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DOE/ET/14801-30
(DE83001643)

GAS/SLURRY FLOW IN COAL LIQUEFACTION PROCESSES
(FLUID DYNAMICS IN A THREE-PHASE FLOW COLUMN)

Final Technical Progress Report for October 1, 1979—March 31, 1982

April 1982

Work Performed Under Contract No. AC22-79ET14801

Air Products and Chemicals, Inc.
Allentown, Pennsylvania

U. S. DEPARTMENT OF ENERGY

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(FLUID DYNAMICS IN A THREE-PHASE FLOW COLUMN)

Final Technical Progress Report
For Period 1 October 1979-31 March 1982

David H. S. Ying
R. Sivasubramanian
Samir F. Moujaes
E. N. Givens

Air Products and Chemicals, Inc.
Allentown, PA 18015

April 1982

Prepared for the United States Department of Energy

Under Contract No. DE-AC22-79ET14801

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ACKNOWLEDGEMENTS

The authors would like to acknowledge the excellent work by Dale Schoeneberger and Lew Fendrick in the construction, maintenance and operation of the equipment used in this project. In addition, we would like to thank Bill Collins and Jeffrey Rabe for their contribution in operating the column and to thank Debbie Wilson, Lori Bedics and Carol Spiess. A special thanks is due to Rebecca Hahn who contributed a lot of time in putting together this final package.

Summary

This is the final report under Contract Number DE-AC22-79ET14801 titled "Gas/Slurry Flow in Coal Liquefaction Processes", covering the period 1 October 1979 to 31 March 1982. This work is a continuation of studies initiated by Air Products and Chemicals, Inc. on the fluid dynamics of 3-phase flow to support the design of a 6000 T/D commercial dissolver. Design studies on the SRC-I 6000 T/D commercial plant began in July, 1978. DOE support for the 3-phase flow studies by Air Products began 1 July 1979 under contract number DE-AC05-78-OR0-3054 and terminated 30 September 1979 at the start of this contract. A background of information developed at Air Products is included in the text.

A commercial coal liquefaction plant will employ vertical tubular reactors feeding slurry and gas concurrently upward through these vessels. In the SRC-I plant design the reactor is essentially an empty vessel with only a distributor plate located near the inlet. Because the commercial plant represents a considerable scale-up over Wilsonville or any pilot plant, this program addressed the need for additional data on behavior of three phase systems in large vessels. Parameters that were investigated in this program were studied at conditions that relate directly to projected plant operating conditions.

The design goal for commercial dissolvers is to ensure that solids passing through the reactors remain suspended in sufficient quantity and distribution to provide the necessary catalytic effect. To achieve this goal, the fluid dynamic behavior of a three-phase upflow system was studied by measuring gas and slurry holdup, liquid dispersion, solids suspension and solids accumulation. The dependent parameters are gas and liquid velocities, solid particle size, solids concentration, liquid viscosity, liquid surface tension and inlet distributor. Two reactor columns (5-inch diameter by 5 feet and 12-inch diameter by 25 feet) were used in this cold-flow simulation which included air/water/sand, nitrogen/tetralin/sand, and air/glycol-water/sand systems. In addition, a model was developed to describe the distribution and accumulation of solid particles in the reactor columns.

Gas holdup in gas/liquid and gas/liquid/solid systems were investigated both with and without liquid flow. Within the range of liquid superficial velocity from 0.0-0.5 ft/sec, gas holdup is found to be independent of liquid flow which agrees with other investigators. The results also confirm our previous finding that gas holdup is independent of column diameter when the column diameter is 5 inches or larger. This independence of column diameter has been tested with a wide range of operating conditions: 0.0-0.6 ft/sec liquid velocity, 0.0-0.4 ft/sec gas velocity, 140 minus - 20 mesh particles, 0.0-20.0 lb/ft³ solids concentration, with and without distributor.

The gas holdup depends strongly on gas flow rate; gas holdup increases with increasing gas velocity. In general, gas holdup is found proportional to $(V_g)^{0.75}$, where V_g is gas superficial velocity. A lower exponential value (0.50-0.65) was measured when either 50% glycol-50% water and 70% glycol-30% water solutions were used.

Effect of solids particles on gas holdup depends on the gas flow rate. At low gas superficial velocities (≤ 0.10 ft/sec), the presence of solids did not change the gas holdup. However, at high gas flow, gas holdup was reduced slightly (3-10%) with both fine and large particles. This effect is true regardless of the type of inlet distributor.

Increasing liquid viscosity and surface tension reduce gas holdup which agrees with other investigators. Because of the complexity of the system, we could not find a single correlation to best fit all the data. However, the correlations developed by Akita and Yoshida, Pilhofer et al., and Hughmark all fit the data with a reasonable degree of accuracy. It is important to note that none of these correlations can be applied to a foaming system.

The degree of liquid backmixing markedly affects chemical changes occurring in the dissolver, such as sulfur removal, and oil and distillate formation. The liquid dispersion coefficient, which is a measure of the degree of backmixing, was found to increase with increasing gas velocity and column diameter. Within the range of liquid physical properties employed in this work, no trend

of dependence on viscosity and surface tension was observed. In a gas/liquid two phase upflow system, a correlation developed by Baird et al., best fit our data. Several other correlations, including one developed by Air Products, underpredict our results slightly.

The liquid dispersion coefficient was found to be independent of liquid velocity and inlet distribution. However, the presence of solid particles reduced the coefficient which decreased with increasing solids concentration, regardless of the particle size. Nevertheless, the liquid dispersion results have clearly shown that the SRC-I Demonstration Plant Dissolver closely resembles a completely backmixed reactor.

The amount of hydrogen gas transferred to the bulk liquid is an important factor in the dissolver design. Chemical reactions, such as solvent rehydrogenation, will be limited by the gas/liquid mass transfer rate if the hydrogen transfer is less than the rate consumed by the reactions.

Gas/liquid mass transfer rates were investigated in both air/water and air/water/sand systems using a 5-inch and 12-inch diameter column. Our results indicate that the mass transfer rate increases with increasing gas velocity. However, its dependence on gas velocity behaves differently as the column diameter varies. In both the 5-inch and 12-inch diameter columns, the mass transfer coefficient increases with increasing gas velocity, but the rate of increase in the larger vessel decreases with increasing gas velocity whereas in the 5-inch diameter column, the rate of increase is almost constant.

The gas/liquid mass transfer coefficient is independent of column height and inlet distribution because of the highly backmixed system. However, the presence of solids reduces the transfer rate. A correlation developed by Yoshida and Akita for a gas/liquid two-phase system was found to fit our air/water data. This correlation was modified to describe the mass transfer rate in a gas/liquid/solid system.

Gas velocity is the key element for maintaining solid particles in suspension. Above the critical gas velocity, which is defined as the minimum gas velocity to achieve full solids suspension, solids accumulation is practically independent of gas velocity. However, solids accumulation was found to increase linearly

with decreasing liquid velocity at a fixed gas input rate. The effect of liquid and gas velocities on solids accumulation was studied in a 5-inch and 12-inch diameter column using 60/80 and 140 minus mesh sand particles. The superficial gas and liquid velocities were varied from 0.01 to 0.43 ft/sec and from 0.0 to 0.05 ft/sec, respectively, covering the flow conditions designed for the SRC-I demonstration plant. At a half slurry turndown velocity, the accumulation of the fines (140 mesh minus) did not increase substantially but the amount of large particles (60/80 mesh) increased tremendously. Of more concern is the settling of the larger particles which have critical gas velocities higher than the half gas turndown velocity.

Increasing column diameter and liquid viscosity reduces the accumulation of solids in the dissolver. As the column diameter increases, dispersion in the system also increases, thereby reducing the solids concentration gradient and accumulation. Since particle terminal velocity decreases with increasing liquid viscosity, the solid particles can be carried through the column by the bulk liquid flow easier in a more viscous liquid. Therefore the accumulation is reduced. These effects of column diameter and liquid viscosity are more apparent on the large particles because the distribution of fine particles is nearly homogeneous whereas the large ones show a substantial concentration gradient.

A solid dispersion model was developed to predict the solids concentration gradient and accumulation in a gas/liquid/solid upflow system. This model was tested successfully with 40 runs conducted in water, tetralin, and 50% glycol-50% water solution under a wide range of operating conditions covering the SRC-I plant design. We also found that there is little particle-particle interaction to affect the solids distribution profiles of different particle sizes. Therefore the model-predicted concentration profile of each individual particle size can be superimposed together to provide a complete description of the solids distribution in the dissolver.

When the critical gas velocity of any particle exceeds the operating flow conditions, the solid particles will settle at the reactor bottom, creating an undesirable situation. Our results indicate that a solid withdrawal system

located at the reactor bottom can reduce solids accumulation and significantly alter the solids distribution profile. Although all solid particles, large or small, were removed together at the bottom of the column, our data indicate that the larger particles were removed faster because they settled to the bottom more quickly. However, these results were obtained under non-steady-state operation. It is recommended to perform steady-state solids removal experiments in the future to determine the optimal operating conditions.

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1.0 OBJECTIVE

The overall objective of this project was to study solids accumulation and suspension of various gas/liquid/solid systems in cold-flow tubular columns to provide data to design the coal dissolver for the SRC-I Demonstration Plant.

The specific objectives of this program were:

1. to study the effects of slurry velocity, gas velocity, solid particle size, solids concentration, liquid viscosity, and surface tension on the performance of a cold-flow tubular column;
2. to develop an effective technique to withdraw slurry from the bottom of a tubular column as a means to control solid concentration;
3. to study the performance of cold-flow tubular columns both with an improved distributor and without a distributor.

A major part of the coal-dissolution section of any liquefaction plant is the dissolver. Although considerable liquefaction occurs in the preheater, most of the necessary chemical changes occur in the dissolver, including sulfur removal, oil and distillate formation, and solvent rehydrogenation.

Vertical tubular reactors are used in all of the major processes currently under consideration for commercial coal liquefaction. In the Solvent Refined Coal (SRC), Exxon Donor Solvent (EDS), and H-Coal processes, slurry and gas are concurrently fed upward through these vessels. In the EDS and SRC processes, the reactors are basically empty vessels, whereas in the H-Coal process a bed of ebullating catalyst is maintained in the reactor. Between the EDS, SRC-I, and SRC-II processes, the major differences in dissolver operation are the compositions of the feed streams and reactants within the dissolver. Different hardware such as the distributor plates, draft tubes, or recycle loops can also affect the behavior of slurries in these vessels.

In order to design a technically feasible and cost effective dissolver, the physical behavior of three-phase systems in tubular columns must be clarified. All of the major processes under development require an understanding of backmixed three-phase systems. In each process, part of the dissolver volume is backmixed. As the design of the 6,000-ton-per-day (tpd) SRC-I Demonstration Plant progresses, the increased vessel size (and other considerations) may dictate the use of reactors in series, which would decrease the overall backmixed characteristic of the commercial plant.

The SRC-I Demonstration Plant dissolver will be scaled up considerably from the Wilsonville and Ft. Lewis Pilot Plant dissolvers. The relative sizes of the dissolvers for the pilot plants and the SRC-I Demonstration Plant are compared below (at a residence time of 0.56 hr, a gas-feed rate of 20,000 mscf per ton of dried coal, and a 38% coal slurry):

<u>Plant</u>	<u>Size</u>	<u>Volume</u>	<u>Dissolver</u>	<u>Height</u>	<u>Superficial Velocities (ft/sec)</u>	
	<u>(tpd)</u>	<u>(ft³)</u>			<u>Liquid</u>	<u>Gas</u>
Wilsonville	6-10	18.1	12 in.	23	0.012	0.074
Ft. Lewis	50-61	106.8	24 in.	34	0.017	0.10
SRC-I Demo Plant	6,000	10,454.0	11 ft	110	0.06	0.36

The reactor volumes and diameters differ dramatically. Thus, to make intelligently good decisions about the designs, we need more information on the flow properties of three phase systems in large vessels.

Also important is the effect of differences in gas and liquid superficial velocities on slurry behavior. A fivefold difference in velocity exists between Wilsonville and the commercial SRC-I design. This difference can have a considerable impact on the process because the gas and liquid superficial velocities have a strong effect on (a) gas void volume, (b) actual solids concentration in the dissolver, and (c) the relative degree of backmixing. As velocity through the dissolver increases, the tendency for solids to remain behind diminishes, causing a decrease in the actual concentration of ash particles in the reactor. Those particles that do remain will tend to be larger. Considerable evidence shows that reactor solids have a definite catalytic effect. Larger particles will have fewer surface areas exposed, and thus will probably have diminished catalytic activity. Knowing the particle sizes that can accumulate under commercial flow conditions will give us some indication of the size of dissolver solids that should be examined for catalytic activity.

3.0 BACKGROUND

3.1 Gas/Liquid Flow

Numerous investigators (1-28) have studied the behavior of the gas and liquid components in two-phase flow. With the fluid as the continuous phase, the gas-flow pattern can be categorized as bubbly, slug, or churn-turbulent. The pattern is determined primarily by the competing rates of bubble coalescence and breakup, which, in turn, depend on liquid-phase properties such as viscosity, surface tension and density. During bubble coalescence, as the bubbles draw together, the thin film of liquid separating the bubbles ruptures. Apparently, gas bubbles coalesce by the capture of one bubble in the wake of another rising bubble; this has been extensively studied by Crabtree and Bridgwater (18). Calderbank et al. (24) reported that increased liquid viscosity enhances the rate of bubble coalescence. Bubble breakup is due to disturbance at the interface. In the early fifties, Davies and Taylor (17) performed a theoretical study on the instability of liquid surfaces. Clift and Grace (23) proposed in 1972 that bubble breakup is due to this instability.

One of the most important aspects of gas flowing through a tubular reactor is its availability at the reaction site. The transfer of gas to that site, be it in a homogeneous liquid-phase transition state or a heterogeneous surface-activated state, is ultimately related to the interfacial area between the liquid and gaseous phases. In a two-phase gas/liquid system, the interfacial area is related to the volume fraction actually occupied by gas. This volume fraction, which is typically referred to as the gas holdup or void fraction, is a value that is extremely important to optimum reactor design.

In a two-phase flow reactor, gas residence time is governed by the gas holdup at the operating conditions and mass transfer is governed by gas contact at the interfacial phase barrier. Numerous investigators (8,10,11,13,14) have studied gas holdup with many different systems, covering a wide range of liquid viscosity (0.58-152.0 cp), surface

tension (22.3-76.0 dynes/cm), and density (0.79-1.70 gm/cm³). Several correlations have been developed to predict the gas-void fraction in two-phase flow systems. In general, the gas-void fraction is expressed in terms of gas superficial velocity, liquid surface tension, viscosity, and density. A partial list of the existing correlations is provided in Table 1.

Generally, gas-void fraction does not depend on liquid input rate. Gas holdup is usually higher for a smaller column diameter because of the wall proximity effect. Ellis and Jones (25) studied the effect of column diameter on gas holdup and concluded that, for diameters greater than 3 in., gas holdup is independent of column diameter. Directionally, work at Air Products agrees with their results.

