer coefficients for 656 m²/m³ (compared to those of 1476 m²/m³) is that the liquid spread in the tangential direction due to the Coriolis force is more pronounced for 656 m²/m³ packing with larger pore size (allowing larger deviation from the radial direction). This means that the wetted surface area, a_w , may be higher for 656 m²/m³ packing than that of 1476 m²/m³ packing, resulting in higher volumetric mass transfer coefficients. More work needs to be done to reinforce the above statements.

5.2.3. Effect of Liquid Flowrate:

To investigate the liquid flowrate effects, the HTU dependency on the acceleration fields was determined at liquid flowrates of 36 kg/sec-m² and 18 kg/sec-m². The gas flowrate was held at 1.0 kg/sec-m², and the packing with specific surface area of 1476 m²/m³ was used. The results are presented in Figure 15. The slope of the best line drawn through these data, using least square fit technique, is -0.33 for higher liquid flowrate in comparison with the slope of -0.30 at the lower liquid flowrates. Note that higher values of HTU were obtained at higher liquid flowrate, something to be expected.

**Figure 15. Effect of Liquid Flowrate
on HTU at High Accelerations**

5.2.4. Effect of Gas Flowrate:

Tests were also performed at gas flowrate of 0.66 kg/sec-m^2 at varying centrifugal accelerations, while the liquid flowrate kept constant at 36 kg/sec-m^2 . The results of these tests are presented in Figure 16. Although lower values of volumetric mass transfer coefficients were measured for these experiments in comparison with the higher flowrate (1.0 kg/sec-m^2), the power dependency of HTU on gravitational field turned out to be the same as before (slope of the line is -0.30). It is important to point out that the fact that HTU increases as the gas flowrate decreases is consistent with the trend observed in conventional columns.

5.2.5. Effect of Viscosity:

Increasing the liquid viscosity decreases mass transfer rate due to reduced diffusivity and mixing in the liquid phase. To investigate the effect of viscosity on mass transfer in rotating beds, a solvent-hexane-air system was studied, where the hexane was air stripped from the solvent at 70°C and 1 atm. The solvent viscosity was 3 cp at 70°C .

At rotational speeds of 700 to 1100 rpm, HTU ranges from 1.5–2.5 cm were achieved. The results are presented in Table 5. The packing with specific surface area of $1476 \text{ m}^2/\text{m}^3$ was used.

5.3. Comparison with Models:

The experimentally obtained values of HTU from all the performed experiments at varying gravitational fields, as well as different packing surface areas, and varying gas and liquid flowrates, are compared with the HTU values estimated from both models, using equations (26) to (29). The results are presented in Table 6. This comparison is illustrated in Figure 17. Figure 17 shows

**Figure 16. Influence of Gas Flowrate
HTU at High Accelerations**

TABLE 5. EXPERIMENTAL RESULTS OF HIGH VISCOSITY SOLVENT

T_{liquid} (° C)	T_{gas} (° C)	Liquid Flow (Kg/s-m ²)	Gas Flow (Kg/s-m ²)	#g _{ave.}	# stages
79.3	92.8	28	1.10	75	4.15
84.2	88.9	26	1.20	75	4.39
87.0	93.9	26	1.12	75	4.13
85.7	94.5	29	1.11	75	1.80
84.7	88.9	26	1.11	165	6.18
82.1	93.9	28	1.10	184	7.90
86.2	94.4	29	1.12	208	10.14
86.9	94.4	29	1.11	343	12.60
90.5	94.4	29	1.11	447	14.0
87.8	94.4	28	1.11	474	13.2
85.2	93.9	28	1.12	388	6.50
89.0	88.9	27	1.12	422	7.50
93.0	88.9	27	1.12	539	12.50
86.4	93.9	26	1.12	150	5.63

TABLE 6. COMPARISON OF MEASURED AND CALCULATED VALUES OF HTU

#g	Measured HTU (cm)	Calculated HTU (cm) Using Penetration Theory	Calculated HTU (cm) Davidson's Model
LIQUID FLOW: 36 Kg/s-m ² , GAS FLOW: 1.0 Kg/s-m ² , a _p : 1476 m ² /m ³			
50	3.67	3.29	3.60
100	2.80	2.92	3.10
182	2.30	2.50	2.80
272	2.00	2.40	2.60
LIQUID FLOW: 18 Kg/s-m ² , GAS FLOW: 1.0 Kg/s-m ² , a _p : 1476 m ² /m ³			
50	2.98	2.50	2.70
100	2.25	2.19	2.39
182	1.98	1.92	2.10
272	1.79	1.83	2.00
LIQUID FLOW: 36 Kg/s-m ² , GAS FLOW: 1.0 Kg/s-m ² , a _p : 2952 m ² /m ³			
50	2.22	1.58	1.71
100	1.83	1.41	1.51
182	1.67	1.22	1.33
272	1.22	1.16	1.26
LIQUID FLOW: 36 Kg/s-m ² , GAS FLOW: 1.0 Kg/s-m ² , a _p : 656 m ² /m ³			
50	2.80	7.01	5.60
100	2.19	6.09	5.00
182	2.01	5.48	4.40
272	1.96	5.18	4.2

Figure 17. Comparison of Measured and Calculated Values of HTU

the results are within ± 30 , except for the results of $656 \text{ m}^2/\text{m}^3$ packing. It is interesting to observe that the estimated values agree fairly well with the experimental values. Neither model seems to show any superiority over the other within these range of experimental conditions. However, Davidson's model seems to underpredict the HTU values. The good agreement between the estimated and experimental values is not observed for the values corresponding to $656 \text{ m}^2/\text{m}^3$ packing because of possible reasons discussed earlier.

6. POWER REQUIREMENT FOR ROTATING BEDS

6.1. THEORY:

The power consumption arises from three sources: 1) the power required for the acceleration of the liquid to the tip speed of the rotor; 2) the power consumed in overcoming the frictional windage drag of the vapor; and 3) the power required for the bearing friction. In addition, there is a small power requirement associated with the flow of vapor between the rotating disk and the stationary housing, and the flow of vapor through the rotor bed.

6.1.1. Bearing Friction:

The power consumed by the bearing friction will be strongly dependent upon the type of the bearing used. It is reasonable to assume that the frictional loss of bearing is directly proportional to speed of rotation. The proportionality may be evaluated experimentally.

6.1.2. Windage Effect:

Windage losses may be evaluated by studying the flow between two disks, one spinning and one stationary. There have been numerous studies of this type of flow in the literature [17–21]. In particular, the flow pattern of disk in a housing with large gap between them (width of the gap compared with the radius of the disk) was investigated for the laminar and for the turbulent flow by Von F. Shultz–Grunow [20]. It was suggested that for a large gap, an additional boundary layer will be formed on the stationary housing, as well as on the rotating disk. The

fluid in the boundary layer on the revolving disk is centrifuged outwards and this is compensated by a flow inwards in the boundary layer on the stationary housing. For the turbulent flow case, the tangential velocity was assumed to obey the 1/7-th-power law. The expressions for the torque both for the laminar and turbulent cases are as follows:

$$T = 2.67 \rho_g \omega^2 R^5 (\text{Re})^{-1/2} \quad \text{Laminar} \quad (30)$$

$$T = 0.0622 \rho_g \omega^2 R^5 (\text{Re})^{-1/5} \quad \text{Turbulent} \quad (31)$$

Here, T is the torque, ρ_g is the gas density, R is the disk radius, and Re is Reynold's number defined as:

$$\text{Re} = \rho_g \omega R^2 / \mu_g$$

where μ_g is the gas viscosity.

The power is, of course, the product of the torque and angular velocity:

$$\text{Power} = T \times \omega \quad (31)$$

6.1.3. Windage Effect with Inflow:

The flow pattern for a rotating disk in a stationary housing is modified when there is inflow [17, 19]. There is a significant influence of inflow on shear stresses

at the wall and the disk [17]. The fluid rotates faster due to inflow, and the tangential shear stress is reduced at the rotating disk and increased at the stationary housing (similarly for the radial shear stresses). The influence of inflow on shear stresses is more pronounced for larger gap width. The windage loss may be defined as follows:

$$\text{Power} = \int_{r_i}^{r_o} 2\pi r^2 \omega \tau_{td} dr \quad (32)$$

where τ_{td} is the tangential stress on the rotating disk, r_o and r_i are the outside and inside radii respectively, and r is the radial direction. From equation (32) it can be seen that the power loss due to windage is reduced due to inflow since τ_{td} is reduced. Von. F. Shultz-Grunow [20] proposed the following expression for τ_{td} :

$$\tau_{td} = \mathcal{F}_t \rho \frac{(r\omega - U)^2}{2} \quad (33)$$

where $\mathcal{F}_t$ is friction factor defined as:

$$\mathcal{F}_t = 0.0996 (\text{Re})^{-1/5} \times \left(\frac{1}{1 - r/r_o} \right)^{2/5} \quad \text{Turbulent} \quad (34)$$

$$\mathcal{F}_t = 3.66 (\text{Re})^{-1/2} \times \left(\frac{1}{1 - r/r_o} \right) \quad \text{Laminar} \quad (35)$$

Here, U is the tangential velocity. The knowledge of tangential velocity profile is needed in order to evaluate τ_{td} . The equation of motion in tangential direction may be written as follows:

$$\rho_g V \frac{dU}{dr} - \rho_g \frac{U V}{r} = (\tau_{td} - \tau_{tw})/s \quad (36)$$

Here, V is the radial velocity, s is the gap between the rotor and the stationary housing, and τ_{tw} is the tangential shear stress on the wall of stationary housing defined as follows:

$$\tau_{tw} = \mathcal{F}_t \rho_g \frac{U^2}{2} \quad (37)$$

Equations (33) to (37) will be used to evaluate the tangential velocity, then the power may be estimated using equation (32). A 4th order Runge-Kutta method was used to solve equation (36). Equation (32) was solved numerically using the composite trapezoidal rule.

6.1.4. Liquid Acceleration:

The largest contributor to power consumption is the acceleration of the liquid to the top speed of the rotor. The theoretical power required to bring the liquid to angular velocity of the rotor may be calculated as follows:

$$\text{Power} = \text{Change in Kinetic Energy} = (K.E.)_{out} - (K.E.)_{in}$$

Assuming that the liquid enters the rotor with no angular velocity, $(K.E.)_{in} = 0$, the power may be estimated as:

$$\text{Power} = (K.E.)_{out} = I\omega^2/2 \quad (38)$$

Here, I is the moment of inertia $= m r_{out}^2$, where m is the liquid flowrate, and r_{out} is the outside radius of the rotor. ω is the angular velocity of the rotor $= 2\pi N$, where N is the rotational speed.