Mass transfer in bubble columns has been studied extensively (1,7,9,12, 13,16,19,21,22,26). Because the mass-transfer resistance in the gas phase is negligible compared with that in the liquid phase, the overall mass-transfer coefficient is dominated by that of the liquid phase. In general, the volumetric liquid-phase mass-transfer coefficient ($k_L a$) depends on gas-input velocity, column diameter, bubble diameter, and liquid properties (diffusivity, surface tension, viscosity, and density). Several correlations have been developed (7,12,16,19-22,24). The dependence of mass transfer on liquid velocity and gas properties (density and viscosity) has also been proposed (7,19,20) but this is not unanimously accepted. A higher gas-input rate increases mass transfer because it enhances interfacial area and liquid turbulence.

Another significant aspect of gas/liquid flow is the dispersion (back-mixing) of the fluid phase. In studies at Air Products, we found that in a gas/liquid flow column an extremely small amount of gas flow can trigger a high degree of dispersion in the liquid phase. Shah, et al., (15) have extensively reviewed backmixing in gas/liquid reactors. In general, backmixing in the liquid phase depends on column diameter, gas velocity, and the nature of the gas distributor plate. Most investigators agree that the axial dispersion coefficient is independent of liquid-flow rate and liquid-phase properties (surface

Table 1

Correlations for Predicting Gas Holdup

- Hughmark (16)

$$\epsilon_g = \frac{V_{sG}}{U_s} \left[\left(\frac{62.4}{\rho_L} \right) \left(\frac{72}{\sigma} \right) \right]^{1/3}$$

- Akita and Yoshida (26)

$$\frac{\epsilon_g}{(1-\epsilon_g)^4} = C_1 (N_{Bo})^{1/8} (N_{Ga})^{1/12} (N_{Fr})$$

- Hikita and Kikukawa (3)

$$\epsilon_g = 0.505 V_{sG}^{0.47} (72/\sigma)^{2/3} (1/\mu)^{0.05}$$

where ϵ_g = gas-void fraction

V_{sG} = gas superficial velocity (ft/sec)

U_s = slip velocity (ft/sec)

ρ_L = liquid density (lb/ft³)

σ = liquid surface tension (dynes/cm)

$N_{Bo} = gD^2\rho_L/\sigma$ (Bond Number)

$N_{Ga} = gD^3/\nu_L^2$ (Galileo Number)

$N_{Fr} = V_{sG}/\sqrt{gD}$ (Froude Number)

C_1 = 0.20 for nonpolar solutions, 0.25 for polar solutions

D = column diameter (ft)

ν_L = liquid kinematic viscosity (ft²/sec)

g = gravitational acceleration (32.2 ft/sec²)

μ = liquid viscosity (Cp)

tension, viscosity, and density). Many correlations have been developed to predict the axial dispersion coefficient in gas/liquid flow reactors. Table 2 lists some correlations for gas as the dispersion phase and liquid as the continuous phase.

At low gas velocities, the agreement among these published correlations is poor. For example, for a 5-in.-diameter column, we estimated a sixfold difference in the predicted dispersion coefficient at a 0.02 ft/sec superficial gas velocity:

Axial Dispersion Coefficients (ft²/sec) from Published Correlations

<u>Reference</u>	<u>Low Gas Velocity</u> <u>(0.02 ft/sec)</u>	<u>High Gas Velocity</u> <u>(.33 ft/sec)</u>
28	0.047	0.190
1	0.088	0.204
2	0.022	0.053
3	0.134	0.219
4	0.129	0.229
6	0.091	0.239

However, at high gas velocities these correlations give better agreement. Because of the inconsistencies in these published correlations, guidelines are unavailable for predicting the dispersion coefficient, which is an essential value for designing a reactor.

To anyone initiating studies in this area, experiments in two-phase flow should be performed to make certain that experimental techniques are adequate to reproduce existing data. Air Products has performed such studies and has developed standard techniques that do reproduce existing data. From these studies, a keen understanding of the value of the correlative models has been gained, which will be applied in the proposed program.

Table 2

Correlations for Predicting Axial Dispersion Coefficients

Kato and Nishiwaki(4)

$$E_{ZL} = \frac{V_G D (1 + 6.5 Fr_G^{0.8})}{13 Fr_G}$$

Towel and Ackerman(28)

$$E_{ZL} = 1.23 D^{1.5} V_G^{0.5}$$

Cova(2)

$$E_{ZL} = 0.0759 V_G^{0.32} \rho_L^{0.07}$$

Deckwer, et al(1)

$$E_{ZL} = 0.678 D^{1.4} V_G^{0.3}$$

Hikita and Kikukawa(3)

$$E_{ZL} = (0.366 + 0.674 V_G^{0.77}) D^{1.25} (1/\mu_L)^{0.12}$$

Baird and Rice(6)

$$E_{ZL} = 0.35 D^{4/3} (V_G g)^{1/3}$$

Ying(49)

$$E_{ZL} = 0.27 D V_G \left(\frac{Dg}{V_G^2} \right)^{0.32}$$

where $Fr_G = \text{Froude Number} = \sqrt{V_G^2/gD}$

E_{ZL} = axial dispersion coefficient (ft²/sec)

V_G = gas superficial velocity (ft/sec)

D = column diameter (ft)

ρ_L = liquid density (g/cm³)

μ_L = viscosity of liquid (cP)

g = acceleration of gravity (ft/sec²)

3.2 Gas/Liquid/Solid Flow

Adding solid particles to a gas/liquid flow reactor greatly increases the complexity of the system. The presence of the solid itself, as well as the particle size, size distribution, and density, can affect the behavior significantly. The rate of bubble coalescence increases in the presence of solid particles. Kato et al. (29, 30) observed larger coalesced bubbles in a gas/liquid/solid system. However, this enhancement in the rate of bubble coalescence decreases to an insignificant level in a turbulence regime at high gas velocities.

Studies of gas holdup in a three-phase flow reactor (Air Products work) showed that the gas-void fraction decreases in the presence of solid particles at low gas velocities. At high gas velocities, the dependence of gas holdup on solid particles disappears. These effects of solid particles on gas holdup are directly in line with the observation of the bubble coalescence rate in the presence of solid particles.

A unique feature of three-phase flow is the contraction of the fluidized bed (34, 35, 37, 39-42) upon addition of gas to the system, as first reported by Turner (34, 35). Empirical expressions have been developed and an iterative method has been employed to determine the volume fraction of the phases (37, 39, 41).

Several investigators have also studied the effects of solid particles on liquid backmixing in three-phase flow systems (31, 38). Some data of (31) indicate that the presence of solid particles reduces liquid backmixing.

The effects of particle size, gas velocity, and liquid velocity are interrelated. An empirical correlation was developed by Kato et al. (29, 30) to predict the dispersion coefficient of liquid in a slurry. The correlation does not depend on particle size or liquid velocity.

$$E_{ZSL} = \frac{V_{SG} D (1 + 8 Fr_G^{0.85})}{13 Fr_G}$$

where E_{ZSL} = dispersion coefficient of liquid in a slurry (ft^2/sec).

Comparison of this correlation with an earlier correlation (4) for liquid dispersion in a gas/liquid system indicates that the liquid-phase dispersion coefficient is higher in a slurry than in a clear fluid, except when the Froude number is very small. This result differs from the findings of Vail et al.

Kato et al., (29) also studied the dispersion of the solid phase in a gas/liquid system. Their results lead to two empirical correlations, depending on the particle size. For small particles the dispersion coefficients for both the liquid and solid phases are identical. For large particles, the empirical condition includes the effect of particle diameter, as shown in the following expression.

$$E_{ZS} = \frac{V_{SG} D (1 + 8 Fr_G^{0.85})}{13 Fr_G (1 + 0.009 Re_p Fr_G^{-0.8})}$$

where $Re_p = d_p V_p / \nu_L$;

E_{ZS} = dispersion coefficient of solid in a slurry (ft^2/sec);

d_p = particle diameter (ft);

V_p = particle terminal velocity in stagnant fluid (ft/sec); and

ν_L = liquid kinematic viscosity (ft^2/sec).

The suspension of solid particles in a batch column depends on the gas-input rate. Roy et al., (36) and Imafuku et al. (32) have studied the critical gas velocities for a complete solid suspension. Critical gas velocity is defined as the minimum gas-input rate needed to fully suspend the solid particles. Imafuku et al., (32) observed that the critical gas velocity depends on the shape of the bottom inlet. They recommend conical bottom; the position of the gas distributor plate is also important to obtain complete solid suspension.

Our experience in three phase flow behavior indicates that pertinent data to define behavior in these systems is lacking.

3.3 Solids Accumulation

High solids accumulation has been experienced in the dissolvers at Ft. Lewis and Wilsonville Pilot Plants. Hydrocarbon Research, Inc., (HRI) when running an SRC study (44), also observed solids buildup in their experimental reactor. However, the most quantitative data on solids accumulation are those available from Wilsonville. For over 2 years, they have reported extensive data on rates of accumulation, quantities, particle sizes, and chemical composition of accumulated materials.

To show the levels of solids that can accumulate, data for four Wilsonville runs, each on a different coal, are presented in Table 3. The highest level of accumulation was in Run 100 with Amax Belle Ayr coal from which 1,300 pounds of reactor solids was discharged. This corresponds to a solids concentration of 72 lb/ft³. On a volume basis, these solids occupy about 40% of the total dissolver space. Whether or not these solids are completely suspended at this concentration is unknown. Obviously, such information is critical for designing a commercial vessel that has a diameter of about 11 ft, as proposed for the SRC-I Demonstration Plant.

In addition to the data from Wilsonville on solids composition, an extensive evaluation of reactor solids was carried out at Pennsylvania State University (45). That study led to an understanding of the mechanism of solids formation. The relative growth of particles was related to the calcium content in the feed coal: as calcium content increased, particle growth and the quantity of accumulated solids also increased.

Wilsonville also generated considerable data on the particle-size distributions of the recovered solids that accumulated when plant grind coal (95% through 200 mesh) was fed. The solids recovered from

Table 3

Wilsonville Solids Accumulation

Coal	Run length (days)	LHSV ^a in dissolver (hr ⁻¹)	Superficial velocities (ft/sec)		Solids in reactor (lb/ft ³)	% ash in solids
			Slurry	Gas		
Kentucky 9 & 14	68	1.7	0.011	0.09	33	83
Monterey	95	1.7	0.011	0.05	22	70
Indiana V	129	0.9	0.006	0.03	53	61
Amax-Bell Ayr.	100	1.1	0.007	0.06	72	71

^a LHSV, liquid hourly space velocity

the reactor grew considerably, as demonstrated by the amounts of material that accumulated on 25- and 50-mesh screens. For example, the particle-size distribution for the solids collected after Run 68, (Table 3) was as follows:

<u>Mesh size</u>	<u>wt %</u>
25	5
25-50	7
50-100	5
100-200	50
200-325	24
325	9

Obviously, the larger particles would not have been in the feed coal. Other Wilsonville reports shown even higher concentrations of larger particles in the reactor solids.

Since mid-1976, attempts have been made to control the concentration of solids in the Wilsonville reactor. Reactor solids were added at the start of individual runs to bring the dissolver to a rapid line-out. More recently, a solids withdrawal system was added to maintain a steady solids concentration in the reactor. With the capability to both add and withdraw solids, valuable data will undoubtedly be obtained that can be used for the commercial plant design. However, additional data at higher flows and with different geometries will be needed to effectively use such a withdrawal system on a larger scale.

At commercial design conditions, accumulation of solids will almost certainly occur. Exxon's results on 8-mesh feed coal showed the necessity for a solids removal system (46). Also, the DOE Project Review Meeting on Preheater Design held at Oak Ridge on 21 March 1979, C. Ackerman (P & M Mining Co.) reported 1/16-1/8 in. rock accumulation in the Ft. Lewis dissolver after only 3-12 days operation with 1/8 in. coal feed. Data on Ft. Lewis reactor design (see Section 2)

and data from Wilsonville (Table 3) Run 68 show that relative superficial gas velocities are very similar, i.e., 0.11 and 0.09 ft/sec, respectively. At these velocities, solids that accumulate when plant grind coal is used have particle sizes less than 100 mesh. Certainly much higher levels of accumulation would be expected with the larger coal feed size planned for the commercial plants.

The relevance of the above data at superficial gas velocities of 0.1 ft/sec is apparent when the alternate option of running the 6,000 tpd dissolver at 0.18 ft/sec (parallel dissolvers) is a distinct possibility. Knowing the behavior of solids in a column at these flow velocities is good engineering practice.

Very little data on the retention of solids in dissolvers are available from other sources. Ft. Lewis data have been limited largely to the time that a considerable plug was discovered in their dissolver (48). Otherwise, the behavior of solids in their system has not been studied systematically. H-Coal researchers have studied the flow rates that allow retention of solid catalysts within their reactor. However, the catalyst particles used in the H-Coal process are larger than the ash particles anticipated to accumulate in the SRC dissolver. The H-Coal process will probably use a 1/16 in. extrudate for their commercial design.

Apparently, particle growth within the dissolver causes the large accumulation of solids having particle sizes greater than those present in the feed coal. The program at Pennsylvania State University has shown that the particles can be attributed to a shell growth effect.

At Air Products, to understand the problem of solids accumulation, critical gas velocities were determined for different particle sizes in three liquid media using a 5-in.-diameter column. Critical gas velocity is the superficial velocity beyond which the particles will be in complete suspension.

Table 4 summarizes the results of the critical gas velocity study. The effect of particle size on solids suspension or critical gas velocity is significant; the critical gas velocity increases substantially as particle size increases. When the gas velocity dropped below the critical value, sand particles settled to the bottom and the gas channeled through the settled bed. The amount of settled solids decreased with increasing gas velocity. As the critical gas velocity was reached, the momentum transferred from the gas phase was large enough to suspend all the sand particles.

Also shown in Table 4 is the sieve opening (microns) for each particle range. Assuming spherical particles have diameters equal to the sieve opening, the settling velocities of sand particles were calculated based upon particle drag in an infinite water medium (Table 5). It is reasonable to assume that the largest particle, or the maximum sieve opening in each group, would govern the critical gas velocity. It is then interesting to note that the critical gas velocity in each group (Table 5) is approximately twice the settling velocity of the largest particle of the corresponding group (with the exception of Group II). Extrapolating these results, the Group VI particles (20-30 mesh) are predicted to have a critical gas velocity of about 0.7 ft/sec. In any event, the results of this critical gas velocity study strongly suggest the problem of settled solids when large particles are used.

The physical properties of a liquid also have a significant impact on solid suspension (Table 4). When a mixture of ethanol and water was tested, the critical gas velocities for Groups II and V were reduced by approximately 30% of those measured in pure water. After the physical properties of these two solutions were examined, two possible causes for this reduction were apparent: surface tension and viscosity. However, the experiment with pure ethanol eliminates the possibility of the surface tension effect. The surface tension of the mixed solution was determined experimentally to be 37.43 dynes/cm, which is more than a factor of two lower than that of pure water. If liquid surface tension were responsible for the observed reduction of the

Table 4

Effects of Particle Size and Liquid Properties
On Critical Gas Velocities

Group No.	Particle size		Critical gas velocities (ft/sec) at various liquid phases		
	Mesh	μm	Water	Ethanol/water ^a	Ethanol
I	140 minus	Less than 105	0.13	---	---
II	80-120	177-125	0.137-0.165	0.10-0.11	0.137-0.165
III	60-80	250-177	0.193-0.217	---	---
IV	45-60	345-250	0.298-0.362	---	---
V	30-45	595-354	0.482-0.503	0.344	---
VI	20-30	841-595	No full suspension detected up to 0.57-ft/sec gas velocity	---	---

^a 30 wt % ethanol/70 wt % water mixture.