6.2. Experimental Results and Discussion:

Figure 18 shows the total power consumption as a function of liquid flowrate at three different accelerations.

In order to sort out the factors contributing to the power consumption, experiments were performed to isolate the contributors.

6.2.1. Bearing Friction:

To isolate the bearing contribution of power consumption, the rotor was dismantled from the shaft, and the power was measured as the shaft was allowed to rotate in the open housing. Figure 19 presents the results. It can be seen that it is reasonable to assume that the bearing frictional loss is directly proportional to rotational speed, $\text{Power} = 0.2 \text{ (rpm)}$, for our equipment.

6.2.2. Windage Effects:

To investigate the windage effects, the power measurements were made at varying acceleration fields (varying speed of rotation) with the liquid and gas flows off. The bearing power consumption was subtracted. The experimentally obtained values of the power were compared with the values estimated by

Figure 18. Total Power as a Function of Liquid Flowrate

+ 600 rpm ▲ 840 rpm ● 960 rpm

Figure 19. Windage and Bearing Power Consumption vs Rotational Speed

equations (20) and (21) proposed by Von. F. Shultz-Grunow [20]. The model predicts extremely low values for the windage effects compared with the experimental values (order of magnitude less at low speeds). The model predicts that the windage loss varies with the cube of rotational speed. However, Figure 19 shows a $3/2$ power dependency of power consumption on speed, $\text{Power} = 8.79 \times 10^{-3} (\text{rpm})^{1.5}$. One of the factors for the low prediction may come from the mismatch of the apparatus setup. Shultz-Grunow's disk was free within housing, while in our setup the rotor is engaged to a lip seal, which in turn, tends to increase the power consumption due to frictional loss.

6.2.3. Windage Effect with Inflow:

The power consumption was measured with gas flow-on and liquid flow-off at varying accelerations. The bearing power consumption was subtracted. These experimentally obtained values of power were compared with the estimated values of the windage effect with inflow as discussed in Section 6.1.3. The results are presented in Figure 20. The rotational speed was varied from 400 to 1200 rpm, and the gas flow rate ranged from 0.34 to 1.34 kg/sec-m².

6.2.4. Liquid Acceleration:

Experiments were carried out to measure the power at varying rotational speeds at three different liquid flowrates, 1.40, 1.0, and 0.65 liter/sec, and zero gas flowrate. The bearing and windage power consumption were subtracted. The results are shown in Figure 21.

Figure 20. Comparison of Measured and Calculated Windage with Inflow

Figure 21. Power Consumed for Liquid Acceleration

6.2.5. Total Power Consumption:

Based on the results of all the components contributing to the power consumption, the total power may be written as:

$$\text{Total Power} = a\omega^2 + b\omega^{1.5} + c\omega$$

where a, b, and c are constants. The total power was calculated from the empirical equation derived for the bearing and windage effects, and the theoretical power required for the liquid acceleration (change in kinetic energy = $I\omega^2/2$). The calculated values are compared with the experimentally obtained values in Figure 22. The difference between the calculated and experimental values is, partly, the result of loss of energy through conversion of mechanical energy to heat. Note also that a reduction in total power consumption is obtained when the gas flows through the rotor. This norm is mainly a result of the inflow effect as discussed earlier.

Figure 22. Total Power
(contribution of all components)

7. Residence Time Distribution

7.1 Theory:

Residence time distribution (RTD) analysis has been a valuable tool in obtaining information on the fluid motion and mixing in a continuous flow system [25,26]. The experimental technique is a stimulus-response technique. The stimulus is a tracer input into the fluid entering the vessel, whereas the response is a time record of the tracer leaving the vessel. A perfect tracer has exactly the same flow properties as the substance it represents, yet it is sufficiently different in some nonflow attribute that it can be detected by an analytical instrument. In aqueous systems the choice usually falls on electrolytes. When a slug of tracer enters the bed, velocity variations, molecular diffusion and turbulent mixing all tend to spread the measured response of the tracer at the bed outlet. The term dispersion is used to denote the combined action of all the dispersive phenomena. Numerous models have been used in the literature to simulate systems with varying degrees of dispersion [26-32]. The proposed models can be divided into two groups: a dispersion model which has been used for empty tubes and packed beds, and a combined model in which complete mixing and plug flow regions are located in series and parallel arrangements. It has been shown in the literature [30,31] that the dispersion model can successfully simulate the behavior of devices in which complex radial and axial flow and transport patterns exist.

7.1.1. Dispersion Model:

The dispersion model assumes a plug flow pattern with no radial variations in velocity and concentration. The dispersion occurs in the axial direction and the extent of the dispersion is sufficient to account for the combined effects of all

dispersive phenomena (turbulent mixing, molecular diffusion, and nonuniform velocities) in the device. The spread in the impulse response takes place in the axial direction with an effective diffusivity, D_e . For $D_e=0$ the model reduces to ideal plug flow, indicating no axial dispersion. When $D_e \rightarrow \infty$, axial dispersion dominates and the model approaches to perfect mixer conditions.

The derivation of the model equations ranging from the most general to the most restricted are discussed in the literature [30–32]. The dimensional analysis of the dispersion model results in a dimensionless group: the Peclet number, defined as follows:

$$Pe = \frac{\bar{V}L_0}{D_e} \quad (39)$$

where $\bar{V}$ is superficial velocity, L_0 is a characteristic length of the device. The degree to which axial dispersion influences the performance of the device is determined by the value of the Peclet number, Pe . A high value of Pe corresponds to a low dispersed flow ($Pe \rightarrow \infty$, or $D_e \rightarrow 0$), while a low value of Pe represents a high dispersed flow ($Pe \rightarrow 0$, or $D_e \rightarrow \infty$).

Statistical techniques such as method of moments may be used for interpretation of the experimental data. The following mathematical expressions are used to characterize the data:

$$\bar{t} = \left(\int_0^{\infty} t C(t) dt \right) / \left(\int_0^{\infty} C(t) dt \right) \quad (40)$$

$$\sigma^2 = \left(\int_0^{\infty} (t - \bar{t})^2 C(t) dt \right) / \left(\int_0^{\infty} C(t) dt \right) \quad (41)$$

Here, $\bar{t}$ is the mean value or the centroid of the distribution, σ represents the spread of the distribution, and $C(t)$ is the measured response at the outlet of the device for the tracer impulse injected at the inlet. However, since it is not always possible to inject a tracer as a perfect impulse, an imperfect pulse method (also known as two peak method) may also be employed. The technique involves monitoring the concentration at two points in the test section (the bed inlet and outlet) instead of at a single point. The injection point may be located anywhere upstream of the monitoring stations. The convolution properties may be used to calculate the mean residence time and the variance of the device as follows:

$$\bar{t}_{\text{bed}} = \bar{t}_{\text{outlet}} - \bar{t}_{\text{inlet}} \quad (42)$$

$$\sigma_{\text{bed}}^2 = \sigma_{\text{outlet}}^2 - \sigma_{\text{inlet}}^2 \quad (43)$$

Once $\bar{t}$ and σ^2 are known for the bed, it can be shown that Peclet number, Pe , may be evaluated by the following expression:

$$Pe = \frac{2\bar{t}^2}{\sigma^2} \quad (44)$$

7.1.2. Liquid Holdup and Film Thickness:

One of the promising features of rotating beds is their short residence time and low liquid holdup. The residence time distribution data provides some indication of liquid holdup.

The liquid holdup in rotating beds may also be estimated using rotating disk analysis. The velocity profiles and the film thickness of liquid film on a rotating disk, a problem related to the spin coating procedure in the semiconductor industry, have been studied by several investigators [33–37]. Methods such as polynomial approximation of velocity components and the momentum integral method have been used to describe the liquid flow on a rotating disk. One of the interesting results of the momentum integral method is the analogy of the asymptotic solution of film thickness for a rotating disk with that of gravity liquid film flow along an inclined plane [22]. The asymptotic solution of liquid flow along a rotating disk has the following form:

$$\delta = [3Q'\mu_1 / \omega^2 r \rho_1]^{1/3} \quad (45)$$

Here, δ is the liquid film thickness, Q' is the liquid flowrate per unit width, μ_1 is the liquid viscosity, ρ_1 is the liquid density, ω is the speed of rotation, and r is the radial distance. In the gravity flow the centrifugal acceleration, the $\omega^2 r$ term, is replaced with the gravitational acceleration, g .

The liquid film thickness, derived from the above analysis of a rotating disk, may be related to the liquid holdup in a rotating bed by the following expression:

$$a_w \times \delta = h_o / \dot{V} \quad (46)$$

Here, h_o is the liquid holdup, $\dot{V}$ is the total packing volume, and a_w is the wetted surface area of the packing. The value of a_w may be estimated using the equation (27). The estimated values of the liquid holdup, using above analysis, may be compared with the values evaluated from RTD experiments.

7.2. EXPERIMENTAL RESULTS AND DISCUSSION:

Typical stimulus-response curves for a rotating bed are shown in Figure 23 (see the Appendix for all the experimental runs). The output trace is noisier than the input, which is not surprising given that the output electrode has a trough shape and collects only a thin non-uniform liquid film thickness. The electrode responses do not reveal any indications of channeling (extra peak as a result of bypassing liquid) or stagnancy (long tail because of a stagnant region). It is believed that the appearance of shallow tails in some of the responses may not be suggestive of a stagnant region in the bed; rather, they are caused partly as a result of slow wash-off of the salt solution on the electrodes surfaces. Proper design of the electrodes (use of teflon material) eliminated these problems to a great extent. The data was smoothed, and the mean residence time was calculated from equations (40) and (42).

7.2.1. Gravitational Effect:

Figure 24 shows the dependency of mean residence time on rotational speed. As expected, the mean residence time decreases as the rotational speed increases (at fixed liquid flowrate and no gas flow). The mean residence time varied from

Figure 23. Typical Tracer Response

Figure 24. Effect of Rotational Speed on Mean Residence Time
 + 9 kg/sec-m² Δ 27 kg/sec-m² ○ 36 kg/sec-m²

0.4 to 1.8 seconds. The outlet responses for high rotational speed were very noisy (particularly at 1000 and 1200 rpm). This is probably a result of more pronounced windage effects in higher rotational speeds, causing nonuniform liquid film over the trough shaped output electrode.