Table 5

Particle Settling Velocity in Infinite Water Medium

<u>Mesh No.</u>	<u>Diameter(μm)</u>	<u>Settling velocity (ft/sec)</u>
80	177	0.0567
60	250	0.0971
45	354	0.1535
30	595	0.2592
20	841	0.3707

critical gas velocity in the mixture, pure ethanol, which has a surface tension (22.75 dynes/cm) lower than that of the mixture, should show a similar reduction. Surprisingly, our results showed that the critical gas velocity for the 80-120-mesh solids in pure ethanol was indistinguishable from that observed in pure water (Table 4). Hence, liquid surface tension is not responsible for the reduction effect observed in the mixture, thereby leaving the liquid viscosity as the sole explanation.

The solid settling velocity was lower in the mixed solution than in pure water because the viscosity between these two solutions differed by a factor of three. Going through the exercise of particle drag calculation, the settling velocities of a 30-mesh (595 microns) particle in the mixed solution and the pure ethanol were found to be 0.13 and 0.24 ft/sec, respectively. The insignificant difference between pure ethanol and pure water clearly explains the indistinguishable critical gas velocities measured in those two solutions. On the other hand, the twofold difference in the settling velocities of the 30-mesh particle in mixed solution and pure water explains the reduction in critical gas velocity. This also agrees with the findings on particle-size effect, which showed that critical gas velocity increased with increasing particle settling velocity. As a rough rule of thumb, the ratio of the critical gas velocity to the particle settling velocity is about 2, in the absence of liquid flow.

The operating conditions for the SRC-I pilot and demonstration plants were presented in Section 2. Results presented in Table 4 show that under demonstration plant conditions, solids smaller than 30-45 mesh would not be expected to accumulate, whereas both Wilsonville and Ft. Lewis dissolvers accumulate solids. The actual distribution of mesh sizes cannot be determined, because larger particles displace smaller ones and any size larger than 80-100 mesh may accumulate at Wilsonville. That this is clearly so at Wilsonville has been reported.

3.4 Data Needs for Dissolver Design

The design goal for the commercial dissolvers is to ensure that solids passing through the reactors remain suspended in sufficient quantity and distribution to provide the necessary catalytic effect. Solids and ash particles can accumulate over an extended operating period. Data from Ft. Lewis on rock accumulation in the dissolver when 1/8-in. coal is fed illustrates the problems that may occur commercially. From a design standpoint the Ft. Lewis results are quite alarming, because Tarrer and co-workers have shown that coal fed to the process unit would immediately dissolve into liquid, leaving extremely small ash particles suspended in the reactor.

Information necessary for dissolver design requires the following data:

- o Determine the necessary superficial gas and slurry velocities needed to suspend these particles at turndown velocities
- o Determine the solids sizes necessary to ensure adequate solids accumulation within the reactor at commercial linear liquid and gas velocities
- o Determine adequate slurry correlations to predict accumulations in the commercial reactors, especially for start-up mode and equilibrium operation after solids removal
- o Determine dispersion (backmixing) as a function of fluid and gas velocities, solids concentration, density, and particle size
- o Determine in a large system the effect of surface tension on gas void fraction and solids suspension
- o Determine factors that influence adequate solids removal

- o Determine the influence of distributor plates on fluid properties as a function of position relative to the plate
- o Determine the efficiency of solids removal with regard to plate design and location
- o Examine solids removal as a function of velocity through the draw-off line and its effect on dissolver behavior.

Upon starting the design of the 6,000-tpd SRC-I Demonstration Plant, Air Products realized the necessity for large-scale cold-flow simulator studies. A glass column 25 ft tall with a 12-in. internal diameter was selected as having the required material strength and affording maximum visual access to the behavior of the fluid within the column. This design also corresponds exactly to the dissolver in use at Wilsonville.

4.0 EXPERIMENTAL SECTION

4.1 Program Plan

This program consisted of the following tasks:

- Task 1 - Experimental Planning
- Task 2 - Two-Phase and Three-Phase Flow in Vertical Columns
- Task 3 - Effect of Distribution on Flow
- Task 4 - Solids Removal Study
- Task 5 - Organic Fluid Phase
- Task 6 - Effect of Liquid Properties
- Task 7 - Final Report

4.2 Cold Flow Model Equipment

Both a 5-in.-diameter by 5-ft-long Plexiglas column and a 12-in.-diameter by 25-ft-long glass column were used in the cold-flow simulation study. These two columns are described separately in the following sections.

4.2.1 5-Inch Diameter Column

A schematic of the 5-in. x 5-ft column is shown in Figure 1. Both the column and the conically shaped bottom inlet (60 degree angle with the horizontal) were fabricated from Plexiglas. The bottom was designed so that it would accept a distributor plate. Five sampling taps (A-E) were located along the column. In typical operation, slurry exited through the topmost opening (E) and gas vented through the uppermost top central opening.

Slurries were prepared by mixing sand and water in a 110-gallon feed tank equipped with a Chemineer agitator (10 in. propeller). Slurry was pumped into the column with a Moyno progressing cavity pump. Gas was co-fed into the bottom through the common opening. Slurry exiting

Figure 1

Schematic Of 5-Inch Diameter Plexiglass Column

the column was stored separately in a second drum to avoid any complications due to partitioning of solids of different particle sizes in the column. (Recycling into the feed vessel under certain conditions would result in a feed composition depleted in the larger particles.) A 10 cm³ syringe was also attached to the column at the bottom in order to inject tracer into the feed stream for dispersion studies.

4.2.2 12-Inch Diameter Column

A schematic of the 12-in. diameter column is shown in Figure 2. The flow paths of slurry and gas were identical in both columns although larger slurry inventory and equipment were required for the 12-in.-diameter column. The column comprised of five sections of 5-ft glass tubing interconnected with machined metal flanges. The column rested on scaffolding that completely surrounded the column. Guide bars at the top kept the column in true vertical alignment by preventing the column from tilting. The supporting structure had additional horizontal bars to serve as a ladder to service the column and sample the column through the ports in each interconnecting flange.

The inlet to the column was a conically shaped glass cone having a 60° angle with the horizontal. The cone was suspended from the support plate and could be removed for insertion of a distributor plate without having to dismantle the entire column. A 400 cm³ tube was attached to the column at the bottom for injecting tracer during dispersion studies. The top of the column was enlarged to disengage the gas from the liquid phase. The column was completely wrapped with metal screen (0.5-in. openings) to protect personnel from flying glass in case of rupture.

Two different distributors were examined in the 12-in. diameter column:

Figure 2
Schematic Of 12-Inch
Diameter Glass Column

o Distributor #1

This distributor which closely resembles that in Wilsonville's dissolver, has seven 9/16-in. openings with a 3-3/4-in. center-to-center spacing. A 3-1/2-in. long, with 9/16-in. internal diameter (i.d.) riser tube was welded to each opening (see Figure 3). Two 1/4-in. by 1-in. rectangular slots face each other at the end of each riser tube. The bottom of the distributor faces the inlet cone of the column. On the top side of the distributor, a 1-1/14-in. by 3-1/2-in. long tube protrudes through each hole in the plate. Each tube was covered by a cap, leaving a 5/32-in. spacing between the tube and the cap. The gas/slurry mixture flows through the riser tube and enters into the upper tube. This mixture continues to climb through the upper tube and then runs down along the spacing between the upper tube and the cap. The material finally flows out of the cap through a 1/4-in. gap between the cap and the distributor plate. The complete assembly is shown in Figure 4.

o Distributor #2

This plate has nineteen 9/16-in. openings with a 2-in. center-to-center spacing, as illustrated in Figure 5. Each hole has a riser tube identical to that in distributor No. 1, but a bubble cap and upper tube are not attached to the other side of the plate. The distributor is shown in Figure 6.

In the study of organic fluids, the entire 12-in.-diameter column was enclosed so that volatile flammable liquids could be handled safely. A schematic drawing of the enclosure and its associated safety features are depicted in Figure 7.

Figure 3

Distributor Number 1

Distributor Number 1
Figure 4

Figure 5

Distributor Number 2

Figure 6

Distributor Number 2

Figure 7

Schematic Diagram of the 12-Inch Diameter Column Enclosure

The enclosure completely encompassing the 12-in. diameter column is made of 1/2-in. thick Plexiglas and 1/8-in. thick aluminum sheets bolted to an aluminum angle-iron framework that stands independently on the floor. Horizontal bars were used to connect the enclosure to the surrounding scaffolding, ensuring a rigid structure. The lower 8 feet of the column is completely leak proof; and the strength of this section was checked hydraulically, to guarantee that it could hold the entire contents of the glass column in case of breakage. A set of damper blades was installed as a top cover for the enclosure and a large blower was installed to provide adequately vent for nitrogen and any entrained volatile fumes.

4.3 Experimental Procedures

The procedures described below are essentially the same for both the 5- and 12-in. diameter columns, the only difference being the quantity of tracer injected into the columns for liquid dispersion experiments. The variable ranges were chosen to include the operating conditions for the SRC-I Pilot and Demonstration Plants. Operation systems included: air/water, air/water/sand, nitrogen/tetralin, nitrogen/tetralin/sand, nitrogen/glycol-water, and nitrogen/glycol-water/sand. Operating conditions were as follows: Liquid superficial velocity, 0-0.05 ft/sec; gas superficial velocity, 0.05-0.50 ft/sec; solid particle size, 20-140 minus mesh; solid concentration, 0-40 lb/ft³.

4.3.1 Gas Holdup

Gas holdup and the effects of solid particles on holdup were measured both without and with fluid flow. In the absence of liquid flow, the column was completely filled with liquid and then gas was passed through the liquid at specified rates. Excess liquid exited the column at the top through a side opening. After waiting a period of 5 minutes to ensure that a steady state was achieved, the bottom valve was

closed to shut off the gas input. The final liquid level was measured; and the difference between the initial and final levels represented the gas holdup at that particular gas-flow rate.

With fluid flow, the liquid and gas entered the column through a centrally located opening at the bottom. Excess liquid exited the column through a side opening at the top. After a steady state was reached, the liquid level was measured. Then a common valve at the bottom was closed, stopping both liquid and gas flow simultaneously. The gas void fraction was measured as described previously.

Two different solid concentrations were used to study the effect of solids on gas holdup. For experiments using large particles, such as 20/30 mesh sand, a known amount of sand was initially placed in the column and the experiments were conducted as described above. The sand particles that were carried out of the column, if any, were collected to account for an average solids concentration during the run. For experiments using fine particles, such as -140 mesh sand, a slurry containing sand and water was pumped into the column at a specific flow rate. The experimental procedure for determining holdup was the same as described above.

In order to qualitatively understand the bubble sizes in the column, we photographed the column. The experimental procedure was similar to the one used to measure gas holdup. Both liquid and gas were allowed to flow through the column at predetermined flow rates. After a steady state was established, a common valve at the bottom of the column was closed, stopping both liquid and gas flow simultaneously. The aerated liquid level dropped as the gas phase left the system. This change of aerated liquid level as a function of time was recorded photographically at 1-sec intervals. From the rate of level drop, the relative bubble size could be determined qualitatively.

4.3.2 Liquid Dispersion

Liquid-phase dispersion was investigated using a tracer detection method. Sodium chloride was used as the tracer, and a conductivity probe mounted at the exit line of the column continuously monitored the ion conductivity of the solution. Liquid flow rates were measured periodically to ensure a steady-state flow. An ultrasonic doppler flow meter was also used to check for fluctuations in the liquid-flow rate. Gas-flow rate was monitored using a flow meter that was precalibrated with a dry-test meter.

In the typical experimental procedure, liquid and gas were passed through the column at the desired rates. A recorder monitored signal from the conductivity probe. After a steady base line was established on the recorder, a pulse of 400-cm³ concentrated NaCl solution (0.243 g/cm³) was injected into the bottom of the 12-in.-diameter column. Typical injection time was between 2 and 5 sec. The conductivity of the existing solution was recorded as a function of time. The experiment was terminated when the recorder signal reached the preset base line; at this point, gas flow was stopped, although water flow is used for 5-10 min to ensure that the column had been purged of residual salt before the next experiment started.

A slightly different procedure was followed with solid particles. To determine the effects of solids on liquid dispersion, a constant concentration of solids must be maintained in the column while other parameters such as gas and liquid flow rates are varied. For practical reasons, slurry was not pumped into the column. In order to perform just one slurry pumping experiment, more than 1 ton of sand would be required to prepare a slurry feed containing 20 lb/ft³ of sand at a slurry feed rate of 17.5 gpm (the highest rate used in these experiments). Handling that volume of sand is extremely difficult and time-consuming. Instead, a different procedure was used that maintained the desired solids concentration and eliminated the need for massive amounts of material sand. First, the column was partially filled with water. A predetermined amount of sand (20/30 mesh) was added to the column.

Then the column was completely filled with water at a low flow rate. The 20/30-mesh sand particles were large enough to prevent their exiting the column during this slow filling process. However, minor amounts of sand did leave the column during the time that a steady state was being achieved. During the experiment, solid concentration in the column was maintained by monitoring the amount of sand that exited the column.

Plots of conductivity vs. time obtained from these experiments were transformed to dimensionless curves, and tracer curves were prepared. The tracer curves were then fitted with theoretical curves predicted from an axial dispersion model, which was based on the analogy between mixing in the actual flow and a diffusional process. A computer model was also developed to generate a family of axial dispersion curves having different Peclet numbers. The dispersion coefficient in the liquid phase was calculated from the Peclet number of the best matching curve.

A different method was used to measure the liquid dispersion coefficient in a nitrogen/tetralin system. The liquid dispersion experiments were performed in batch mode (no liquid flow). Typically nitrogen was bubbled at a desired rate through the column filled with tetralin. A concentrated blue dye solution was injected at the bottom of the 12-in.-diameter column. The blue dye solution was selected as trace for tetralin dispersion because (1) it is completely miscible with tetralin (2), it exhibits a maximum absorbance at a wavelength significantly different from that of tetralin, and (3) only a small quantity is needed to give good measurement sensitivity. Samples were taken 20 ft above the injection point every 20 sec. The amount of sample drawn from the column was kept very small to avoid disturbing the equilibrium level. The concentration of dye in the sample was determined from its adsorbance as measured by a colorimeter.

In the nitrogen/tetralin/sand system, the solid particles in the sample were allowed to settle before adsorbance measurement. Then the adsorbance vs. time plots were transformed into dimensionless curves, from which the liquid dispersion coefficients were determined.

4.3.3 Solid Dispersion

Solid dispersion experiments were conducted in both 5-in. and 12-in.-diameter columns using batch as well as continuous operating modes. In batch operation, gas was bubbled through the column, which was filled with liquid and a known weight of solid particles. During a 30-min bubbling period, steady-state conditions were established for each gas velocity. Then slurry samples were withdrawn from sampling ports at various column heights and the concentration of solids was measured.

During continuous operation, the water/sand slurry flowed continuously through the column. Samples were withdrawn from all ports at periodic intervals over 4-6 hr until a steady-state condition was achieved. Two different types of continuous experiments were conducted using the 5-in.-diameter column. In the first method, a water/sand slurry of known concentration was fed to the column on a once-through basis. This method, however, required a minimum of two feed tanks in series and an enormous amount of effort and material to keep up with the once-through operation. The second method involved recycling the column exit stream back to the feed tank; a slurry of known concentration was prepared in a feed tank, and the exit from the 5-in.-diameter column was returned to the feed tank, thus creating a closed loop.

In experiments with the 12-in.-diameter column, two feed tanks were used, and the exit from the column was returned to the first tank, which in turn was fed to the second tank in series. The advantages of this method are: a) less sand is required per experiment; (b) less labor is needed to prepare a fresh batch of slurry every 10 min.; and (c) there is less possibility of human error than when multiple batches are prepared in a short time. However, the solids accumulation measured from the recycle mode differs significantly from that measured for the once-through operation because the particle-size distribution of the feed material is different. In the recycle mode, the larger particles of the feed initially prepared in

the tank are depleted because they are retained in the column. This leads to a slightly higher estimate of solid accumulation based on the steady-state feed concentration and size distribution. This difference decreased with a decreasing particle-size range in the feed, however. Therefore, using a narrow size range could reduce the error caused by the recycle mode operation. One set of operating conditions was duplicated in both methods to check the experimental reproducibility and sample relationship to each other. This result will be presented in a later part of this report (discussion of solid dispersion).

4.3.4 Solid Removal

Solids removal experiments were similar to solids distribution experiments, with minor modifications. Slurry and gas were allowed to flow through the column in a recycle loop for several hours to achieve steady state. Samples were taken along the column to obtain profiles of steady-state solids concentration. Then the lowest sample port in the column (just above the distributor plate) was opened, and the slurry was allowed to flow into a storage tank while the slurry and gas flow into the column was maintained. Simultaneously, recycle operation was changed to once-through operation. However, in the once-through mode, the length of the experiment was limited by the slurry inventory in the holding tanks. Most of the experiments were terminated 15 min after the lowest sampling port was opened. In this short time, multiple sets of samples were obtained along the axis of the column in order to measure changes in the solids distribution profile.

4.3.5 Particle/Particle Interaction

A special set of experiments was undertaken to investigate the effects of settled particles and particle/particle interaction on solids accumulation in the 12-in.-diameter column. Three consecutive experiments were performed using various particle sizes: -140 mesh, 60/80 mesh, and 20/30 mesh. A slurry containing -140 mesh particles

and water was prepared in the feed tank. This fine-particle slurry recycled through the column at a liquid velocity of 0.05 ft/sec, with gas flowing at a velocity of 0.33 ft/sec. A steady-state solid concentration profile was obtained by taking samples along the column. At the end of the run, a known quantity of 60/80 mesh sand was added to the feed tanks, and the new slurry was recycled through the column at the same operating conditions. Samples were withdrawn from the sampling ports until a new equilibrium state was reached. In these samples, in addition to determining the total solids concentration, the concentration of different particle sizes was also determined. At the end of the experiment, a known amount of 20/30 mesh particles was added to the column to simulate the presence of settled particles. The slurry was allowed to flow through the column at identical operating conditions. Again concentration of different-sized particles and the total solids concentration were measured for each sample withdrawn from the column. At the end of the experiment, the column was drained and the sand retained in the column was weighed.

4.3.6 Gas/Liquid Mass Transfer

Gas/liquid mass-transfer studies were also conducted in the 5-in. and 12-in.-diameter columns, in a batch operation (no liquid flow). The mass-transfer rate was measured as the rate of oxygen dissolving in water. Polarographic probes (stationary and submersible) continuously monitored the concentration of dissolved oxygen in the solution as a function of time. The experimental procedure described below was used with for both columns.

The oxygen content of water in the 5-in.-diameter column was measured with stationary probes installed at various column heights. For the 12-in.-diameter column, the submersible probes were used because the column structure prohibits installing stationary probes. Also, several measurements were made using both types of probes in the 5-in.-diameter column at various heights. Negligible differences in the measurements ensured the validity of using the submersible probes.

At the beginning, nitrogen was allowed to flow through the column filled with water to remove oxygen. After the oxygen concentration in the dissolver leveled out at zero for several minutes, the nitrogen flow was stopped and air was introduced at a predetermined rate. Changes in the dissolver oxygen concentration as a function of time were then recorded at three different positions for the 5-in. diameter column and at two different positions for the 12-in. column. The rate of oxygen uptake was used to calculate the gas/liquid mass-transfer coefficient using a completely backmixed model.

5.0 RESULTS AND DISCUSSION

5.1 Gas Holdup

Gas holdup as a function of several variables was measured in 5-in. and 12-in.-diameter columns using various liquid systems. The variables were: column diameter, liquid velocity, gas velocity, entrance effects, presence of solids, physical properties of the liquid phase. In addition, several correlations were tested to determine the best one that could predict the experimental data reasonably well for all the liquid systems tested.

Table 6 lists the operating conditions for the experimental runs. Sand was used in all experiments using the solid phase. The particle size varied from 20/30 mesh to -140 mesh, and the solid concentrations varied from 5 to 20 lb/ft³. The liquid phases studied were water, tetralin, ethylene glycol, and mixtures of various concentrations of ethylene glycol and water. Three different distributor configurations (absence of a distributor, distributor #1 and distributor #2) and two different columns (5- and 12-in. in diameter) were used. Liquid velocity was varied between 0.0 and 0.05 ft/sec, and gas velocity was varied between 0.0 and 0.45 ft/sec.

The operating conditions encompass the operating conditions at the Wilsonville Pilot Plant and two possible configurations (series vs. parallel) being considered for the demonstration plant dissolvers. In addition to the experiments listed in Table 6, a photographic technique was used to determine the fraction of large bubbles present in the column and the effect of different distributors.

5.1.1 Effect of Column Diameter

For a given superficial gas velocity, gas holdup is shown independently of column diameter. Our earlier results using air/water systems clearly indicated that increasing the column diameter from 5 to 12 in. had no appreciable effect on gas holdup. Results obtained with

Table 6
Operating Conditions

Column diameter (In.)	System	Distributor ^a configuration	Range of			
			Gas velocities (ft/sec)	Liquid velocities (ft/sec)	Solids particle size (mesh)	Solids concentration (lb/ft ³)
12	Air/water	1	0.0-0.37	0.0-0.06	0	0
12	Air/water	1	0.0-0.37	0.0-0.08	-100	5
12	Air/water	1	0.0-0.37	0.0-0.05	-100	11
12	Air/water	1	0.0-0.37	0.0-0.05	20/30	5
12	Air/water	1	0.0-0.37	0.0-0.05	20/30	13
12	Air/water	0	0.0-0.43	0.0-0.05	0	0
12	Air/water	0	0.0-0.43	0.0-0.05	20/30	5
12	Air/water	0	0.0-0.43	0.0-0.04	-140	5
12	Air/water	2	0.0-0.43	0.0-0.05	0	0
12	Air/water	2	0.0-0.43	0.0-0.05	20/30	5-20
12	Air/water	2	0.0-0.43	0.0-0.05	-140	5
5	Nitrogen/tetralin	-	0.0-0.40	0	0	0
5	Air/glycol	-	0.0-0.40	0	0	0
5	Air/90% glycol-10% water	-	0.0-0.40	0	0	0
12	Nitrogen/tetralin	1	0.0-0.39	0	0	0
12	Nitrogen/tetralin	1	0.0-0.39	0.02-0.05	-140	5
12	Nitrogen/tetralin	1	0.0-0.39	0.02-0.05	60/80	5
12	Air/glycol	1	0.0-0.39	0	0	0
12	Air/90% glycol-10% water	1	0.0-0.39	0	0	0
12	Air/70% glycol-30% water	1	0.0-0.39	0	0	0
12	Air/50% glycol-50% water	1	0.0-0.39	0	0	0

^a 1, Distributor #1; 2, Distributor #2; 0, absence of a distributor

tetralin (Figure 8) confirmed this. This finding also agrees with results of other investigators, who concluded that gas holdup is independent of column diameters above a diameter of 3 in. (25).

Most of the published gas holdup data were obtained from small diameter columns (6 inches or less). The effect of column diameter on gas holdup in viscous liquid has not been fully investigated by using large diameter columns. In this study, we found that gas holdup in ethylene glycol (20 cP) is higher in a 5-in. diameter column than in a 12-in. diameter column (Figure 9). It is speculated that the effect of column diameter on gas holdup for viscous materials may differ from that for systems of low viscosity, i.e. the minimum diameter above which column diameter has no effect on gas holdup may be larger for more viscous materials. We hypothesize that the larger gas holdup in the 5-in. column for ethylene glycol is due to wall effects. However, due to the limited data, we could not determine the above mentioned minimum diameter for ethylene glycol system.

5.1.2 Effect of Liquid Velocity

Gas holdup in a bubble column is independent of liquid velocity. This was confirmed for both air/water and nitrogen/tetralin systems and is true irrespective of the presence or absence of solids. Figures 10 and 11 plot typical data from the two different liquid systems used in this study. Data obtained with different distributor configurations also shows that gas holdup is independent of liquid velocity.

5.1.3 Effect of Gas Velocity

Gas holdup increases with an increase in superficial gas velocity, in all the systems studied (Figures 12 and 14). The figures clearly show that the dependence of gas holdup on linear gas velocities can

FIGURE 8
EFFECT OF COLUMN DIAMETER
ON GAS HOLDUP
TETRALIN/NITROGEN

Figure 9

Effect of Column Diameter on Gas Holdup
(Air/Glycol)

○ = 5" column

□ = 12" column

System = air/ethylene glycol

Figure 10

Effect of Liquid Velocity on Gas Holdup
(Air/Water)

Figure 11

Effect of Liquid Velocity on Gas Holdup
(Tetralin/Nitrogen)

$P_s = 140$ minus mesh
 $V_G = .216-.392$ ft/sec
12" Column

Figure 12
Effect of Gas Velocity on Gas Holdup
(Air/Water)

Figure 13
Effect of Gas Velocity on Gas Holdup
(Tetralin/Nitrogen)

- = 100% Glycol
- = 90%
- = 70%
- △ = 50%

Figure 14
Effect of Gas Velocity on Gas Holdup (Air/Glycol)

be described fairly well by a straight line on a log-log plot for all the systems studied. This indicates that gas holdup varies exponentially with linear gas velocity; in other words, gas holdup is proportional to $(V_g)^x$, where V_g is the linear gas velocity and x is an exponent. The value of the exponent was calculated by a regression analysis on the data (see Table 7). It is interesting to note that for four of the six systems studied, the exponent is nearly the same. For low concentrations of glycol (50 and 70 wt%) the exponents were different. This difference is not clearly understood. Perhaps the physical properties of the system may have had an effect on the exponent. Nonetheless, the range of the exponents listed in Table 7 (0.52-0.79) agrees with the values reported by other investigators.

5.1.4 Effects of Solids

The presence of solids has a minimal effect on gas holdup. A decrease in gas holdup was observed at very high gas velocities (0.3 ft/sec) in the presence of fine particles (-100 mesh). In this project, extensive data were obtained using an air/water/sand system, and limited experiments were conducted using a nitrogen/tetralin/sand system.

Figure 15 shows the effect of solid particles on gas holdup using the air/water/sand system. At low superficial gas velocities (up to 0.10 ft/sec), solids did not change the gas holdup appreciably. However, at high velocities of solids decreased gas holdup.

Certain conditions also affected the influence of solids concentration on gas holdup. In the presence of small particles, such as -100 mesh sand, gas holdup decreased, compared to the two-phase system, as gas velocity and solids concentration increased. A 13.9% reduction was measured for a solids concentration of 11 lb/ft³ at a superficial gas velocity of 0.368 ft/sec.

Table 7

Value of Exponent for Various Systems

$$\epsilon_g \propto (V_g)^b$$

ϵ_g = gas holdup

V_g = gas velocity

b = exponent

System	Value of exponent, b
Air/water	0.79
Nitrogen/tetralin	0.73
Air/100% glycol	0.77
Air/90% glycol	0.75
Air/70% glycol	0.52
Air/50% glycol	0.64

Figure 15

Effect Of Solids On Gas Holdup In 12-Inch Diameter Column

The larger particle-size solids decreased gas holdup only at the highest gas velocity and at high solids concentration (12.7 lb/ft^3). The magnitude was less than that for the smaller size material. At lower concentration no change was observed.

Gas holdup was also influenced by particle suspension in this flow regime. The 20-30 mesh sand in this work settled mainly at the bottom of the 12-in. diameter column and reduced gas holdup less than the -100 mesh suspended particles. The suspended particles may have enhanced the coalescence of gas bubbles more effectively than the settled solids, thus reducing in gas holdup more. Therefore, the effect of increasing solid concentration could be related to an increase in bubble coalescence. In any event, with the exception of high concentration of fine particles, the presence of solids did not significantly affect gas holdup.

Figure 16 shows the effect of solid particles on gas holdup using the nitrogen/tetralin/sand system. Addition of 60/80-mesh sand resulted in negligible changes in gas holdup; however, gas holdup data for the -140 mesh minus sand was about 4-5% lower. These findings agree with data obtained for the air/water/sand system.

5.1.5 Effect of Distributors (Entrance Effects)

The presence or absence of a distributor does not seem to markedly affect gas holdup and negligible differences were found in the values of gas holdup for two different types of distributors.

Gas holdup values were determined for the air/water and air/water/sand system for three different distributor configurations: distributor #1 (7 bubble caps), distributor #2 (19 openings but no bubble caps), and without either distributor, gas velocities varied from 0.05 to 0.43 ft/sec, and liquid velocities ranged from 0.01 to 0.05 ft/sec.

FIGURE 16
EFFECT OF SOLID PARTICLES
ON GAS HOLDUP

Overall gas holdup was not influenced by entrance effect, as shown in Figures 17 through 22. These figures compare gas holdup data obtained from the three different distributor configurations under various experimental conditions (presence and absence of liquid flow and solid particles). Except for the case without a distributor at high gas velocities, negligible differences exist in the gas holdup values.

At higher gas velocities, an unusual behavior was observed in the absence of a distributor. The liquid level at the top of the column periodically surged such that the liquid level would rise above the exit line and then suddenly drop a few inches below the exit line. This phenomenon occurred at gas velocities above 0.2 ft/sec. Because of this surging behavior, the holdup data at high gas velocities in the absence of a distributor may be substantially in error. Periodic gas slugs that were observed in the column in the absence of the distributor are suspected to be the major cause of the surging behavior.

However, the presence of these huge bubbles (slugs) did not appreciably change the holdup values, indicating that the true fraction of large bubbles in the column at a given time may be quite small. Gas holdup data alone cannot sufficiently reflect the size of the bubbles present in the column. Because no attempt was made to measure bubble sizes experimentally, a photographic method was developed to identify the fraction of gas-void volume occupied by large gas bubbles.

The photographic method was detailed in Section 4.3. Gas holdup is normally measured by stopping the flow of liquid and gas into the column and measuring the drop in liquid height. In the photographic method, the drop in the liquid level at the top of the column was measured every second after the liquid and gas flows had been shut off. Data were taken at various liquid and gas velocities and with distributors #1 and #2 in place.