7.2.2. Liquid and Gas Flowrates Effects:

Figure 25 shows the dependency of mean residence time on liquid flowrate. At constant rotational speed and no gas flow, the mean residence time decreases as the liquid flowrate increases.

More tests were performed to investigate the effects of the gas flow on liquid residence time distribution. The results show that, statistically, the flowrate does not have significant effects on the distribution. However, the output response tends to be narrower in case of the gas flow-on, indicating shorter residence time. Figure 26 illustrates this effect.

7.3. Comparison with Models:

Equations (40) to (44) were used to evaluate Peclet number, Pe . Peclet numbers ranging from 0.1 to 3 were found, indicating highly dispersed flow conditions. Because of the large extent of backmixing (dispersion), as indicated by the low Pe , it is unlikely that the assumptions and simplifications of the dispersion model will be satisfied [33] for the rotating bed analysis. The mixing patterns in rotating beds are too complex. However, attempts could be made to develop a mixing model for rotating beds based on the combined model.

**Figure 25. Liquid Flowrate Effect
on Mean Residence Time**

**Figure 26. Effect of Gas Flowrate
on Mean Residence Time**

Equations (45) and (46) were used to estimate the liquid holdup. The residence time distribution data also provides some indication of liquid holdup. From the known liquid flowrate, the liquid holdup may be estimated: the mean residence time multiplied by the flowrate gives the apparent holdup. These values are compared and plotted in Figure 27. The deviation at the higher speed (lower liquid film thickness) is because the asymptotic solution, equation (45), does not give a good prediction of liquid film thickness at high speed of rotation[14].

Figure 27. Comparison of Experimental and Estimated Values of Liquid Holdup

8. CONCLUSIONS and RECOMMENDATIONS:

In spite of some similar trends in pressure drop characteristics between a rotating packed bed and a conventional packed tower, the experimental results of the pressure drop indicate that they are some substantial differences in hydrodynamics behavior of gas-liquid in these two devices.

The total pressure drop is the sum of the pressure drop inside of the rotor and the pressure drop outside the rotor (between the rotor and the stationary housing). In this study, up to 10% of the total pressure drop is contributed by the pressure drop outside of the rotor. The agreement between the calculated values and experimental data for the pressure drop was within $\pm 20\%$ error. It would also be very useful to measure the pressure at different positions along the housing to investigate the effects on the pressure drop measurements. The measurements may be improved if a five-hole Pitot tube is used to account for the static and dynamic pressure for the three-dimensional flow between the rotor and the housing (making sure that the dynamic pressure component is not contributing to the total pressure drop measurements).

The experimentally measured height of a transfer unit decreases with the gravitational field with 0.30 to 0.4 power. This means that the volumetric mass transfer coefficient should rise with 0.6 to 0.7 power. This finding is in agreement with the data of Vivian et al [6] and within the range estimated by Dudukovic et al [14,15]. The height of a transfer unit, HTU, is found to be 13 to 30 times smaller in the rotating bed, as compared with that of conventional towers over the same range of liquid mass velocity. The magnitude of the end effects may be studied in

more detail by estimating these effects in all the possible regions that contribute to the end effects. This may provide some information about the discrepancy observed in the measurements of volumetric mass transfer coefficients at different packing specific surface areas.

The predicted values of HTU, using the penetration theory modified for rotating beds, showed a reasonable agreement with the experimentally obtained values of HTU at varying operating conditions.

The required total power, in the range of the experimental operating conditions, was under 1 joule/gram of liquid. The largest contributor to power consumption is the acceleration of the liquid to the top speed of the rotor. The power consumption due to the bearing friction is proportional with the rotational speed. The introduction of the gas reduces the overall power required for rotating beds.

Mean residence times on order of 0.4 to 2 seconds were obtained for the rotating bed as the rotational speeds were varied from 400 to 1200 rpm. The mean residence time decreases as the speed of rotation is increased, or the liquid flowrate is increased. The gas flowrate tends to reduce the liquid mean residence time at constant liquid flowrate and rotational speed.

9. NOMENCLATURE

a_e	effective gas-liquid interfacial area (cm^2/cm^3)
a_p	packing specific surface area (cm^2/cm^3)
a_r	rotational acceleration (cm/sec^2)
a_w	wetted surface area (cm^2/cm^3)
A	coefficient for Ergun's or Morton's model
B	coefficient for Ergun's or Morton's model
C	liquid concentration (moles/ cm^3)
C_o	interfacial concentration (moles/ cm^3)
d_p	packing diameter (cm)
$\mathcal{D}$	diffusion coefficient (cm^2/sec)
f_r	friction factor in radial direction
f_t	friction factor in tangential direction
g	gravitational constant (cm/sec^2)
$\bar{G}$	molar gas flowrate (gmole/sec)
h	axial height of packing (cm)
H	tower height (cm)
$\hat{H}$	Henry's constant (atm/gmole/ cm^3)
h_o	liquid holdup (cm^3)
HTU	height of a transfer unit (cm)
I	moment of inertia (g-cm^2)
k_l	liquid side mass transfer coefficient (cm/sec)
L	superfacial liquid mass flowrate ($\text{g}/\text{sec-cm}^2$)
$\bar{L}$	molar liquid flowrate (gmole/sec)

NTU	number of transfer units
Q	liquid volumetric flowrate (cm ³ /sec)
P	pressure (Pa)
ΔP	pressure drop (Pa)
r	radial distance (cm)
r_i	rotor inside diameter (cm)
r_o	rotor outside diameter (cm)
R	disk radius (cm)
s	gap between the rotor and the stationary housing (cm)
S	empty tower cross section (cm ²)
t	time (sec)
T	torque (dyne-cm)
$\bar{t}$	mean residence time (sec)
U	velocity in tangential direction (cm/sec)
V	velocity in radial direction (cm/sec)
$\bar{V}$	superficial velocity (cm/sec)
$\hat{V}$	packing volume (cm ³)
V_i	interfacial velocity (cm/sec)
x	axial direction (cm)
α_e	equilibrium concentration (gmole/cm ³)
α_i	liquid CO ₂ concentration in the rotor inlet (gmole/cm ³)
α_o	liquid CO ₂ concentration in the rotor outlet (gmole/cm ³)
y	CO ₂ concentration in the gas (gmole/cm ³)
y_i	CO ₂ concentration in the gas inlet (gmole/cm ³)

z	axial direction
Z	characteristic size of the packing (cm)

Greek Notations:

ϵ	packing void fraction
φ_s	packing particle sphericity
ρ_l	liquid density (g/cm ³)
ρ_g	gas density (g/cm ³)
μ_l	liquid viscosity (g/cm sec)
μ_g	gas viscosity (g/cm sec)
τ	shear stress (Pa)
ω	rotational speed (rpm)
σ	variance of residence time distribution (sec)
$\bar{\sigma}$	liquid surface tension (dyne/cm)
σ_c	critical surface tension of packing (dyne/cm)
δ	liquid film thickness (cm)

Dimensionless Groups:

Fr	Froude number= $L^2 a_p / \rho^2 g$
Gr	Grashof number= $Z^3 \rho^2 g / \mu^2$
Re	Reynolds number= $L / a_p \mu$
Re _l	Reynolds number= $4L / \mu a_e$
Sc	Schmidt number= $\mu / \rho \mathcal{D}$
We	Weber number= $L^2 / \rho \bar{\sigma} a_p$

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11. APPENDIX

Section 11.1 shows the stress analysis made for the foam metal packing. Section 11.2 shows the calculations for the shaft design. Section 11.3 presents the the pressure transducers specifications and the calibration curves for the liquid and gas orifices. Section 11.4 presents the representative results of the mass transfer experimental runs.

11.1 Stress Analysis for Aluminum Foam Metal Packing:

When a circular disk rotates, the inertia forces can induce stresses that may become very large at high speed.

Consider a rotating disk with inner radius r_i and outer radius r_o , and a rotational speed of N . The maximum values of radial and tangential stresses can be calculated using the following equations:

$$(\sigma_r)_{\max} = \frac{\rho V^2}{g} \times \frac{3 + \nu}{8} \left(1 - \frac{r_i}{r_o}\right)^2$$

$$(\sigma_t)_{\max} = \frac{\rho V^2}{g} \times \frac{3 + \nu}{4} \left(1 + \frac{1 - \nu}{3 + \nu} \times \left(\frac{r_i}{r_o}\right)^2\right)$$

where,

ρ is the density of the aluminum foam metal

V is the linear velocity $= 2\pi N r_o$

ν is the Poisson ratio $= 0$ for the aluminum foam metal

For the our bed specifications, at 2000 rpm, σ_r and σ_t were 5 psi and 73 psi, respectively.

The permissible tensile stress for the aluminum foam metal, as reported by the manufacturer, is 165 psi. Thus, the packing is strong enough to withstand the applied stress at 2000 rpm (gravitational field of 1000g, based on outside radius)..

11.2. Shaft Design:

Shaft design requires two steps: 1– establishing the minimum shaft diameter, 2– determining the natural frequency of vibration for the shaft.

The bending moment, M, and the Torque, T, can act simultaneously; these loads can be combined and resolved into shear stress and tensile stress that can act on the shaft:

$$\text{Torque} = T = \frac{\text{Power}}{\text{Rotational Speed}} = \frac{P}{N}$$

$$\text{Bending Moment} = M = \frac{\text{Power} \times \text{Shaft Length}}{\text{Rotational Speed} \times \text{Rotor Diameter}} = \frac{P \times L}{N \times D}$$

The minimum shaft diameter will be the greatest of the two values calculated from the following equations to meet the shear stress or tensile stress, respectively:

$$d_s = \left(\frac{16 \times \sqrt{T^2 + M^2}}{\pi \sigma_s} \right)^{1/3}$$

$$d_t = \left(\frac{16 \left[M + \sqrt{T^2 + M^2} \right]}{\pi \sigma_t} \right)^{1/3}$$

Here, σ_s and σ_t are the allowable shear and tensile stresses.

For known rotor specifications, required power, and rotational speed, the diameter of the shaft may be determined. However, this value of the shaft also must meet the requirement of the natural frequency analysis, given by the following expression:

$$N_c = \frac{d^2 F_m}{L \times \sqrt{w \times (L + S_b)}}$$

where,

N_c natural frequency

d shaft diameter

w weight of the rotor

L shaft length

S_b distance between the bearings

F_m material properties factor

The operating speed of the shaft must not exceed the value of N_c calculated from the expression.

11.3. Calibration of the Transducers and Orifices:

To measure the pressure drop across the orifices, differential pressure sensors (SenSym, SCX Series) were used. The sensitivity for the sensors used for the orifices is 6 MV/psi. The full-scale span is 90 MV. The recorded data for various experimental runs were in range of 0–60 MV for the gas orifice and 0–9 MV for the liquid orifice.

The calibration of the gas orifice was carried out using: i) the orifice equation, ii) a rotameter, iii) a pitot tube. The calibration curves for these units are presented in Figures A1, A2, and A3, respectively. The calibration curves for (i) and (ii) cases have the second order feature, and they predict similar flowrate values (within 3%). The calibration curve for the pitot tube does not show a second order functionality which is characteristic for orifice meters and the predicted flowrates are off by more than 50% compared to those of (i) and (ii) cases. A least square fit was employed for the obtained data, and the error analysis shows an error band of $\pm 0.9 \text{ ft}^3/\text{min}$. Considering $\pm 5\%$ full scale accuracy for the rotameter, the measured flowrates are accurate within 6% to 7%. For this to be true, we assume that pressure transducer's inaccuracy is pressure independent; the 5% full scale accuracy for the rotameter is flow independent, and the error propagation is linear and can be superposed. Figure 4A shows the linearity of the pressure transducer response in regard to the differential pressure.

Figure A5 shows the calibration of the pressure sensor used to measure the pressure drop across the bed. A monometer was used to calibrate this pressure transducer. This sensor has a sensitivity of 18 MV/psi, and it has a full scale span

of 18 MV. The liquid orifice was calibrated by weighing the liquid output from the rotor for the periods of 20 second to 30 seconds. Figure A6 shows the calibration curve for the liquid orifice.

Figure A.1. Calibration of Pressure Sensor
Using the Orifice Equation

Figure A.3. Calibration of Pressure Sensor
Using PH α Tube

Figure A.4. Linear Response of the
Pressure Sensors

Figure A.5. Calibration of Pressure Sensor
Used Measuring Bad Pressure Drop

Figure A.6. Calibration of Pressure Sensor
Used for Liquid Qrimea

11.4. Mass Transfer Experimental Runs:

Desorption of carbon dioxide from water into air in rotating packed bed was studied. The carbon dioxide was dissolved in water by dissolving sodium bicarbonate in distilled water and then acidifying the solution to a pH of 5.0 (at pH between 4.8 and 5.2 all the carbonate and bicarbonate are converted to the CO₂). Water containing dissolved carbon dioxide was pumped from an overhead storage tank (760 liters capacity) through an orifice to the inner edge of the rotor.

The data from Dodds et al [A1] were employed for the solubility of CO₂ in water. The accuracy of these data was not very significant in this investigation, since the liquid concentration in equilibrium with gas was quite small compared with the CO₂ concentrations in the main body of the liquid. The equilibrium data was used to compute the hypothetical liquid concentration in equilibrium with the gas phase concentrations at the inlet and outlet of the rotor. In a few cases, gas analyses were taken for the purpose of material balance calculations, as a test of the accuracy of the data. The gas-phase concentration in the entering air was found by Gas Chromatograph to be about 5.5×10^{-4} atmospheres, about twice that of the ordinary atmosphere (3.3×10^{-4}).

TABLE A.1. MEASURED HTU AT VARYING ACCELERATION FIELDS
FOR 1476 m²/m³ PACKING

LIQUID FLOWRATES: 36 Kg/s-m ² , GAS FLOWRATES:1.0 Kg/s-m ²	
HTU (cm)	# g's
3.76	50
3.76	50
3.80	50
3.70	50
2.74	100
2.80	100
3.30	100
2.40	182
2.25	182
2.70	182
2.00	272
2.00	272

TABLE A.2. MEASURED HTU AT VARYING ACCELERATION FIELDS
FOR 1476 m²/m³ PACKING

LIQUID FLOWRATE :18.0 Kg/s-m ² , GAS FLOWRATE: 1.0 Kg/s-m ²	
HTU (cm)	# g's
2.98	50
3.25	50
3.00	50
3.10	50
2.98	50
2.25	100
2.11	100
2.25	100
2.36	100
1.95	182
1.98	182
1.95	182
2.00	182
1.80	272
1.79	272
1.80	272

TABLE A.3. MEASURED HTU AT VARYING ACCELERATION FIELDS
FOR 2952 m^2/m^3 PACKING

LIQUID FLOWRATE: 36 Kg/s- m^2 , GAS FLOWRATE :1.0 Kg/s- m^2	
HTU (cm)	# g's
2.65	50
2.71	50
2.52	50
2.22	50
2.28	50
2.28	50
2.28	50
1.94	100
1.92	100
2.07	100
2.01	100
1.83	100
2.07	100
1.70	182
1.67	182
1.70	182
1.64	182
1.40	272
1.21	272
1.22	272

TABLE A.4. MEASURED HTU AT VARYING ACCELERATION FIELDS
FOR 656 m^2/m^3 PACKING

LIQUID FLOWRATE: 36 Kg/s-m^2 , GAS FLOWRATE :1.0 Kg/s-m^2	
HTU (cm)	# g's
2.74	50
2.74	50
2.67	50
2.74	50
2.82	50
2.74	50
2.20	100
2.20	100
2.01	182
2.00	182
2.06	182
1.95	272
1.92	272
1.86	272

11.5. Residence Time Distribution Experimental Runs:

The normalized electrode responses for varying liquid and vapor loading, and rotational speed are presented in Figures 7A to 28A. The Y-axis represents the electrode response in millivolts, while the X-axis is time in seconds. The solid lines, drawn through the noisy part of the curves, indicates the effort made to smooth the data. A digitizer was used to transfer the smoothed points to a personal computer for the data analysis.

Figure 7A. $L = 36 \text{ Kg/s-m}^2$, $N = 400 \text{ rpm}$

Figure 8A. $L = 36 \text{ Kg/s-m}^2$, $N = 600 \text{ rpm}$

Figure 9A. $L = 36 \text{ Kg/s-m}^2$, $N = 800 \text{ rpm}$

Figure 10A. $L = 36 \text{ Kg/s-m}^2$, $N = 1000 \text{ rpm}$

Figure .11A. $L = 36 \text{ Kg/s-m}^2$, $N = 1200 \text{ rpm}$

Figure 12A. $L = 27 \text{ Kg/s-m}^2$, $N = 400 \text{ rpm}$.

Figure 13A. $L = 27 \text{ Kg/s-m}^2$, $N = 600 \text{ rpm}$

Figure 14A. $L = 27 \text{ Kg/s-m}^2$, $N = 800 \text{ rpm}$.

Figure 15A. $L = 27 \text{ Kg/s-m}^2$, $N = 1000 \text{ rpm}$

Figure 16A. $L = 27 \text{ Kg/s-m}^2$, $N = 1200 \text{ rpm}$

Figure 17A. $L = 9 \text{ Kg/s-m}^2$, $N = 600 \text{ rpm}$

Figure 18A. $L = 9 \text{ Kg/s-m}^2$, $N = 800 \text{ rpm}$

Figure 19A. $L = 9 \text{ Kg/s-m}^2$, $N = 1000 \text{ rpm}$

Figure 20A. $L = 9 \text{ Kg/s-m}^2$, $N = 1200 \text{ rpm}$

Figure 21A. $L=36 \text{ Kg/s-m}^2$, $N=1000 \text{ rpm}$, $G=0.34 \text{ Kg/s-m}^2$

Figure 22A. $L=36 \text{ Kg/s-m}^2$, $N=1000 \text{ rpm}$, $G=1.0 \text{ Kg/s-m}^2$

Figure 23A. $L=36 \text{ Kg/s-m}^2$, $N=1000 \text{ rpm}$, $G=1.3 \text{ Kg/s-m}^2$

Figure 24A. $L=9 \text{ Kg/s-m}^2$, $N=600 \text{ rpm}$, $G=0.34 \text{ Kg/s-m}^2$

Figure 25A. $L=9 \text{ Kg/s-m}^2$, $N=600 \text{ rpm}$, $G=1.0 \text{ Kg/s-m}^2$.

Figure 26A. $L=36 \text{ Kg/s-m}^2$, $N=600 \text{ rpm}$, $G=0.34 \text{ Kg/s-m}^2$

Figure 27A. $L=36 \text{ Kg/s-m}^2$, $N=600 \text{ rpm}$, $G=1.0 \text{ Kg/s-m}^2$.

Figure 28A. $L=36 \text{ Kg/s-m}^2$, $N=600 \text{ rpm}$, $G=1.3 \text{ Kg/s-m}^2$

11.6. Selected Coordinate System for the Control Volume:

PART II

ROTATING BED MASS TRANSFER STUDY OF SULFUR DIOXIDE SORPTION IN WATER-LIME SLURRIES

BY

Adin Coskundeniz

ABSTRACT

The effect of gravitational force upon the overall volumetric mass-transfer coefficients in rotating packed bed has been studied for sulfur dioxide absorption in water and in lime slurry. K_Ga is found to increase logarithmically with gravity for both system. The influence of gravity on K_Ga increases at lower liquid and higher gas flowrates for SO_2 -Air-Water system, varying between the power of 0.40 and 0.05. In SO_2 -Air-Lime system gravity is found to be less effective as $K_Ga \propto g^{0.08-0.11}$ and not sensitive to liquid and gas flowrates. For both systems, the overall volumetric mass-transfer coefficients obtained in rotating packed bed are much higher than the corresponding values in conventional packed towers.

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1. INTRODUCTION:

A radical approach to reduce the capital cost of a chemical plant has been process intensification. Reaction and separation processes are enhanced by the use of high gravitational, electric, acoustic, or magnetic fields. In particular, heat and mass transfer can be intensified by carrying out processes in high gravitational fields.