The data were interpreted assuming that large bubbles rise faster than smaller bubbles. A limitation of the photographic method is that it cannot quantitatively measure bubble size. However, one can

Figure 17. EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP IN THE ABSENCE OF LIQUID FLOW AND SOLID PHASE

Figure 18. EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT LOW LIQUID VELOCITY WITH HIGH CONCENTRATION OF LARGE SOLID PARTICLES

Figure 19. EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT HIGH LIQUID VELOCITY WITH LOW CONCENTRATION OF LARGE SOLID PARTICLES

Figure 20. EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP
AT LOW LIQUID VELOCITY WITH NO SOLIDS

Figure 21. EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT HIGH LIQUID VELOCITY WITH LOW CONCENTRATION OF FINE PARTICLES

Figure 22. EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT LOW LIQUID VELOCITY WITH HIGH CONCENTRATION OF FINE PARTICLES

characterize the bubble population in the column into two distinct groups: (1) a group of bubbles of about the same size rising in the column with a constant velocity and (2) a group of bubbles of various sizes, but larger than those belonging to the first group, and also rising at a faster rate. Observations indicated that the drop in liquid level (after the gas and liquid flows had been shut off) is very unsteady in the beginning, but becomes smooth with time, indicating the possibility of at least two different rates of drop in liquid level. Our objective was to determine these two different rates and the fraction of large bubbles present in the column.

Our results showed that the fraction of large bubbles increases with gas velocity, (see Figures 23 thru 31). Figures 23-26 represent data taken using distributor #1 and Figures 27-31 represent data taken with distributor #2. The figures indicated that two different rates of drop in the liquid level do exist. The bars indicated that the liquid level was unsteady and represented the low and high values observed from the photograph. Looking at a series of photographs taken at 1-sec intervals one can see the change in the behavior of liquid level. Figures 32 and 33 are two typical photographs taken 3 and 21 sec respectively, after the gas and liquid flows were stopped by closing the bottom valve. In Figure 32 the liquid level was not horizontal, and the liquid was very turbulent. Twenty-one seconds after the valve was closed, the liquid level was still dropping; however, as can be seen from Figure 33 the liquid level was now horizontal.

The type of distributor used does not affect the fraction of large gas bubbles, indicating that the bubble size distribution in this column is practically independent of the distributor design. Figures 23-31 clearly indicate that, for similar operating conditions, use of either distributor produced identical results. Hence, the results will be discussed together.

Figure 23.

RATE OF THE AERATED LIQUID LEVEL DROP

Figure 24.

RATE OF THE AERATED LIQUID LEVEL DROP

Figure 25.

RATE OF THE AERATED LIQUID LEVEL DROP

Figure 26.

RATE OF THE AERATED LIQUID LEVEL DROP

Figure 27. RATE OF THE AERATED LIQUID LEVEL DROP

Figure 28. RATE OF THE AERATED LIQUID LEVEL DROP

Figure 29.

RATE OF THE AERATED LIQUID LEVEL DROP

Figure 30. RATE OF THE AERATED LIQUID LEVEL DROP

Figure 31. RATE OF THE AERATED LIQUID LEVEL DROP.

Figure 32

Photograph of Liquid Level (Three seconds after closing the valve)

Figure 33

Photograph of Liquid Level (Twenty-one seconds after closing valve)

The photographic technique confirms some of our earlier results and provides a few qualitative conclusions: bubble size distribution as well as gas holdup are independent of liquid velocity. A change in liquid velocity from 0.02 to 0.05 ft/sec did not change the rate of liquid level drop or the final liquid level. This behavior was observed at two different gas velocities (0.25 and 0.43 ft/sec), which suggests that this behavior is more than coincidence.

The fraction of large bubbles in the column increases with gas velocity. At very low gas velocities, the transition between the two distinct regions could not be seen because of the obstruction caused by the connecting flanges in the column. However, Figure 27 suggests that the fraction of large bubbles in the column, if any, is very small at a gas velocity of 0.05 ft/sec. Increasing the velocity from 0.25 to 0.43 ft/sec results in an increase of the large bubbles from 31 to 43%. The graphs also reveal that the average size of the smaller bubbles (a group of bubbles of about uniform size) is independent of gas velocity, even though the fraction of large bubbles in the column increases with gas velocity. Superimposing Figures 23-31, which are plotted for different gas velocities, reveals that the data points are parallel, indicating that the slopes are the same. Because the slopes correspond to the average velocity of the small bubbles, this finding of parallel data points at different gas velocities means that the small-bubble rise velocities are independent of gas velocity.

Our results indicate that distributor configuration does not affect bubble-size distribution or the average gas holdup. However, we have not measured or attempted to measure local gas holdup values. These values might differ in the vicinity of the distributor. The large L/d ratio (25) of the 12-in.-diameter column may help to diminish entrance effects on gas holdup.

5.1.6 Effect of Physical Properties

Physical properties of the liquids affect gas holdup. In this project, the viscosity of the liquids used varied from 1 to 19 cP, the surface tension from 32 to 72 dynes/cm and the density from 0.97 to 1.17 g/mL (see Table 8). Because more than one property was varied at any given time, the effect of physical properties cannot be completely analyzed. However, based on our data and published data, a few qualitative conclusions can be drawn.

Both liquid viscosity and surface tension affect gas holdup. A decrease in either seems to increase gas holdup. Comparing our results with water and tetralin (Figures 12 and 13), one can see that gas holdup was larger when tetralin was used. The surface tension of tetralin is 32 dynes/cm, compared to 72 dynes/cm for water but tetralin has a higher viscosity (2.65 vs. 1 cP). Even though these are opposite effects, the increased holdup indicates that surface tension may have a greater effect than viscosity. Hikita and Kikukawa's correlation confirms this conclusion.

Experiments with glycol/water mixtures show that both viscosity and surface tension affect gas holdup: decreasing the glycol content from 100 to 50 wt % increased gas holdup. At the same time, both viscosity and surface tension were changed. Due to the significant drop in viscosity from 19.00 to 4.1 cP gas holdup should have increased although the surface tension was increased slightly from 48.2 to 56.7 dynes/cm.

Although density was varied only slightly, it also appeared to affect gas holdup. This effect is obvious when comparing results with tetralin and 50 wt % glycol systems. Both the viscosity and surface tension of tetralin are lower than those of the 50 wt % glycol/water mixture. Hence one would expect tetralin to exhibit a higher gas holdup. However, the differences in holdup values for the two systems are negligible. Tetralin had a lower density than the 50 wt % glycol/water mixture, indicating that liquid density also may play an important role.

Table 8

Physical Properties of Different Liquid Systems

Liquid system	Temp.(°C)	Density (g/ml)	Viscosity,(cP)	Surface tension (dynes/cm)
100% glycol	25	1.17	19.00	48.2
90 wt. glycol	21.6	1.107	13.08	45.9
70 wt. glycol	21.6	1.091	6.40	48.8
50 wt. glycol	18.5	1.078	4.10	56.7
Tetralin		0.97	2.65	32
Water		1	1	72

Our results show that physical properties do affect the gas-holdup values. However, because our experiments were limited, a detailed discussion of the effect of every macroscopic physical property is not feasible. In addition, one should point out that foaming characteristics of these liquids also play an important part in determining gas holdup. Unfortunately, just by knowing the values of the density, surface tension, and viscosity, one cannot predict the foaming characteristics. At this point, a rational approach is to predict gas-holdup values for the demonstration plant dissolver with reasonable accuracy.

5.1.7 Correlations

Several correlations for determining gas holdup in bubble columns have been published. Recently, Irwin et al., (50) examined several correlations in order to determine the most suitable for modelling coal-liquefaction dissolvers. Correlations were within $\pm 20\%$ of each other, indicating that more than one correlation can be successfully used to predict gas holdup. Because several correlations can be used successfully, developing a new correlation is not really necessary; in addition, the limited amount of data available from this program is not sufficient to develop a good correlation. Hence, in this section, we attempt to identify the correlation(s) that can be used to predict our experimental data.

Table 9 lists the correlations that were used to predict gas holdup in bubble columns. These correlations show that different investigators found that physical properties of the fluid systems affect gas holdup differently. In general, most observers found that gas holdup can be simply expressed by:

$$\epsilon_g \propto V_g^n$$

where V_g is the superficial gas velocity. The values of the exponent vary widely in this investigation between 0.52 and 0.79 for the systems that were studied. Very few correlations consider the gas

Table 9

Correlations for Predicting Gas HoldupAkita and Yoshida (26)

$$\frac{\epsilon_g}{(1 - \epsilon_g)^4} = C_1 (N_{Bo})^{1/8} (N_{Ga})^{1/2} (N_{Fr})$$

Hikita and Kikukawa (3)

$$\epsilon_g = 0.505 V_G^{0.47} \left(\frac{72}{\sigma}\right)^{2/3} \left(\frac{1}{\mu}\right)^{0.05}$$

Hughmark (16)

$$\epsilon_g = \frac{V_{sG}}{U_s} \left[\left(\frac{62.4}{\rho_L}\right) \left(\frac{72}{\sigma}\right) \right]^{1/3}$$

Pilhofer et al. (51)

$$\frac{\epsilon_g}{1 - \epsilon_g} = 0.115 [V_g^3 / (u g \Delta \rho / \rho_L)]^{0.23}$$

 ϵ_g = gas holdup V_g = superficial gas velocity (ft/sec) U_s = slip velocity (ft/sec) ρ_L = liquid density (lb/ft³) σ = dynes/cm D = diameter of column (ft) C_1 = 0.20 for nonelectrolytes

0.25 for electrolytes

 u = kinematic viscosity (ft²/sec) g = gravitational acceleration
32.2 (ft/sec²) μ = liquid viscosity (cp) $N_{Bo} = g D^2 \rho_L / \sigma$ (Bond number) $N_{Ga} = g D^3 / u_L^2$ (Galileo number) $N_{Fr} = V_g / \sqrt{g D}$ (Froude number)

properties. However, for the system we studied, gas properties were similar because nitrogen and air were used. Most investigators agree that surface tension dramatically affects gas holdup, even though the magnitude of the effect varies. Pilhofer et al. (51) did not include the effect of surface tension in their correlation. Several investigators related gas holdup to a ratio of the surface tension of the liquid studied and that of water. The effect of viscosity on gas holdup has also been recognized by most investigators; however, a few have ignored the effect of liquid viscosity. Also, few investigators have considered the effect of liquid density on gas holdup.

Earlier, we showed that gas holdup is affected by both surface tension and viscosity. Hence a single correlation that could fit all our data should include both viscosity and surface tension. Akita and Yoshida's correlation (12) fits in this mold. A few other correlations were also investigated. Akita and Yoshida's correlation predicted the results from air/water reasonably well, but underpredicted our results for tetralin and glycol/water mixtures. For tetralin, correlations of Pilhofer et al. (15) and Hughmark (16), predict the data reasonably well (Figure 34). Hughmarks's correlation does not predict the glycol/water very well, possibly because viscosity is not considered.

Pilhofer's correlation does not reasonably predict gas holdup for the glycol/water mixture at concentrations we studied; for pure glycol it does provide a reasonable fit with the data. Akita and Yoshida's correlation has a proportionality constant; the value recommended for this constant (C_1) is 0.20 for nonelectrolytes and 0.25 for electrolytes. If one considers ethylene glycol as a weak electrolyte, then the use of 0.25 for the constant results in good predictions for the gas-holdup values. Tetralin is not an electrolyte. However, the use of 0.25 for C_1 in the case of tetralin also results in a reasonably good agreement with experimental data.

Figure 34

Comparison of Hughmark's Correlation with Experimental Data

One can conclude that a single correlation is not effective in predicting our experimental gas-holdup values. Akita and Yoshida's correlation is quite reasonable except for the tetralin data, but with minor corrections, it can be used effectively.

5.2 Liquid Dispersion

The axial liquid dispersion coefficient (E_{zL}) is an important quantity that is needed to describe the hydrodynamics taking place in the dissolver. Its magnitude indicates how much turbulent exchange takes place axially in the column. Usually, the assumption is made that radial liquid dispersion is much higher than axial. Hence, uniformity of liquid velocity and dye concentration in the radial direction can be assumed.

Most of the experiments conducted for axial liquid dispersion were in the 12-in. column, although two runs with the tetralin were conducted in the 5-in. column. We used two methods to measure the axial dispersion number: batch (no liquid flowing) and continuous (liquid flow).

5.2.1 Theoretical Background

5.2.1.1 Batch Model

A basic one dimensional diffusion model was applied to our data. We used an equation from Ohki et al. (5):

$$\frac{\partial c}{\partial t} = E_{zL} \frac{\partial^2 c}{\partial z^2} \quad (1)$$

where c = tracer concentration, z = axial distance (ft), and E_{zL} = liquid axial dispersion coefficient (ft^2/sec).

Assuming an impulse injection, the approximate solution is obtained with equation 2.

$$C' = 1 + 2 \sum_{n=1}^{\infty} \left[\left(\cos \frac{\pi n}{L} Z \right) \exp \left(-n^2 \frac{\pi^2}{L^2} E_{zL} t \right) \right] \quad (2)$$

where C' = nondimensional tracer concentration and L = total length of aerated liquid.

This approximate solution is within 1% of the complete analytical solution. Equation 2 was used to obtain the value of E_{zL} . Hence by obtaining Z/L at a particular sampling site and determining C' as a function of t from the experiment, one can calculate E_{zL} . Ideally, one point on the curve of C' and t should be sufficient to determine E_{zL} . In actuality, three points were selected in the range of C' between 0.3 and 0.8, from which three values of E_{zL} were calculated. The average E_{zL} value was then reported for that particular gas velocity. A separate method was used to determine E_{zL} and to compare with the average value obtained from the technique described above. This method involves at least-square fits program to optimize the value of E_{zL} to best fit the entire tracer curve. A comparison of these two methods will be discussed later.

5.2.1.2 Continuous Model

The basic one-dimensional model used for the continuous-flow experiments is:

$$\frac{\partial c}{\partial t} + \frac{\partial c}{\partial z} = \frac{E_{zL}}{uL} \frac{\partial^2 c}{\partial z^2} \quad (3)$$

where c = non-dimensional tracer concentration; t = non-dimensional time from injection; z = non-dimensional axial distance; E_{zL} = axial liquid dispersion coefficient (ft^2/sec); L = length of dispersion vessel considered (ft); u = liquid flow velocity (ft/sec); $E_{zL}/uL = 1/\text{Pe}$; and Pe = Peclet number. Boundary conditions used for this model are for closed vessel. The model was solved numerically.

5.2.2 Air/Water and Air/Water/Sand Systems

Liquid dispersion experiments were performed in the 12-in. diameter column for both two- and three-phase systems. The ranges of variables investigated are summarized in Table 10. Fits of all the experimental tracer curves with theoretical curves from the axial dispersion model are displayed in Appendices A and C. In these curves, circles represent the experimental tracer curve and the solid line represents the theoretical predicted curve.