High gravitational fields are achieved in rotating packed beds in the form of a rigid metal foam in the shape of a disk with a hole in the center. In gas-liquid mass-transfer operations, liquid is sprayed into the center of the bed and flows outward through the bed because of the centrifugal force. The gas introduced at the outer surface of the packing is forced inward countercurrently to the liquid by the imposed pressure gradient. Intense mass-transfer occurs as high shear forces produce very thin liquid films, rapidly renewed surfaces and substantial turbulence, enhancing the volumetric mass-transfer coefficients. Rotating packed bed has a number of features which differ significantly from more conventional mass transfer devices. Benefits include:

- small size and weight
- reduced solvent inventory
- short residence time
- rapid attainment of steady state

Rotating packed bed has a potential to be used in flue gas desulfurization (FGD) which is a broadly applied technology for limiting

sulfur dioxide emissions from power plants, particularly utilizing the high-sulfur coals. Lime slurries is widely used in wet scrubbing methods.

This study is primarily concerned with the effect of gravitational force upon the overall volumetric mass-transfer coefficients for sulfur dioxide absorption in water and lime slurry.

2. PREVIOUS WORK:

Vivian et al. [1] presented a correlation for gas absorption in short wetted-wall columns:

$$\frac{k_L h}{d} = 0.443 \left(\frac{4\Gamma}{\mu} \right)^{0.4} Sc^{1/2} \left(\frac{g h^3 \rho^2}{\mu^2} \right)^{1/6}$$

Later, Davidson [2] obtained an exponent of 1/6 on the gravity group by theoretical considerations.

Onda et al. [3] proposed a dimensionless correlations for liquid-phase mass-transfer coefficients in Raschig ring columns. Although the interfacial area, a , is assumed to be independent of gravity in previous correlations, k_L and a are combined in the expression by Onda et al.

$$\frac{(k_L a) d^2}{D} = 0.023 Sc^{1/2} \left(\frac{g d^3 \rho^2}{\mu^2} \right)^{0.38} \left(\frac{d L}{\mu} \right)^{1/2} \left[1 - 1.02 \exp \left[-(0.15) \left(\frac{d L}{\mu} \right)^{0.4} \right] \right]$$

For the first time, Vivian et al. [4] experimentally studied the effect of gravity on liquid-phase volumetric mass-transfer coefficient for carbon dioxide desorption. In their study, a Raschig-ring cylindrical packed column which is mounted on a rotating centrifuge was used in 1 to 6.4 times the gravity force. Their experimental results showed that $k_L a$ varies with gravity to a power between 0.41 and 0.48, the effect of gravity increasing as the liquid flow rate decreases.

ICI, Imperial Chemical Industries [5,6] introduced the HIGEE packed bed

operated at mean acceleration fields of 200 to 500 g, and reported that volumetric mass transfer coefficient varied with the gravitational force to the 0.14 to 0.54 powers.

Duducovic et al [7,8,9,10] studied the effect of rotational speed on gas-liquid interfacial area and the liquid-side gas-liquid mass-transfer coefficients, separately. They concluded that the volumetric mass-transfer coefficient increases with the gravitational force as a fractional power of 0.30 to 0.37 which is in agreement with the data of Vivian et al. and within the range estimated by ICI. It is shown that experimentally measured gas-liquid interfacial area increases with the gravity as $\propto g^{0.14}$ to $g^{0.21}$ depending on the packing type and liquid flow rate. The liquid-side gas-liquid mass-transfer coefficient, k_L , varies with gravity as $k_L \propto g^{0.16}$. Power dependency of 1/6 also matches the results obtained from the theoretical predictions based on film flow on rotating disk and rotating blades.

Desorption of carbon dioxide from water into air in rotating packed bed was studied by Keyvani [11]. Results indicates that liquid-phase volumetric mass-transfer coefficient increases with 0.3 to 0.4 power of gravity.

3. EQUIPMENT and PROCEDURE:

3.1. Schematic of Rotor:

A simplified schematic of the rotor is shown in Figure 1. The liquid entering from the inner pipe of the stationary double pipe (1) is distributed on the inside edge of the bed (2) and then is accelerated into the housing (3) by the effect of centrifugal force. Due to imposed pressure gradient, the gas entering the housing is flown inward countercurrently to the liquid through the rotating bed. The liquid collects in the bottom of the housing and serves as a gas trap. The gas exits on the inside edge of the bed through the double pipe (4).

3.2. Packing:

The rotating packed bed consists of a packing unit, made up of a nickel plated aluminum foam metal in the shape of a torous. The outside diameter is 18 inches and the thickness is 1.75 inches. The packed bed used in the experiments has specific surface area of $900 \text{ ft}^2/\text{ft}^3$ and porosity of 0.92. The number of pores per inch is 40.6 and pore size is 0.01. Packing unit is dimensionally stable in the high gravitational fields.

3.3. Liquid Distributor:

The liquid distributor has been designed to eliminate the liquid entrainment problems and to provide even liquid distribution at the inner edge of the rotating bed where entrainment of liquid by gas is most probable to occur due to high gas and low liquid flow rates. Distributor consists of a

Figure 1. Schematic of Rotating Bed

stainless steel pipe ($3/4$ " ID, $3-1/2$ " long) with a flattened end. About $1/16$ "-thick liquid sheet is produced by using flattened end which forms a $1-1/2$ " by $1/16$ " slit. The liquid distributor is placed $1/2$ " above the inner edge of the rotating bed.

3.4. Rotor Assembly and Seals:

The rotor assembly is holded by a 2 inches diameter-shaft which is supported by two bearings. The assembly is driven by a 5 hp-motor (3 phase, 230 V) with a belt and a speed controller (Parajust Model M01712-A00).

Silicon rubber radial lip seals are placed in the shaft and the double pipe-gas exit assembly.

3.5. Flow Diagram:

Flow diagram of the experimental apparatus which have been used in the absorption of SO_2 both in water and lime slurry, is shown in Figure 2. The maximum liquid flow rate of 30 gpm can be achieved by using the feed and recycle pumps. Up to 60 acfm of air can be delivered to the system by the compressor used. A static mixer is placed in the piping system in order to provide a well-mixed SO_2 -Air mixture before gas inlet.

3.6. Measurement of Operating Parameters:

The gas flow rates were measured by correlating the pressure drop across restriction orifices which was determined by using differential pressure transducers. Pure SO_2 flow rate is adjusted by a rotameter.

The liquid flow rate is determined by Magnetic Flowmeter (Magnavox

Figure 2. Experimental Apparatus

Model Mag 3570) which is capable of measuring flowrates up to 10 gpm.

Orifice has not been used for measuring lime-slurry flow because of its highly abrasive nature.

A pressure gage was used to monitor the pressure of the liquid entering the rotating bed.

In order to measure pressure drop across the bed, probes for the transducers were placed at the inlet and outlet of the rotor as shown in Figure 1.

Thermocouples are placed in the inlet and outlet of the rotor to monitor the temperature across the bed which has great influence on the solubility data.

The torque required for the rotating bed was measured by a sprocket/pulley torque transducer (Lebow Model 1388-102) which is connected to the end of the drive shaft. The speed of rotation was measured by a magnetic pick-up tachometer (Amtex Model 1736).

Sulfur dioxide concentrations in the gas mixture were determined by infrared SO_2 analyzer (Horiba Model VIA-300) which is continuously connected to the inlet and outlet of the rotor.

3.7. Data Acquisition:

The gas and liquid flow rates, pressure drop across the bed, rotational speed and torque are interfaced to a personal computer (IBM-XT). Labtech Notebook, a menu driven software package is used for data acquisition in actual units.

4. THEORY:

4.1. Two Film Theory:

The two film theory, originally proposed by Whitman [12] is widely used to explain mass transfer operations, such as absorption. In this theory, the interface between the gas phase and the liquid phase is assumed to consist of a gas film adjacent to a liquid film. It is presumed that the main body of gas and liquid phase are completely mixed by turbulent flow, while the gas and liquid film are laminar or stagnant. There is no solute concentration gradient in the main body of either phase, but a solute concentration gradient exist across both the gas film and the liquid film. At the interface, solute concentration in the gas film is assumed to be in equilibrium with the solute concentration in the liquid film. There is no resistance to mass transfer at the interface. In the absorption operations where solute is transferred from a vapor phase to a liquid phase, the solute concentration in the vapor phase should be greater than the equilibrium concentration in the liquid phase in order to drive the mass transfer. A solute in the main body of gas phase should first pass through the gas film, then through the liquid film, and finally dissolve into the main body of liquid.

The solute transferred per unit time from the main body of gas phase through the gas film is:

$$N = A Z k_G a (p - p') \quad (1)$$

Partial pressure of solute in the gas phase is related to mole fraction as follows:

$$p = y P_T \quad (2)$$

Equation (1) and (2) may be combined as:

$$N = A Z k_G a P_T (y - y') \quad (3)$$

Similarly, the solute transferred per unit time through the liquid film to the main body of liquid phase is given by equation (4):

$$N = A Z k_L a (x' - x) \quad (4)$$

where N = solute transferred (lb-mol/h)

A = column cross-sectional area (ft²)

Z = height of packed tower (ft)

k_G = gas-phase mass-transfer coefficient (lb-mol/h·ft²·atm)

k_L = liquid-phase mass-transfer coefficient (lb-mol/h·ft²·mol/mol)

a = interfacial area (ft²/ft³)

P_T = total system pressure (atm)

y = mol fraction of solute in gas phase

y' = mol fraction of solute in the gas phase at interface

x = mol fraction of solute in liquid phase

x' = mol fraction of solute in the liquid phase at interface

The interfacial area is considered to be the actual mass transfer area between liquid and gas phases. The product of a and $A Z$ gives the total interfacial area within the entire packed bed.

Since concentration differences, such as $(y - y')$, take place over extremely small distances, it is difficult to determine the values of y' and x' in practical cases. However, it is assumed that the concentrations in the gas and liquid

phases are correlated by a linear equilibrium relationship over a small solute concentration change. In this case, the rate of mass transfer is proportional to the differences between the main body concentration in one phase and the concentration which would be equilibrium with the main body concentration in the second phase. Therefore, the solute transferred can be written in terms of overall mass-transfer coefficients K_G and K_L shown in equation (5) and (6).

$$N = A Z K_G a P_T (y - y^*) \quad (5)$$

$$N = A Z K_L a (x^* - x) \quad (6)$$

Here, y^* is the solute mole fraction in the gas phase which would be in equilibrium with the liquid phase and similarly, x^* is the the solute mole fraction in liquid phase which would be in equilibrium with the gas phase. In these equations, overall gas and liquid-phase mass transfer coefficients, K_G and K_L , are combined with the interfacial area, a , to produce overall gas and liquid-phase volumetric mass-transfer coefficients, $K_G a$ and $K_L a$. Equilibrium curve and operating line in a packed absorber is shown in Figure 3. The equilibrium curve slope, m , is defined from Figure 3.

$$m_1 = \frac{y - y'}{x^* - x'} \quad (7) \quad \text{or} \quad m_2 = \frac{y' - y^*}{x' - x} \quad (8)$$

Equation (3) can be rewritten as:

$$y - y' = \frac{N}{A Z k_G a P_T} \quad (9)$$

Figure 3. Equilibrium Curve and Operating Line in an Absorber

Equation (8) is substituted into equation (4), so that

$$y' - y^* = \frac{m_2 N}{A Z k_L a} \quad (10)$$

and adding the driving forces yields:

$$y - y^* = y - y' + y' - y^* = \frac{N}{A Z k_G a P_T} + \frac{m_2 N}{A Z k_L a} \quad (11)$$

The interfacial concentration is eliminated, giving equation (12).

$$y - y^* = \frac{N}{A Z} \left(\frac{1}{k_G a P_T} + \frac{m_2}{k_L a} \right) \quad (12)$$

The substitution of equation (5) into equation (12) gives:

$$\frac{1}{K_G a P_T} = \frac{1}{k_G a P_T} + \frac{m_2}{k_L a} \quad (13)$$

Here, $K_G a$ is overall gas-phase volumetric mass-transfer coefficient expressed in the units of (lb-mol/h·ft³ · atm). This equation shows how the overall gas-phase mass-transfer coefficient is related to the individual gas and liquid-phase coefficients. The reciprocals of mass-transfer coefficients in the above equation, are regarded as the resistances to mass transfer, consisting of a gas-phase resistance and a liquid-phase resistance in series.

Similarly, the overall resistance to mass transfer in terms of liquid phase can be found as follows:

$$x^* - x = x^* - x' + x' - x \quad (14)$$

Equation (4) and (7) can be substituted into equation (14).

$$x^*-x = \frac{N}{m_1 A Z k_{Ga} P_T} + \frac{N}{A Z k_L a} \quad (15)$$

When equation (6) is taken into account, it becomes

$$\frac{1}{K_L a} = \frac{1}{m_1 k_{Ga} P_T} + \frac{1}{k_L a} \quad (16)$$

Here, $K_L a$ is the overall liquid-phase volumetric mass-transfer coefficient expressed in the units of (lb-mol/h·ft³·mol/mol).

When equilibrium line is linear in which the case of ($m=m_1=m_2$) and k_G is small compared to k_L or m is small (equilibrium line flat), solute is very soluble in the liquid. In equation (13), $m/k_L a$ term becomes minor and $1/k_G a$ term represents the major resistance. The absorption is considered to be gas-phase mass-transfer controlled in which $K_G a = k_G a$.

Conversely, when k_L is small compared to k_G or m is very large (equilibrium line vertical), solute is insoluble in the liquid. In equation (16), $1/m k_G a P_T$ term can be neglected and the major resistance ($1/k_L a$) is considered to exist in the liquid-phase ($K_L a = k_L a$). The absorption, having these characteristics are called liquid-phase controlled.

$K_G a$ is most useful in designs when composition of the gas phase is of primary interest and $K_L a$ where composition of the liquid phase is of primary interest.

4.2. Calculation of $K_L a$ and $K_G a$ in Rotating Bed:

Generally, height of a transfer unit is used to present data on absorption column performance. Because of simplicity and similarity to plate-column calculation methods, design procedures are based on this concept.

The total height of a packed bed is given by:

$$Z = (HTU) \times (NTU) \quad (17)$$

where

Z = depth of packing (ft)

HTU = height of a transfer unit (ft)

NTU = number of transfer units

Figure 4 shows the rotating packed bed together with the nomenclature that will be used in developing the equations.

When operating and equilibrium lines are linear over the concentration range of the column, it is practical to employ the use of a logarithmic mean driving force in the rate equations in which the overall volumetric mass-transfer coefficients are used.

The rate of interphase transfer can be expressed in terms of $K_L a$:

$$N = G (y_i - y_o) = L (x_o - x_i) = K_L a A Z \rho (x^* - x)_{LM} \quad (18)$$

where

$$(x^* - x)_{LM} = \frac{(x^* - x)_o - (x^* - x)_i}{\ln \frac{(x^* - x)_o}{(x^* - x)_i}} \quad (19)$$

Assuming that gas and liquid flowrates are constant and Henry's law is obeyed for the dilute system,

Figure 4. Countercurrent Absorption in Rotating Packed Bed

$$p = H c \quad (20)$$

Here, p is solute partial pressure in gas phase (atm), c is solute concentration in liquid (lb-mol/ft³), H is Henry's law constant (atm/ (lb-mol/ft³)), and H' is also Henry's law constant which is the reciprocal of H .

Henry's law constant can be correlated to the slope of equilibrium line as:

$$m = \frac{H \rho}{P_T} \quad (21)$$

The liquid phase solute concentration which is in equilibrium with gas phase in terms of m is:

$$x^* = y / m \quad (22)$$

The depth of packing can be estimated by equation (23).

$$Z = \frac{L}{K_L a A \rho} \frac{(x_0 - x_i)}{(x^* - x)_{LM}} \quad (23)$$

where

$$HTU = \frac{L}{K_L a A \rho} = \frac{L'}{K_L a \rho} \quad (24)$$

$$NTU = \frac{(x_0 - x_i)}{(x^* - x)_0 - (x^* - x)_i} \ln \frac{(x^* - x)_0}{(x^* - x)_i} \quad (25)$$

In the above equations, N is SO_2 transferred from gas phase to liquid phase

(lb-mol/h), G is the molar gas flow rate (lb-mol/h), L is molar liquid flow rate (lb-mol/h), L' is the average of the molar liquid flow rates per unit area based on the inner and outer cross-sectional area of rotating bed (lb-mol/ft² · h), P_T is total system pressure (atm), K_L is the overall liquid-phase mass-transfer coefficient (lb-mol/h · ft² · (lb-mol/ft³)) or (ft/h), A is packing cross-sectional area (ft²), ρ is the average molar liquid density (lb-mol/ft³), x_i is SO₂ mol fraction in inlet liquid, x_o is SO₂ mol fraction in outlet liquid, and x^* is the mole fraction of SO₂ in the liquid phase in equilibrium with the gas phase.

The average liquid and gas flow rates per unit area, L' and G' , are calculated from the inner and outer cross-sectional areas of rotating bed.

$$L' = L_i + L_o / 2 \quad (26) \quad G' = G_i + G_o / 2 \quad (27)$$

$$\text{For liquid, } L_i = L / A_i \quad (28) \quad L_o = L / A_o \quad (29)$$

$$\text{where } A_i = 2 \pi r_i d \quad (30) \quad A_o = 2 \pi r_o d \quad (31)$$

Here, A_i and A_o are inner and outer cross-sectional areas, r_i and r_o are inner and outer radii of packed bed.

In rotating packed bed, HTU is defined as:

$$\text{HTU} = (r_o - r_i) / \text{NTU} \quad (32)$$

NTU can be found by measuring gas inlet and outlet SO₂ concentrations and then by calculating hypothetical liquid concentrations in equilibrium with the gas phase at the inlet and outlet of the rotor from the equilibrium relation in equation (22) to be used in equation (25). Since packing depth, Z ($r_o - r_i$) in

rotating bed, and NTU are known, overall liquid-phase volumetric mass-transfer coefficient, $K_L a$, can be determined by equation (33)

$$K_L a = \frac{L}{\text{HTU } A \rho} = \frac{L'}{\text{HTU } \rho} \quad (33)$$

Similarly, the rate of solute transferred from the gas phase can be expressed in terms of $K_G a$ and a logarithmic mean driving force.

$$G (y_i - y_o) = K_G a A Z P_T (y - y^*)_{LM} \quad (34)$$

where

$$(y - y^*)_{LM} = \frac{(y - y^*)_i - (y - y^*)_o}{\ln \frac{(y - y^*)_i}{(y - y^*)_o}} \quad (35)$$

Thus, packing depth:

$$Z = \frac{G}{K_G a A P_T} \frac{(y_i - y_o)}{(y - y^*)_{LM}} \quad (36)$$

where

$$\text{HTU} = \frac{G}{K_G a A P_T} = \frac{G'}{K_G a P_T} \quad (37)$$

and

$$\text{NTU} = \frac{(y_i - y_o)}{(y - y^*)_i - (y - y^*)_o} \ln \frac{(y - y^*)_i}{(y - y^*)_o} \quad (38)$$

Here, $K_G a$ is overall gas-phase volumetric mass-transfer coefficient

(lb-mol/h-ft² · atm), y_1 is SO₂ mol fraction in gas inlet, y_o is SO₂ mol fraction in gas outlet, and y^* is SO₂ mol fraction in gas phase that would be in equilibrium with the liquid phase.

In a similar manner, HTU in equation (37) may be obtain with a known packing depth, Z , by using equation (32). NTU for gas phase can be estimated by equation (38), once SO₂ concentrations in gas phase at the inlet and outlet of rotor and also the SO₂ concentration in gas phase which is in equilibrium with the liquid phase are determined. Overall gas-phase volumetric mass-transfer coefficient, $K_G a$, can be determined from equation (39).

$$K_G a = \frac{G'}{HTU P_T} \quad (39)$$

The relationship between $K_L a$ and $K_G a$ is expressed by:

$$\frac{1}{K_L a} = \frac{1}{k_L a} + \frac{H'}{k_G a} = \frac{H'}{K_G a} \quad (40)$$

Absorption of SO₂ in water has been studied by Whitney and Vivian [13] and Pearson et al. [14]. In their studies it is shown that both liquid and gas film resistances exert an appreciable effect and absorption becomes both liquid and gas film controlled.

The equations in this section, are going to be used in determining $K_L a$ and $K_G a$ for SO₂-Air-Water system in the rotating packed bed with a countercurrent design.

4. 3. Chemistry and Absorption-Rate Steps:

In the SO_2 absorption, a portion of sulfur dioxide which is dissolved in water ionizes.