Matching experimental and theoretical curves was found to be sensitive to column height. Because the dispersion number (E_{zL}/uL , a dimensionless quantity) is inversely proportional to the height of the column, determination of the dispersion number by the best-fit technique became more difficult with increasing column height. The curve-matching technique was extremely sensitive to any tailing in the curve. Since some fits were less perfect than others because of tailing problems, a consistent method of selecting the dispersion numbers was developed. Dispersion numbers were always determined by fitting the peak time (peak position) of the tracer curve with the dispersion model, because peak time was not affected by tailing in the experimental curve. Tailing only changed the height of the curve. As long as the mean residence time could be measured accurately, peak position was the best way to match dispersion numbers.

Fluctuations in liquid flow could cause errors in the determination of mean residence time if averages of several discrete measurements were used. In addition, mean residence times, which were calculated using the experimental tracer curve, were compared with the values calculated using average flow rates. Whenever a large (greater than 10%) discrepancy existed between the two numbers, the residence time calculated using the experiment tracer curve was used to generate the tracer curves. This procedure also improved data analysis.

Table 10

Liquid Dispersion: Two- and Three-Phase Flow

Variable Studies

Liquid velocity	0.01-0.07 ft/sec
Gas velocity	0.0-0.43 ft/sec
Particle size	20/30 mesh, 60/80 mesh, -100 mesh, and -140 mesh
Solids Concentration	2-20 lb/ft ³
Distributor	No distributor, distributors #1 and #2

Incorporating all these improvements, a new method was tried to determine the liquid dispersion coefficients for experiments using distributor No. 2. Based on the theoretical curves for the axial dispersion model, a curve showing the relationship between peak time and Peclet number was developed. This curve is shown in Appendix B. Knowing the peak time from the experimental data, the Peclet number can be determined using the above curve. This method is less time consuming than trying to match the experimental and theoretical curves. In order to check the precision of the method, a few liquid dispersion coefficients were determined by using both methods; namely, by matching the experimental and theoretical curves and by using the peak time vs. Peclet number curve. The comparison is shown in Table B-1 (Appendix B). As can be seen from the table, this method is equally good as the method comparing theoretical and experimental curves. Since this method is less time consuming, this method was used to determine the liquid dispersion coefficients for experiments using distributor No. 2.

The liquid dispersion coefficients and their operating conditions are tabulated in Table 11. The axial dispersion coefficients were not reported for those experiments in the absence of gas flow. The tracer curves in the absence of gas flow approached plug-flow regime: the curves rose steeply and the peak times were very close to a dimensionless time of unity. However, tailing problems existed in the experimental tracer curves and could not be handled by the axial dispersion model.

As discussed in the Experimental Section, a constant solid concentration should be maintained in the column in order to determine the effects of solids on liquid dispersion. For practical purposes, slurry was not pumped into the column. Instead, a predetermined amount of sand was added to the column partially filled with liquid. Then the liquid dispersion experiments proceeded with the liquid flowing through the column at various rates. For large particles (20/30 mesh and 60/80 mesh), only slight amounts of solid were lost as entrained particles in the existing stream. However, for very fine particles

Table 11
Liquid Dispersion Experimental Results

Run no.	Liquid velocity (ft/sec)	Gas velocity (ft/sec)	Particle size (mesh)	Solids concentration (lb/ft ³)	dispersion ^a no.	Axial dispersion coefficients (ft ² /sec)
Distributor #1						
XIII - 1	0.0489	0.327	--	0.0	0.50	0.617
- 2	0.0505	0.194	--	0.0	0.39	0.497
- 3	0.0410	0.194	--	0.0	0.47	0.486
- 4	0.0302	0.194	--	0.0	0.65	0.495
- 5	0.0199	0.194	--	0.0	1.00	0.502
- 6	0.0099	0.194	--	0.0	2.00	0.500
- 7	0.0506	0.097	--	0.0	0.38	0.485
- 8	0.0499	0.050	--	0.0	0.32	0.403
- 9	0.0048	0.050	--	0.0	3.2	0.406
- 10	0.0503	0.0	--	0.0	Very low	Very low
- 11	0.0097	0.327	--	0.0	2.5	0.612
- 12	0.0190	0.327	20/30	5.0	1.00	0.479
- 13	0.0400	0.327	20/30	5.0	0.47	0.474
- 14	0.0598	0.327	20/30	5.0	0.32	0.482
- 15	0.0092	0.327	20/30	5.0	2.20	0.511
- 16	0.0295	0.327	20/30	5.0	0.65	0.483
XIV - 1	0.0536	0.050	20/30	5.0	0.22	0.297
- 2	0.0541	0.097	20/30	5.0	0.25	0.341
- 3	0.0535	0.194	20/30	5.0	0.27	0.364
- 4	0.0533	0.327	20/30	5.0	0.35	0.471
- 5	0.0534	0.0	20/30	5.0	Very low	Very low
XV - 1	0.0548	0.50	20/30	20.0	0.22	0.304
- 2	0.0532	0.097	20/30	20.0	0.22	0.295
- 3	0.0538	0.194	20/30	20.0	0.27	0.366
- 4	0.0497	0.327	20/30	20.0	0.30	0.376
- 5	0.0537	0.0	20/30	20.0	Very low	Very low
- 6	0.0404	0.327	20/30	20.0	0.35	0.356
- 7	0.0285	0.327	20/30	20.0	0.70	0.503
- 8	0.0255	0.327	20/30	20.0	0.60	0.385
- 9	0.0130	0.327	20/30	20.0	0.60	0.196

Table 11 (Continued)

Run no.	Liquid velocity (ft/sec)	Gas velocity (ft/sec)	Particle size (Mesh)	Solids concentration (lb/ft ³)	dispersion ^a no.	Axial dispersion coefficients (ft ² /sec)
XVI - 1	0.016	0.080	-140	2.06 to 19.33	0.9	0.33 to 0.38
- 2	0.043	0.327	100	4.88 to 9.47	0.5	0.53 to 0.55
XVII - 1	0.017	0.327	60/80	5.0	1.00	0.43
- 2	0.031	0.327	60/80	5.0	0.5	0.38
- 3	0.049	0.0	60/80	5.0	Very low	Very low
- 4	0.035	0.327	60/80	5.0	0.5	0.44
- 5	0.055	0.5	60/80	5.0	0.22	0.30
- 6	0.069	0.097	60/80	5.0	0.20	0.35
- 7	0.050	0.327	60/80	5.0	0.33	0.42
- 8	0.057	0.194	60/80	5.0	0.27	0.38
- 9	0.954	0.327	60/80	5.0	0.35	0.48
XVIII - 1	0.045	0.050	60/80	20.0	0.20	0.23
- 2	0.050	0.194	60/80	20.0	0.27	0.34
- 3	0.041	0.327	60/80	20.0	0.43	0.45
- 4	0.018	0.327	60/80	20.0	1.0	0.46
(b) XIX - 1	0.049	0.327	-140	5.0	0.35	0.43
Distributor #2						
XXV - 1	0.053	0.05	--	--	0.27	0.361
- 2	0.048	0.10	--	--	0.35	0.424
- 3	0.052	0.194	--	--	0.37	0.485
- 4	0.050	0.33	--	--	0.38	0.479
- 5	0.058	0.43	--	--	0.42	0.615
- 6	0.048	0.0	--	--	Very low	Very low
- 7	0.036	0.33	--	--	0.67	0.609
- 8	0.021	0.33	--	--	1.02	0.540

Table 11 (Continued)

Run no.	Liquid velocity (ft/sec)	Gas velocity (ft/sec)	Particle size (mesh)	Solids concentration (lb/ft ³)	dispersion no. ^a	Axial dispersion coefficients (ft ² /sec)
XXVI - 1	0.056	0.05	20/30	5.0	0.17	0.240
- 2	0.053	0.10	20/30	5.0	0.30	0.401
- 3	0.051	0.194	20/30	5.0	0.30	0.386
- 4	0.050	0.33	20/30	5.0	0.32	0.403
- 5	0.034	0.43	20/30	5.0	0.65	0.558
- 6	0.038	0.33	20/30	5.0	0.45	0.431
- 7	0.020	0.33	20/30	5.0	0.85	0.429
- 8	0.048	0.05	20/30	20.0	0.19	0.230
- 9	0.048	0.10	20/30	20.0	0.20	0.242
- 10	0.048	0.194	20/30	20.0	0.27	0.327
- 11	0.048	0.33	20/30	20.0	0.30	0.363
- 12	0.048	0.43	20/30	20.0	0.35	0.424
- 13	0.037	0.33	20/30	20.0	0.345	0.322
- 14	0.021	0.33	20/30	20.0	0.70	0.371
XXVII - 1 ^b	0.049	0.03	-140	3.74	0.33	0.408
XXVIII - 1 ^b	0.059	0.05	-140	3.74	0.16	0.238
- 2	0.058	0.10	-140	3.74	0.25	0.366
- 3	0.050	0.194	-140	3.74	0.35	0.442
- 4	0.047	0.43	-140	3.74	0.46	0.545
- 5	0.046	0.33	-140	3.74	0.36	0.418
- 6	0.025	0.33	-140	3.74	0.72	0.454

Table 11 (Continued)

Run no.	Liquid velocity ft/sec	Gas velocity ft/sec	Particle size mesh	Solids concentration lb/ft ³	dispersion ^a no.	Axial dispersion coefficients ft ² /sec
XXIX - 1 ^b	0.057	0.05	-140	14.94	0.19	0.273
- 2	0.050	0.33	-140	14.94	0.32	0.404
- 3	0.052	0.10	-140	14.94	0.33	0.433
- 4	0.051	0.194	-140	14.94	0.34	0.437
- 5	0.049	0.43	-140	14.94	0.33	0.408
- 6	0.038	0.33	-140	14.94	0.37	0.355
- 7	0.028	0.33	-140	14.94	0.48	0.39
No Distributor						
XX - 1	0.046	0.327	--	--	0.80	0.928
- 2	0.053	0.194	--	--	0.56	0.749
- 3	0.050	0.050	--	--	0.18	0.227
- 4	0.051	0.100	--	--	0.25	0.324
- 5	0.048	0.0	--	--	Very low	--
- 6	0.041	0.327	--	--	0.50	0.513
- 7	0.015	0.327	--	--	1.30	0.475
XXI - 1	0.049	0.194	20/30	5.0	0.38	0.472
XXI - 1R	0.053	0.194	20/30	5.0	0.33	0.437
- 2	0.027	0.327	20/30	5.0	0.62	0.418
- 3	0.055	0.050	20/30	5.0	0.25	0.348
- 4	0.021	0.194	20/30	5.0	0.88	0.466
XXII - 1 ^b	0.053	0.194	-140	5.0	0.30	0.404
- 2	0.054	0.05	-140	5.0	0.15	0.204
XII - 3 ^b	0.024	0.327	-140	5.0	0.62	0.369
- 4	0.023	0.194	-140	5.0	0.78	0.452

^a Dispersion number is the reverse of the Peclet number.

^b Continuous operation.

(-140 mesh and -100 mesh), large amounts of sand were carried out of the column by the liquid. This resulted in an enormous difference in solid concentration between the beginning and end of the run as shown in Runs XVI-1 and XVI-2 of Table 11. Therefore, averaging extreme concentration values was not justified for very fine particles. Hence, the axial dispersion coefficients for Runs XVI-1 and XVI-2 are presented as a range rather than as a single number. Since Run XIX-1 shown in Table 11 was carried out by pumping a continuous slurry (with a known solids concentration) into the column, we were able to independently (a) determine the liquid axial dispersion coefficient in the presence of fine particles at a uniform concentration of solids and (b) compare the effect of varying solids concentration on liquid axial dispersion coefficients. As discussed earlier, unusual behavior was observed in the absence of a distributor plate. The liquid level at the top of the column periodically surged so that the liquid level rose above the exit line and then suddenly drop a few inches below the exit line. Periodic gas slugs were usually observed in the column and were suspected to be the major cause of the surging behavior. Because of the surging behavior, the electronic noise in the experimental tracer curve increases at higher gas and liquid velocities reducing data accuracy. For example, Run numbers XX-1 and XX-2 shown in Table 11 were conducted under severe surging conditions. The dispersion coefficients obtained from these two experiments were completely out of line with the rest of data and were excluded in the following discussion.

5.2.2.1 Effect of Liquid Velocity

In general, liquid velocity had no effect on the liquid dispersion coefficient. In the presence of distributor #1, the results for two particle sizes (20/30 and 60/80 mesh) at three concentrations (0, 5, and 20 lb/ft³) are summarized in Table 12. The data were taken at a constant gas velocity of 0.327 ft/sec., while the liquid velocity ranged from 0.009 to 0.059 ft/sec. Without solids, a fivefold increase of liquid velocity from 0.009 to 0.049 ft/sec. did not change the value of the liquid dispersion coefficient. A similar lack of dependence

Table 12

Effect of Liquid Velocity on Axial Liquid Dispersion
Coefficient in the Presence of Distributor #1

Gas Velocity = 0.327 ft/sec

Liquid velocity (ft/sec)	No solids	Axial dispersion coefficient (ft ² /sec)			
		20/30 mesh		60/80 mesh	
		$C_S^a = 5.0$ lb/ft ³	$C_S = 20.0$ lb/ft ³	$C_S = 5.0$ lb/ft ³	$C_S = 20.0$ lb/ft ³
0.009	0.61	0.51			
0.013			0.20		
0.018				0.43	0.46
0.026			0.38		
0.030		0.48	0.50	0.38	
0.035				0.44	
0.040		0.47	0.36		0.45
0.049	0.62		0.38	0.42	
0.053		0.47		0.48	
0.059		0.48			

^a C_S = concentration of solids.

on liquid velocity was observed in the presence of solid particles. At a low solid concentration (5 lb/ft^3), the liquid dispersion coefficient was independent of liquid velocity for both 20/30 and 60/80 mesh particles. However, at high concentrations (20 lb/ft^3), the data for 20/30 mesh particles all scattered, at 0.013 and 0.030 ft/sec liquid velocity. At these conditions, discrepancies were also observed between the predicted and theoretical curves. These discrepancies are not clearly understood. Since the large particles (20/30 mesh) were not completely suspended at these operating conditions, the settled solids at high concentrations could be the cause of the observed discrepancies. In any event, such discrepancies were not observed for high concentrations of 60/80 mesh particles. In general, liquid velocity has no effect on the liquid dispersion coefficient in the presence of distributor #1.

Similarly the results for distributor #2 also show that the liquid dispersion coefficient is independent of liquid velocity (Table 13). Although larger data scattering was observed without solids present, there is no trend that indicates dependence on liquid velocity. Furthermore, the data obtained in the presence of solids clearly indicated that the liquid dispersion coefficient was independent of liquid velocity at all conditions.

Although limited data were available because of the surging behavior in the absence of a distributor plate, the results consistently indicated that liquid velocity did not effect the liquid dispersion coefficient. For example, in the absence of solids, increasing liquid velocity from 0.015 to 0.041 ft/sec resulted in an insignificant change in axial dispersion coefficients (0.475 to $0.513 \text{ ft}^2/\text{sec}$). Similarly, increasing liquid velocity from 0.021 to 0.053 ft/sec in the presence of 20/30 mesh and -140 mesh particles resulted in a negligible change in axial dispersion coefficient, from 0.466 to 0.472 and from 0.452 to $0.404 \text{ ft}^2/\text{sec}$, respectively. All these results show consistently that liquid velocity has no effect on axial dispersion coefficients.