(41)

check H on
equation

(42)

(43)

The absorptivity of sulfur dioxide in an aqueous system can be increased by reducing the hydrogen ion concentration or by removing HSO_3^- or SO_3^- .

When lime used, the main reaction steps may be summarized as follows:

lime dissolution,

(44)

(45)

reaction with dissolved SO_2 ,

(46)

(47)

(48)

oxidation,

(49)

(50)

(51)

Equilibrium species in SO_2 absorption versus pH is shown in Figure 5.

In the removal of SO_2 from flue gases by lime slurry, the following

Figure 5. SO_2 Absorption species versus pH

individual steps take place.

1. Diffusion of SO_2 to the gas/liquid interface through the gas film
2. Dissolution of SO_2 in water at the interface
3. Ionization of dissolved SO_2 to H^+ and HSO_3^-
4. Dissociation of HSO_3^- to form $\text{SO}_3^{=}$
5. Diffusion of HSO_3^- , H^+ , and $\text{SO}_3^{=}$ ions from the liquid/gas interface through liquid film into the liquid interior
6. Dissolving of $\text{Ca}(\text{OH})_2$
7. Ionization of dissolved $\text{Ca}(\text{OH})_2$ to Ca^{++} and OH^- ions
8. Reaction of Ca^{++} with $\text{SO}_3^{=}$ to form Ca SO_3
9. Precipitation of $\text{Ca SO}_3 \cdot 1/2 \text{H}_2\text{O}$
10. Oxidation of $\text{SO}_3^{=}$ to $\text{SO}_4^{=}$
11. Reaction of Ca^{++} with $\text{SO}_4^{=}$ to form Ca SO_4
12. Precipitation of $\text{Ca SO}_4 \cdot 2 \text{H}_2\text{O}$

Since ionic reactions are very fast, rate controlling steps can be step 1, gas phase mass transfer, step 5, liquid phase mass transfer and step 6, dissolution of $\text{Ca}(\text{OH})_2$. Because dissolution rate $\text{Ca}(\text{OH})_2$ is fast enough and $\text{Ca}(\text{OH})_2$ has high surface area, step 6 is not controlling. High solubility of SO_2 in lime slurry due to chemical reaction makes step 5, liquid phase transfer, have little effect. The absorption of SO_2 in lime slurry becomes gas phase mass transfer limiting.

4.4 Calculation of K_{Ga} for SO_2 -Air-Lime system in Rotating Bed:

SO_2 being removed from the gas phase can be expressed in terms of K_{Ga} in equation (52).

$$N = G (y_i - y_o) = K_{Ga} A Z P_T (y - y^*)_{LM} \quad (52)$$

where

$$(y - y^*)_{LM} = \frac{(y - y^*)_i - (y - y^*)_o}{\ln \frac{(y - y^*)_i}{(y - y^*)_o}} \quad (53)$$

$Ca(OH)_2$ reacts rapidly with SO_2 absorbed in the liquid phase. Therefore, the vapor pressure above the liquid phase is eliminated or substantially reduced, causing Henry's Law constant, H , to approach to zero. In this case, mass transfer driving force, $(p - p^*)$, is maximized. Assuming y^* to be zero at the inlet and outlet of the rotor, equation (52) is reduced to the following term.

$$G (y_i - y_o) = K_{Ga} A Z P_T \frac{(y_i - y_o)}{\ln (y_i / y_o)} \quad (54)$$

Packing depth, Z , is:

$$Z = \frac{G}{K_{Ga} A P_T} \ln \frac{y_i}{y_o} \quad (55)$$

where

$$HTU = \frac{G}{K_{Ga} A P_T} = \frac{G'}{K_{Ga} P_T} \quad (56) \quad \text{and} \quad NTU = \ln \frac{y_i}{y_o} \quad (57)$$

Here, N is transfer rate of SO_2 from the gas phase to liquid phase (lb-mol/h), G

is molar gas flow rate (lb-mol/h), G' is the average molar gas flow rate based on inner and outer cross-sectional area of rotating bed (lb-mol/ft² · h), y_i and y_o are the SO₂ concentrations at the inlet and outlet gas, respectively.

By measuring the inlet and outlet SO₂ concentrations in the gas phase for a known packing depth, HTU and then K_{Ga} can be calculated according to the equation below.

$$HTU = \frac{Z}{NTU} = \frac{G'}{K_{Ga} P_T} \quad (58)$$

In equation (58) K_{Ga} is the overall gas-phase volumetric mass-transfer coefficient (lb-mol/ft² · h · atm), a is the interfacial area per unit volume of packing (ft²/ft³).

For SO₂-Air-Lime system in which H is very large, overall and individual mass-transfer coefficients K_{Ga} and k_{Ga} are equal to each other according to equation (13).

4. 5. The Effect of High Gravitational Field on Absorption with compared to Conventional Packed Tower:

The packed towers which are usually filled with randomly oriented packing material, are extensively used for absorption operations to provide a significant gas-liquid contact area. The liquid phase is introduced over the top of the packed bed. The liquid trickles down and wets the packing which has a large surface area. The solute-containing gas phase enters the column below

the packing at the bottom of the tower and flows upward through the packing elements countercurrent to the liquid flow and exits at the top of the tower. While liquid flows downward under the influence of gravity, the gas which flows upward must overcome the resistance caused by the tower packing elements.

The rate of mass transfer from the liquid to the gas phase can be increased by increasing the gas and liquid flow rates through the tower. The increased flow rates cause convection and turbulent mixing to be increased in both gas and liquid phases. However, an upper limit called flooding region exists for the gas flow rates in which the entrainment of liquid by gas is begone. Most packed towers is operated below the flooding region, resulting in a limitation in the mass transfer rate. In a packed tower relation for pressure drop and flooding is given by Sherwood correlation shown in Figure 6. In equations (24), (37) and (56), HTU is a function of gas and liquid flow rates, the characteristics of packing, gas and liquid-phase mass-transfer coefficients. NTU depends on concentration of solute in the inlet and outlet and phase equilibria properties of solute in equations (25), (38), (57)

When absorption is carried out in the high gravitational field by using rotating packed bed, higher gas and liquid flow rates, and higher surface area of packing can be used without flooding of the bed. Assuming that Sherwood equation is conservative in the rotating bed, this effect can be observed since gravitational constant, g , is included in the denominator.

Volumetric mass transfer coefficients are enhanced, resulting from the

$$\frac{L}{G} \left(\frac{\rho_g}{\rho_L}\right)^{0.5}$$

Figure 6. Sherwood Correlation

use of high gas flow rate and high surface area. Other contributing factors to the higher coefficients are an increase in the effective gas-liquid contacting efficiency, a_e/a , and an increase in the liquid-phase mass-transfer coefficient, k_L . Increase in overall mass-transfer coefficients and interfacial area of packing reduces HTU. As a result, a significant reduction in absorber size can be achieved.

5. EXPERIMENTAL RESULTS AND DISCUSSIONS:

The overall volumetric mass-transfer coefficients are determined in rotating packed bed for SO₂ absorption in water and lime slurry. The effect of acceleration field, the liquid and gas flowrates on the overall volumetric mass-transfer coefficients were investigated for both system. Runs were repeated in the same conditions to determine experimental error and reproducibility. Experimental results are summarized in Appendix.

5.A. SO₂-Air-Water System:

5.A.1. Effect of Gravity:

Experimental runs were made in constant gas flowrates of 20 and 40 acfm and liquid flowrates of 2, 3, 4, 5 gpm. SO₂ inlet concentration in the gas mixture is kept constant at 1000 ppm. The rotor speed was set at 250, 700, 850, 1000, 1300, 1500 rpm to achieve a centrifugal acceleration of 12.4, 97.3, 143.5, 198.7, 335.8, and 447 g, respectively (arithmetical average of gravity based on the inner and outer radius of the rotor). Figure 7 depicts the influence of gravity on calculated K_{Ga} for different liquid flowrates. Overall volumetric mass-transfer coefficients increase logarithmically with the gravity. As shown in Figure 8, at gas flowrate of 40 acfm and liquid flowrate of 3 gpm, linear regression analysis of the logarithm of the K_{Ga} vs #g data yields $K_{Ga} \propto G^{0.37-0.40}$ in agreement with Dudukovic and Vivian but regression coefficient for the exponent is poor (0.86-0.92) indicating considerable scatter. It should be noted that agreement with Dudokovic's result is only valid for liquid

SO₂-Air-Water
G=20 acfm, y₁=1000 ppm

Figure 7

SO₂-Air-Water System
 $G=40\text{acfm}$ $y_i=1000\text{ppm}$

Figure 8

flowrate of 3 gpm.

5.A.2. Effect of Liquid Flowrate:

At constant gas flowrate of 40 acfm and 20 acfm, the influence of liquid flowrates of 1 - 5 gpm is investigated as shown in Figure 9 and Figure 10. The slopes of the best line drawn through these data vary considerably with liquid flowrates. Slopes vary between 0.4-0.10 at $G=40$ acfm, and 0.14-0.053 at $G=20$ acfm.

5.A.3. Effect of Gas Flowrate:

At constant liquid flowrates, gas flowrates are changed from 20 acfm to 40 acfm. For higher gas flowrates, higher volumetric mass-transfer coefficients are obtained. The influence of gravity on K_Ga at lower gas flowrates was smaller. The same trend are observed for different liquid flowrates. Figure 11 shows the effect of gas flowrate on the influence of gravity for constant liquid flowrate of 5 gpm.

5.A.4. Discussion:

The observation of lowering gravity effect for higher liquid flowrates may be explained as follows. For liquid-phase mass-transfer controlling systems, the improvement of volumetric mass-transfer coefficient is achieved as liquid film gets thinner in high gravity. The chance of having thin liquid film on the surface of packing is reduced at higher liquid flowrates. Also higher gas flowrates help liquid be dispersed better and consequently reduce liquid film thickness. It should be noted that SO_2 absorption in water is liquid

SO₂-Air-Water System
 $G=40\text{acfm}$ $y_1=1000\text{ppm}$

Figure 9

SO₂-Air-Water System
 $G=20 \text{ acfm}$ $y_i=1000 \text{ ppm}$

Figure 10

SO₂-Air-Water System
 L=5 gpm y_i=1000ppm

Figure 11

and gas-phase mass-transfer limiting, therefore calculated K_{Ga} contains liquid and gas-phase effects together.