Table 13

Effect of Liquid Velocity on Axial
Dispersion Coefficients in the Presence of Distributor #2

Gas Velocity = 0.327 ft/sec

Axial dispersion coefficients (ft ² /sec)					
V_L (ft/sec)	No solids	20/30 mesh		-140 mesh	
		$C_s = 5 \text{ lb/ft}^3$	$C_s = 20 \text{ lbs/ft}^3$	$C_s = 3.7 \text{ lb/ft}^3$	$C_s = 14.9 \text{ lb/ft}^3$
0.02	0.54	0.43	0.37	0.45	0.34
0.04	0.61	0.43	0.32	0.42	0.36
0.05	0.48	0.40	0.36	0.41	0.40

5.2.2.2 Effect of Gas Velocity

Increasing gas velocity increased the liquid backmixing as well as the liquid dispersion coefficient. The effects of gas velocity in the presence of distributors #1 and #2 at various solids concentrations are shown in a log-log plot of liquid dispersion coefficient gas superficial velocity in Figure 35. All the data at different combinations of distribution plate, solid concentration, and particle size consistently show that the dispersion coefficient increases with increasing gas velocity.

The data shown in Figure 35 can be separated into three different groups. The open circles and hexagons (above the shaped data points) represent the absence of solids. The rest of the open symbols that fall below the shaped data points represent results at high solid concentration. Although the data are scattered, the presence of solid particles clearly decreased the liquid dispersion coefficient. Furthermore the coefficient continues to decrease with increasing solid concentration. Within each solid concentration (0, 5 and 20 lb/ft³), the results suggest that the type of distributor and the particle size have no effect.

However, without a distributor plate no consistent pattern was observed as a function of gas velocity and solid particle size. Table 14 summarizes the effect of gas velocity and solids on axial dispersion coefficients from the 12-in.-diameter column in the absence of a distributor. In this table, wherever possible, dispersion coefficients were averaged over values obtained at different liquid velocities. When solids were absent, the axial dispersion coefficient increases with gas velocity, as expected. However, the presence of solids complicates the pattern. For both -140 and 20/30 mesh particles, the dispersion coefficient initially increases with gas velocity, but then decreases. Furthermore, at the highest gas velocity employed, the presence of solids decreased in the axial dispersion coefficients. This is in agreement with earlier results obtained in the presence of a distributor. However, at a gas velocity of 0.05 ft/sec, the presence

Figure 35

Liquid Axial Dispersion Coefficient

Table 14

Effect of Gas Velocity and Solids on Axial
Dispersion Coefficients in the Absence of a Distributor

Gas velocity (ft/sec)	Axial dispersion coefficients (ft ² /sec)		
	No solids	20/30 Mesh Solids	-140 Mesh Solids
0.05	0.277	0.348	0.204
0.10	0.324	--	--
0.194	--	0.458 ^a	0.428 ^a
0.327	0.494 ^a	0.418	0.369

^a Average values.

of -140-mesh solids decreased the value of the axial dispersion coefficient whereas the presence of 20/30 mesh particles resulted in an increase. These results indicate that the absence of a distributor plate results in some unusual patterns which may be attributed to the surging behavior discussed previously.

5.2.3 Nitrogen/Tetralin and Nitrogen/Tetralin/Sand Systems

Figures 36 to 47 plot the data from data from different dispersion runs in tetralin. Figure 36 shows two runs obtained in the 5-in.-diameter column using the batch method with the tracer injection at the bottom and the top of the column. The square data points were obtained from the samples collected at the top of the column with the dye tracer injected at the column bottom. The circle data points represented the results by reversing the sampling and injection location. As shown in Figure 36, the data obtained from the reversed injection are very close. This run was performed to ensure that the liquid dispersion coefficient (E_{zL}) was not affected by different geometries and inlet conditions. The solid line shows the fit of the axial dispersion model using the least-squares fit technique predicting a value of $0.093 \text{ ft}^2/\text{sec}$ for E_{zL} .

Figures 37 to 45 show how concentration varies with time for different gas velocities in the 12" column. The data were analyzed by both averaging and least-squares fit techniques described earlier. Comparison of the E_{zL} values obtained from these two methods was summarized in Table 15. The values are very close. Most of the differences between the values are less than 5%. This good comparison provides confidence in the data analyses to determine the E_{zL} values.

Table 15 shows an increase of E_{zL} with increasing gas velocity, though the change is quite insensitive between 0.133 and 0.216 ft/sec gas velocities (illustrated more clearly in Figure 45). In addition, the liquid dispersion coefficient was reduced in the presence of solid particles and with increasing solid concentration. These findings agree with the results from the air/water/sand system.

Figure 36

Comparison of Theoretical and Experimental Concentration

Figure 37

Comparison of Theoretical and Experimental Concentration

Figure 38

Comparison of Theoretical and Experimental Concentration

Figure 39
Comparison of Theoretical and Experimental Concentration

Figure 40
Comparison of Theoretical and Experimental Concentration

Figure 41

Comparison of Theoretical and Experimental Concentration

Figure 42

Comparison of Theoretical and Experimental Concentration

Figure 43

Comparison of Theoretical and Experimental Concentration

Figure 44

Comparison of Theoretical and Experimental Concentration

Figure 45

Effect of Gas Velocity on Liquid Dispersion Coefficient

Table 15

Comparison of E_{ZL} by Averaging and Least-Squares Fitting Technique

Column diameter (in.)	V_g (ft/sec)	E_{ZL} (least sq.)	E_{ZL} (averaged)	Remarks
5	0.05	0.0937	0.093	Sampled at D
5	0.05	0.0876	0.093	Sampled at A
12	0.043	0.4503	0.45	
12	0.133	0.6867	0.65	
12	0.176	0.6412	0.651	
12	0.216	0.6393	0.651	
12	0.267	0.6885	0.641	
12	0.356	0.739	--	Averaging was not done
12	0.392	0.8641	0.85	No solids
12	0.392	0.7805	0.79	#140-mesh sand $C_2 = 5 \text{ lb/ft}^3$
12	0.392	0.7231	0.70	#140-mesh sand $C_s = 22 \text{ lb/ft}^3$

Table 16 summarizes the comparison between the results of these tests and predictions from published correlations. The range of gas velocity simulates those designed for the SRC-I plant, i.e., 0.043-0.392 ft/sec. Each correlation indicates an increase in axial dispersion due to an increase in turbulent mixing resulting from gas agitation of the liquid. These correlations show that diameter, gas velocity, liquid viscosity, and liquid density could influence the values of the axial liquid dispersion coefficient. Of the six correlations compared in Table 16 those of Towel and Baird are the best.

Two additional experiments were run with tetralin using the continuous mode with two liquid velocities (Figures 46 and 47). The following data shows the differences between the continuous and batch modes:

V_g ft/sec	V_l ft/sec	E_{zL} (ft ² /sec)	
		Batch	Continuous
0.392	0.00	0.85	
0.392	0.02		1.02
0.392	0.05		1.33

The E_{zL} values for the continuous operation mode were determined by matching the peak time between the tracer curve and the model as described in the experimental section. Unlike the sodium chloride tracer test in the air/water system, the dye tracer study requires discrete samples collected at fixed interval of time. Usually, a flat region exists at the top of the graph which makes the resolution of which point is an assumed peak rather questionable. In fact the curve fits shown in Figures 46 and 47 were poor. Since limited runs in continuous operation mode can be performed with tetralin due to accumulation of dye concentration, effect of liquid flow was not studied at other gas velocity. At this point in time, little can be said about liquid velocity effect on E_{zL} in tetralin system. Since the air/water system had determined that there is practically no liquid flow effect on E_{zL} , it is unlikely that the physics will change in the tetralin system.

Table 16

Comparison of Experiments and Correlations
for Tetralin Liquid Dispersion Coefficient

Column Diameter (in.)	V_g (ft/sec)	E_{zL} Exp.	E_{zL} (Towel)	E_{zL} (Cova)	E_{zL} (Deckwer)	E_{zL} (Hikita)	E_{zL} (Baird)	E_{zL} (Ying)
5	0.05	0.093	0.074	0.176	0.081	0.155	0.128	0.088
12	0.043	0.45	0.255	0.167	0.268	0.447	0.390	0.264
12	0.133	0.65	0.449	0.240	0.370	0.627	0.565	0.397
12	0.176	0.65	0.516	0.263	0.403	0.701	0.620	0.439
12	0.216	0.65	0.572	0.281	0.428	0.767	0.664	0.472
12	0.276	0.65	0.636	0.300	0.456	0.846	0.712	0.516
12	0.392	0.85	0.770	0.340	0.512	1.028	0.808	0.585

Figure 46

Dimensionless Concentration Versus Dimensionless Time

Figure 47

Dimensionless Concentration Versus Dimensionless Time

Column Diameter = 12"

System: Tetralin

$V_g = .392$ ft/sec

$E_{ZL} = 1.33$ ft²/sec

$V_L = .05$ ft/sec

□ Experimental

— Dispersion Model

5.2.4 Nitrogen/Glycol-Water and Nitrogen/Glycol-Water/Sand Systems

Figures 48-57 graph the results of continuous experiments conducted with the glycol mixture to determine E_{zL} . Different liquid velocities at constant gas velocities were tested to determine the effect of liquid velocity. Table 17 lists the calculated E_{zL} numbers for the runs. Although the data were scattered, no definite trend of dependence on liquid velocity was observed.

The average value of E_{zL} over five-fold change in liquid velocity was plotted as a function of gas superficial velocity in Figure 58. The liquid dispersion coefficient increases with increasing gas velocity, in perfect agreement with the earlier observations in air/water and nitrogen/tetralin systems.

Table 18 compares experimental and calculated liquid dispersion coefficients. The data indicate that Towel's and Baird's correlations agree reasonably well with the averaged E_{zL} value. This agreement is consistent with the findings in nitrogen/tetralin system.

In the presence of solids, the E_{zL} value was similar to that found in two-phase system as shown in Table 17. However, such one-data-point comparison is probably inconclusive because of the large scatter in the two-phase system. Therefore, the true effect of solid particles on liquid dispersion coefficient will be based on the large data base from the air/water/sand and nitrogen/tetralin/sand systems.

5.3 Solids Dispersion

Solid dispersion experiments were performed in this study on a 5-inch and a 12-inch diameter column. The liquids used were water, tetralin and different concentration of ethylene glycol and water mixtures.

In the batch mode, gas was bubbled through the column, which was filled with liquid and a known weight of solid particles. During a 30-minute bubbling period at each gas velocity, steady-state conditions

Figure 48
Liquid Dispersion Coefficient

Figure 49

Liquid Dispersion Coefficient

Figure 50
Liquid Dispersion Coefficient

Figure 51
Liquid Dispersion Coefficient

Figure 52
Liquid Dispersion Coefficient

Figure 53
Liquid Dispersion Coefficient

Figure 54

Liquid Dispersion Coefficient

Figure 55
Liquid Dispersion Coefficient

Figure 56
Liquid Dispersion Coefficient

Figure 57

Liquid Dispersion Coefficient

Table 17

Axial Liquid Dispersion Values (E_{zL}) in
a 50% Glycol Mixture with a 12-in. Column

V_g ft/sec	V_L ft/sec	ft^2/sec	Average Value (ft^2/sec)	Remarks
0.05	0.00	0.381		No solids
0.05	0.01	0.479		No solids
0.05	0.03	0.434	0.496	No solids
0.05	0.05	0.575		No solids
0.10	0.00	0.568		No solids
0.197	0.00	0.607		No solids
0.197	0.01	0.618		No solids
0.197	0.03	0.618	0.57	No solids
0.197	0.05	0.475		No solids
0.40	0.00	0.717		No solids
0.40	0.01	0.841		No solids
0.40	0.03	1.1	0.891	No solids
0.40	0.05	0.915		No solids
0.40	0.05	0.915	---	#60/80-mesh sand used 16.5-lb/ft ³ average concentration

FIGURE 58
COMPARISON OF E_{ZL} FOR BATCH
AND CONTINUOUS EXPERIMENTS (50% GLYCOL)

Table 18

Liquid Dispersion Coefficients in
50% Glycol with a 12-in. Diameter Column

V_g (ft/sec)	Avg. exp.	Towel	Cova	Deckwer	Hikita	Baird	Ying
0.05	0.438	0.275	0.177	0.276	0.440	0.409	0.279
0.10	0.568	0.389	0.221	0.340	0.538	0.515	0.358
0.20	0.589	0.550	0.276	0.418	0.704	0.647	0.459
0.40	0.804	0.778	0.344	0.515	0.987	0.813	0.590

were established. Then slurry samples were withdrawn from sampling ports at various heights of the column to determine solids concentration profile. In continuous operation, slurry flowed continuously through the column, and samples were withdrawn from all ports at periodic intervals over a few hours until a steady-state condition was achieved. Samples from each port were usually obtained near the wall region and the center of the tube. The average of these two samples is the average concentration considered to exist at this axial position.

5.3.1 Theoretical Background

In a batch operation (with no liquid flowing), at any cross section of the column, the mass balance of solid particles at steady state conditions results in the following expression:

$$V_p C_s + E_{zp} \frac{dC_s}{dL} = 0 \quad (4)$$

where V_p = settling velocity of solid particles (ft/sec); C_s = concentration of solid particles in liquid (lb/ft³); E_{zp} = dispersion coefficient of solid particles (ft²/sec); and L = Distance from the bottom of the column (ft).

Equation 4 can be rewritten as:

$$\frac{d \ln C_s}{dL} = -V_p/E_{zp} \quad (5)$$

Therefore, a plot of $\ln C_s$ vs L should yield a straight line, provided that both V_p and E_{zp} are not functions of either solid concentration or column level. Figures 59 to 88 are semilogarithmic plots of C_s vs L as a function of gas velocity for all the experiments conducted in this research program. In general, the experimental points follow the straight-line relationship suggested by the theory.