5.B. SO_2 -Air-Lime System:

Tests were also performed for SO_2 absorption in lime slurries. Hydrated lime containing 93 % $Ca(OH)_2$ and 2 % $Mg(OH)_2$ is used. Stoichiometric ratio of 1.3 is used to provide 30 % excess $Ca(OH)_2$ in slurry. In this conditions gas-phase controls mass transfer. Overall volumetric mass-transfer coefficient, K_{Ga} , is equal to individual gas-phase mass-transfer coefficient, k_{Ga} .

5.B.1 Effect of Gravity:

Liquid and gas flowrates were held constant, while the rotor speed was set at 250, 400, 500, 700, 800, 1000, 1300, and 1500 rpm to achieve a centrifugal acceleration of 12.4, 31.7, 49.6, 97.3, 143.5, 198.7, 335.8 and 447 g, respectively (g is calculated as the average centrifugal acceleration based on the inner and outer radius of the rotor). Figure 12 and 13 show the effect of gravity on K_{Ga} at $L=2$ gpm and $G=40$ acfm: for two runs performed in the same conditions. K_{Ga} values is observed to be logarithmically increasing with increasing numbers of gravity. Linear regression analysis of the logarithm of K_{Ga} vs # g yields as $K_{Ga} \propto g^{0.083-0.088}$ in Figure 14 and 15.

5.B.2 Effect of Liquid Flowrate:

The K_{Ga} dependency on acceleration fields was determined at liquid

SO₂-Air-Lime**G=40 acfm, L=2 gpm, y₁=1000 ppm**

Figure 12

SO₂-Air-Lime**G=40 acfm, L=2 gpm, y₁=1000 ppm**

Figure 13

SO₂-Air-Lime System
G=40acfm L=2gpm y_i=1000ppm

Figure 14

SO₂-Air-Lime System
G=40acfm L=2gpm y_i=1000ppm

Figure 15

flowrates of 1 and 2 gpm, while gas flowrates are held constant at 55 acfm. The slopes of the best line drawn through data, using least square technique are 0.115-0.09 for 1 gpm and 0.078-0.073 for 2 gpm in Figure 16. Figure 17 depicts the effect of liquid flowrate at $G=40$ acfm, giving slopes of 0.105-0.088. It turns out to be that the liquid flowrates does not affect the power dependency of K_Ga on gravitational field. It can be recognized that differences between slopes are within the range of experimental error.

5.B.3. Effect of Gas Flowrate:

Liquid flow rate of 1 gpm is kept constant, and gas flowrates of 40 and 55 acfm are used. As Figure 18 illustrates, for higher gas flowrates, higher K_Ga values are obtained but slopes do not change (0.115-0.105) within the range of experimental error. It can be said that gas flowrates does not affect the power dependency of K_Ga on gravitational field.

5.B.4. Discussion:

SO_2 absorption in lime slurry is an example for the absorption accompanied by chemical reaction. Since SO_2 molecules reacts on the gas-liquid interface, liquid film distance through which SO_2 passes are greatly reduced. Therefore, the effect of gravity in terms of reducing the liquid film thickness is less pronounced for gas-phase mass-transfer limiting systems.

SO₂-Air-Lime System
 $G=55\text{acfm}$ $y_1=1000\text{ppm}$

Figure 16

SO₂-Air-Lime System
 $G=40\text{acfm}$ $y_i=1000\text{ppm}$

Figure 17

SO₂-Air-Lime System
L=1 gpm $y_1=1000\text{ppm}$

Figure 18

6. CONCLUSION:

For SO_2 -Air-Water system, the overall volumetric mass-transfer coefficient, K_{Ga} , increases logarithmically with average gravity to a power between 0.40-0.15. Liquid flowrate has great influence on the gravity effect. The effect of gravity increases as the liquid flowrate decreases. In higher gas flowrates higher K_{Ga} values are obtained and gravity effect is increased.

In SO_2 -Air-Lime system volumetric mass-transfer coefficient, K_{Ga} , is found to increase logarithmically as $K_{Ga} \propto g^{0.08-0.11}$. Gravity effect on K_{Ga} is less pronounced for gas-phase mass-transfer limiting system. The effect of gravity on K_{Ga} is not found to be influenced by gas and liquid flowrates.

For both systems, the overall volumetric mass-transfer coefficients obtained in rotating bed are much higher than the corresponding values in conventional packed towers.

7. NOMENCLATURE:

a_e	effective gas-liquid interfacial area (ft^2/ft^3)
a	total packing surface area (ft^2/ft^3)
A	packing cross-sectional area (ft^2)
A_i	inner cross-sectional area of rotating bed (ft^2)
A_o	outer cross-sectional area of rotating bed (ft^2)
c	SO_2 concentration in liquid ($\text{lb-mol}/\text{ft}^3$)
d	diameter (or characteristic dimension) of packing (ft)
d	width of packed bed (ft)
D	liquid-phase diffusion coefficient (ft^2/h)
g	gravitational acceleration (ft/sec^2)
G	molar gas flowrate ($\text{lb-mol}/\text{h}$)
G'	average of molar gas flowrates per unit area based on inner and outer cross-sectional area of rotating bed ($\text{lb-mol}/\text{ft}^2 \cdot \text{h}$)
G_i	molar gas flowrate per unit area based on inner cross-sectional area of rotating bed ($\text{lb-mol}/\text{ft}^2 \cdot \text{h}$)
G_o	molar gas flowrate per unit area based on outer cross-sectional area of rotating bed ($\text{lb-mol}/\text{ft}^2 \cdot \text{h}$)
h	height of short wetted-wall column (ft)
H	Henry's law constant ($\text{atm}/(\text{lb-mol}/\text{ft}^3)$)
H'	Henry's law constant ($(\text{lb-mol}/\text{ft}^3)/\text{atm}$)

HTU	height of a transfer unit (ft)
k_G	gas-phase mass-transfer coefficient (lb-mol/ h·ft ² · atm)
k_L	liquid-phase mass-transfer coefficient (lb-mol/h·ft ² · (lb-mol/ft ³)) or (ft/h)
K_{Ga}	overall volumetric gas-phase mass-transfer coefficient (lb-mol/h·ft ³ · atm)
K_{La}	overall volumetric liquid-phase mass-transfer coefficient (1/h)
L	molar liquid flow rate (lb-mol/h)
L'	average of molar liquid flowrates per unit area based on the inner and outer cross-sectional area of rotating bed (lb-mol/ft ² · h)
L_i	molar liquid flowrate per unit area based on the inner cross-sectional area of rotating bed (lb-mol/ft ² · h)
L_o	molar liquid flowrate per unit area based on the outer cross-sectional area of rotating bed (lb-mol/ft ² · h)
m	slope of equilibrium curve
N	solute transferred (lb-mol/h)
NTU	number of transfer units
p	partial pressure of solute (atm)
p'	partial pressure of solute at gas-liquid interface (atm)
p^*	partial pressure of solute which is in equilibrium with liquid (atm)
P_T	total pressure (atm)
r_i	rotor inside radius (ft)

r_o	rotor outside radius (ft)
x	SO ₂ mol fraction in liquid-phase
x'	mol fraction of solute in the liquid-phase at interface
x^*	SO ₂ mol fraction in the liquid-phase in equilibrium with the gas-phase
x_i	SO ₂ mol fraction in the rotor inlet
x_o	SO ₂ mol fraction in the rotor outlet
y	SO ₂ mol fraction in gas-phase
y'	mol fraction of solute in the gas-phase at interface
y^*	SO ₂ mol fraction in the gas-phase in equilibrium with the liquid-phase
y_i	SO ₂ mol fraction in the rotor inlet
y_o	SO ₂ mol fraction in the rotor outlet
Z	depth of packing (ft)

Greek Letters:

ϵ	packing void fraction
Γ	liquid flowrate per unit perimeter in wetted-wall column (lb/h · ft)
μ	liquid viscosity (lb/h · ft)
ρ	liquid density (lb-mol/ft ³) or (lb/ft ³)

Dimensionless Groups:

Gr Grashof number = $g d^3 \rho^2 / \mu^2$

Reynolds number = $d L / \mu$

Schmidt number = $\mu / \rho D$

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9. APPENDIX:

TABLE 1

SO₂-Air-Water G=40 acfm* y_i=1000 ppm*

# g	L=3 gpm*	L=3	L=4	L=4	L=5	L=5
12.41	144.15	160.17	111.86	114.11	106.58	99.79
97.36	210	295	144.77	150.86	134.42	118.71
143.56	530.43	384.41	-	167.81	146.29	-
198.70	-	556.32	-	182.12	157.4	-
335.8	-	-	179.27	210.11	174.51	139.79
447	-	-	-	222.82	186.37	-

TABLE 2

SO₂-Air-Water G=20 acfm y_i= 1000 ppm

# g	L=2	L=3	L=4	L=5
12.41	77.3	58.43	56.83	58.83
97.36	111.54	71.23	65.65	66.24
143.56	114.92	75.32	68.23	68.25
198.7	120.34	76.64	69.02	69.53
335.8	125.22	77.54	71.49	69.97
447	128.33	78.95	72.36	70.87

TABLE 3

SO₂ -Air-Lime G=40 acfm y_i=1000 ppm

#g	L=1	L=1	L=2	L=2
12.4	65.87	67.57	67.04	66.59
31.79	71.22	71.22	-	73.62
49.67	79.49	79.19	77.19	76.39
97.36	87.05	86.33	79.29	82.27
143.56	90.87	90.87	82.16	85.41
198.7	92.27	92.10	84.92	87.27
335.8	92.52	92.52	88.67	89.64
447	93.80	92.94	91.11	90.88

TABLE 4

SO₂ -Air-Lime G=55 acfm y_i=1000 ppm

#g	L=1	L=1	L=2	L=2
12.41	82.41	89.06	-	69.29
27.94	-	-	-	74.84
31.79	94.02	92.11	-	-
49.67	95.67	98.08	72.74	-
97.36	111.2	105.63	-	-
143.56	115.44	109.86	-	81.24
198.7	118.48	113.99	78.86	85.55
263.69	119.53	115.94	-	-
335.8	121.14	117.45	-	-
447	121.7	118.48	86.88	93.22

* acfm: actual cubic feet per minute

ppm: part per million by volume

gpm: gallon per minute

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