Figure 59

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 5" COLUMN USING 140/170 MESH GLASS BEADS
AT LOW CONCENTRATION**

FIGURE 60

SOLIDS CONCENTRATION vs COLUMN HEIGHT FOR THE 5" COLUMN USING 140/170 MESH GLASS BEADS AT HIGH CONCENTRATION

Figure 61

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 5" COLUMN USING -140 MESH SAND
AT LOW CONCENTRATION**

Figure 62

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 5" COLUMN USING -140 MESH SAND
AT HIGH CONCENTRATION**

Figure 63

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 12" COLUMN USING -140 MESH SAND
AT LOW CONCENTRATION**

Figure 64

SOLIDS CONCENTRATION vs COLUMN HEIGHT FOR THE 12" COLUMN USING -140 MESH SAND AT HIGH CONCENTRATION

Figure 65

Concentration Versus Length in a 5-Inch Column

Figure 66

Concentration Versus Length in a 5-Inch Column

Figure 67

Concentration Versus Length in a 5-Inch Column

Figure 68

Concentration Versus Length in a 12-Inch Column

Figure 69

Concentration Versus Length in a 12-Inch Column

Figure 70

Solids Distribution Concentration Versus Axial Length

System: 90% Glycol

P_s = 140 minus mesh sand

12" Column

Figure 71

Solids Distribution Concentration Versus Axial Length

Figure 72

Concentration Versus Length in a 12-Inch Column

Figure 73

Concentration Versus Length in a 12-Inch Column

System: 50% Glycol

$P_s = 140$ minus mesh sand

Figure 74

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 5" COLUMN USING 60/70 MESH GLASS BEADS
AT LOW CONCENTRATION**

Figure 75

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 5" COLUMN USING 60/70 MESH GLASS BEADS
AT HIGH CONCENTRATION**

Figure 76

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 5" COLUMN USING 60/70 MESH SAND
AT LOW CONCENTRATION**

Figure 77

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 5" COLUMN USING 60/70 MESH SAND
AT HIGH CONCENTRATION**

Figure 78

**SOLIDS CONCENTRATION vs COLUMN HEIGHT
FOR THE 12" COLUMN USING 60/80 MESH SAND
AT LOW CONCENTRATION**

Figure 79

SOLIDS CONCENTRATION vs COLUMN HEIGHT FOR THE 12" COLUMN USING 60/80 MESH SAND AT HIGH CONCENTRATION

Figure 80

Concentration Versus Length in a 5-Inch Column

Figure 81
Concentration Versus Length in a 5-Inch Column

Figure 82

Concentration Versus Length in a 5-Inch Column

Figure 83

Concentration Versus Length in a 12-Inch Column

System: Tetralin

$P_s = 60/80$ mesh

$C_s = 6.65 \text{ lb/ft}^3$

□ $V_g = .133 \text{ ft/sec}$

○ $V_g = .216 \text{ ft/sec}$

▲ $V_g = .308 \text{ ft/sec}$

+ $V_g = .356 \text{ ft/sec}$

* $V_g = .392 \text{ ft/sec}$

Figure 84
Concentration Versus Length in a 12-Inch Column

Figure 85

Solids Distribution Concentration versus Axial Length

Figure 86

Solids Distribution Concentration Versus Axial Length

Figure 87

Concentration Versus Length in a 12-Inch Column

Figure 88
Concentration Versus Length in a 12-Inch Column

Tables 19 to 21 list batch experiments that were conducted using water, tetralin and glycol. The range of gas velocities and solid concentrations included those of expected in the SRC-I dissolver.

5.3.2 Fine particles (Batch, 140/170 Mesh Glass Beads and -140 Mesh Sand)

Figures 59 to 73 graph the axial distribution of fine particles in the 5-in. and 12-in. columns. In general all these profiles are flat, with very small slopes, which is expected because the terminal velocity of small particles is usually small. In fact, as the particles become smaller, the interfacial drag per unit volume usually increases, thus decreasing the terminal velocity. Particles of this size, therefore, will stay well-mixed in the dissolver under a wide range of gas velocities.

In addition, the distribution of solids concentration was independent of gas velocity which implies that the solid dispersion coefficient is also independent of gas velocity. However, if the effective V_p in Equation 4 varies with gas velocity, the solid dispersion coefficient will be proportionally dependent on gas velocity.

The V_p/E_{zp} values measured from the slopes of the straight line plots shown in Figure 59 through 73 are summarized in Tables 22 through 25. The results obtained from the air/water/solid system (Table 22) showed that the glass beads had consistently higher V_p/E_{zp} values than sand for both concentrations; the V_p/E_{zp} value for sand is about 10-15% of that for glass beads. This difference can be attributed to a higher settling velocity for glass beads because the average particle size for the 140/170 mesh glass beads were at least a factor of two larger than that of the 140 mesh minus sand. Based upon Stoke's Law, the V_p for glass beads would be at least a factor of four higher than the sand. Therefore, the V_p/E_{zp} values for sand is expected to be at least less than 25% of the value for glass beads for constant E_{zp} . This will explain the difference in V_p/E_{zp} values observed between glass beads and sand particles.

Table 19
List of Batch Experiments

Liquid Phase-Water

Table No.	Column diameter, (in.)	Type of solid particles	Particle size (mesh)	Solid concn ₃ (lb/ft ³)	Range of gas velocities, (ft/sec)
B-1	5	Glass beads	60/70	7.4	0.05-0.43
B-2	5	Glass beads	60/70	28.2	0.05-0.43
B-3	5	Glass beads	140/170	7.6	0.05-0.33
B-4	5	Glass beads	140/170	28.6	0.05-0.33
B-5	5	Sand	60/70	7.6	0.05-0.43
B-6	5	Sand	60/70	29.4	0.05-0.43
B-7	5	Silica	-140	7.6	0.05-0.43
B-8	5	Silica	-140	29.4	0.05-0.43
B-9	12	Sand	60/80	6.2	0.05-0.37
B-10	12	Sand	60/80	25.2	0.05-0.37
B-11	12	Silica	-140	6.2	0.05-0.43
B-12	12	Silica	-140	25.2	0.05-0.43

Table 20

List of Batch Experiments

Liquid Phase-Tetralin

Column diameter, (in.)	Type of solid particles	Particle size (mesh)	Solid Conc. (lb/ft ³)	Range of gas velocities (ft/sec)
5	Glass beads	140/170	8.32	0.1-0.40
5	Glass beads	140/170	25.1	0.1-0.40
5	Glass beads	60/70	8.28	0.1-0.40
5	Glass beads	60/70	26.41	0.1-0.40
5	Sand	-140	8.28	0.1-0.4
5	Sand	-140	25.74	0.1-0.4
5	Sand	60/80	8.295	0.1-0.4
5	Sand	60/80	26.11	0.1-0.4
12	Sand	-140	6.19	0.133-0.392
12	Sand	-140	22.01	0.133-0.392
12	Sand	60/80	6.65	0.133-0.392
12	Sand	60/80	23.66	0.133-0.392

Table 21
List of Batch Experiments
Liquid Phase-Glycol/Water Mixture

Column diameter (in.)	Type of solid particles	Particle size (mesh)	Solid concn (lb/ft ³)	Mixture type	Range of gas velocities, (ft/sec)
12	Sand	-140	20.75-22.93	90% glycol by wt	0.043-0.392
12	Sand	-140	21.43-24.3	70% glycol by wt	0.043-0.392
12	Sand	-140	6.66-21.68	50% glycol by wt	0.05-0.4
12	Sand	60/80	22.885-23.079	90% glycol by wt.	0.043-0.392
12	Sand	60/80	17.39-23.17	70% glycol by wt.	0.043-0.392
12	Sand	60/80	4.32-28.35	50% glycol by wt.	0.05-0.4

Table 22

Summary of V_p/E_{zp} for Fine Particles as a Function of Gas Velocity

Particle Size = 140/170 Mesh for Glass Beads Experiments

Particle Size = -140 Mesh for Sand Experiments (Water)

Gas velocity ft/sec	$V_p/E_{zp} \text{ (ft}^{-1}\text{)}$					
	$C_s = 7.5 \text{ lb/ft}^3$			$C_s = 28.6 \text{ lb/ft}^3$		
	5" Column		12" Column	5" Column		12" Column
	Glass Beads	Sand	Sand	Glass Beads	Sand	Sand
-161	0.10	0.307	0.029		0.013	
	0.15	0.312	0.018	0.156	0.014	0.006
	0.20	0.292	0.047	0.157	0.016	0.005
	0.24		0.019			0.007
	0.28		0.016			0.005
	0.30	0.034				
	0.33	0.330	0.017	0.162	0.019	0.006
	0.37		0.015			0.006
	0.43	0.275	0.073	0.148		0.006

Table 23

Summary of V_p/E_{zp} for Fine Particles as a
Function of Column Diameter and Gas Velocity

Particle Size = 140/170 Mesh for Glass Beads

Particle Size = -140-Mesh for Sand (Tetralin)

Gas velocity ft/sec	V_p/E_{zp} (ft ⁻¹)					
	$C_2 = 6.19-8.32$ lb/ft ³			$C_s = 22.01-25.74$ lb/ft ³		
	5-in. column		12-in. column	5-in. column		12-in. column
	Glass beads	Sand	Sand	Glass beads	Sand	Sand
0.100	0.147	0.29		0.101	0.008	
0.133			0.0123			0.0038
0.150	0.182	0.64		0.115	0.032	
0.200	0.203	0.67		0.117	0.023	
0.216			0.0150			0.0052
0.250	0.179	0.48		0.114	0.032	
0.308			0.0180			0.0045
0.330	0.183	0.70		0.109	0.028	
0.365			0.0172			0.0049
0.392			0.0160			0.0031
0.400	0.172	0.42		0.121	0.030	

Table 24

Values of V_p/E_{zp} for -140-Mesh Sand in Glycol/Water Mixtures
(12-in. diameter column)

Gas velocity (ft/sec)	V_p/E_{zp} (ft^{-1}) (90% glycol)	V_p/E_{zp} (ft^{-1}) (70% Glycol)
0.216	0.0043	0.0050
0.392	0.0040	0.0046
0.043	0.0030	0.0044
Viscosity (cp)	13.08	6.2

Table 25

Summary of V_p/E_{zp} for Fine Particles as a Function of Gas Velocity
in a 12-in. diameter column

Particle Size = -140 Mesh

Liquid = Glycol Mixture (50% by Wt)

Gas velocity (ft/sec)	C_s (lb/ft ³)	V_p/E_{zp} (ft ⁻¹)
0.2	6.57	0.0210
	18.37	0.0096
	20.76	0.0065
0.4	6.57	0.0161
	18.37	0.0107
	20.76	0.0071
0.5	6.57	0.0163

Effect of column diameter and solid concentration on V_p/E_{zp} are quite obvious as shown in Table 22. V_p/E_{zp} decreased with both increasing column diameter and solids concentration. The dependence on column diameter is not surprising because the degree of liquid backmixing and liquid dispersion coefficient increased with increasing column diameter, hence the solid dispersion coefficient will be expected to increase if there is any column diameter effect. The observed decrease in V_p/E_{zp} value with increasing column diameter directly reflects an increase in E_{zp} which is in qualitative agreement with the above expectation. The reason for the dependence on solid concentration is not clear, however. It is speculated that the particle settling velocity is hindered by the neighbor particles. This will qualitatively explain the decrease in V_p/E_{zp} with increased solids loading.

The fine particles behave similarly in tetralin (Table 23) and glycol-water mixtures (Tables 24 and 25) as in the aqueous system (Table 22). The V_p/E_{zp} values were consistently higher when glass beads were used instead of the fine silica (sand) particles. The V_p/E_{zp} ratio decreased as the sand concentration and the column diameter were increased. Consistently the V_p/E_{zp} ratio was found to be independent of gas velocity in all three system.

In the 5-in. column with glass beads, the values of V_p/E_{zp} decreased by 30 to 40% when water was substituted by tetralin at the same velocities. This result shows the effect of viscosity on solids distribution and V_p/E_{zp} . As viscosity increases, the solid settling velocity (V_p) decreases, thereby decreasing the V_p/E_{zp} values.

However, the viscosity effect is less distinct with the fine sand particles. For an order of magnitude increase in viscosity from 1 cp (pure water) to 13 cp (90% glycol solution), only a two-fold difference in V_p/E_{zp} was measured. For this fine particle size sand, the concentration profiles were almost completely flat. The value of V_p/E_{zp} measured from the slope of a flat curve is a very small quantity. Therefore the viscosity effect is less apparent. It is also likely that E_{zp} decreases with increasing viscosity, further dampening the viscosity effect on V_p/E_{zp} ratio.

5.3.3 Large Particles (Batch, 60/70-Mesh Glass Beads and 60/80 Mesh Sand)

Marked differences are seen in the behavior of large particles compared to fine particles. Significant gradients in the solids concentration were measured for both sand and glass beads. Complete suspension of these large particles could not be achieved at low gas velocities. Sampling from the bottom port is extremely difficult and the reliability of this sample is questionable. Hence, data from the lowest sampling port were excluded from the analysis.

The V_p/E_{zp} values for these large particles measured from the slopes of the plots shown in Figures 74-88 at different solids concentration and column diameter are summarized as a function of gas velocity in Tables 26-28. V_p/E_{zp} increases markedly in comparison with the case of fine particles primarily because the particle terminal velocity increases due to an increase in particle size.

In most of these experiments, complete suspension was achieved only at higher gas velocities. The critical gas velocity, which is defined as the velocity above which all particles are in complete suspension, for the 60/80 mesh sand particles in water was determined to be between 0.193 and 0.217 ft/sec. The results shown in Table 26 indicated that above the critical gas velocity, all the V_p/E_{zp} values showed no systematic change with gas velocity. With the limited amount of data available, the V_p/E_{zp} values seemed to be independent of gas velocity which is consistent with the behavior observed with the fine particles.

The results with water (Table 26) and tetralin (Table 27) as the liquid phase show that increasing the column diameter resulted in a decrease in V_p/E_{zp} values. This means an increase in solid dispersion coefficient with increasing column diameter, and is consistent with the observation using fine particles. Furthermore, the results also indicated that the V_p/E_{zp} values for the sand particles are slightly less than that for the glass beads, although the same size range of particles, namely 60/70 mesh, were used. This slight difference in

Table 26

Summary of V_p/E_{zp} for Fine Particles as a Function of Gas Velocity

Particle Size = 60/70 Mesh for 5-in.-Column Experiments

Particle Size = 60/80 Mesh 12-in.-Column Experiments

Gas velocity ft./sec	V_p/E_{zp}					
	$C_s = 7.5 \text{ lb/ft}^3$			$C_s = 28.6 \text{ lb/ft}^3$		
	5-in. Column		12-in. Column	5-in. Column		12-in. Column
	Glass beads	Sand	Sand	Glass Beads	Sand	Sand
-167-	0.10					
	0.15	0.959	0.174	0.117		
	0.20	0.916	0.757	1.115	0.789	0.193
	0.24		0.181			0.179
	0.28		0.205			0.152
	0.30		0.740	0.828		
	0.33	0.864	0.662	0.982	0.979	0.154
	0.37					0.154
	0.43	0.846	0.729	0.940		0.946

Table 27

Summary of V_p/E_{zp} for Fine Particles as a
Function of Column Diameter and Gas Velocity

Particle Size = 60/80 Mesh for Glass Beads

Particle Size = 60/80 for Sand (Tetralin)

Gas Velocity ft/sec	V_p/E_{zp}					
	$C_2 = 6.65 - 8.295 \text{ lb/ft}^3$			$C_s = 23.66 - 26.4 \text{ lb/ft}^3$		
	5" Column	12" Column		5" Column	12" Column	
	<u>Glass Beads</u>	<u>Sand</u>	<u>Sand</u>	<u>Glass Beads</u>	<u>Sand</u>	<u>Sand</u>
0.100	0.541	0.369		0.77	0.216	
0.133			0.128			0.120
0.150	0.573	0.455		0.468	0.466	
0.200	0.604	0.479		0.431	0.605	
0.216			0.118			0.104
0.250	0.641	0.488		0.514	0.584	
0.308			0.115			0.109
0.330	0.652	0.484		0.634	0.580	
0.365			0.131			0.0956
0.392			0.127			
0.400	0.691	0.484		0.465	0.497	0.9961

Table 28a

V_p/E_{zp} for 90% and 70% Glycol in 12-in.

Column Using 60/80-Mesh Sand

Gas velocity (ft/sec)	90% glycol	70% glycol
0.216	0.0216	0.0690
0.392	0.0220	0.0699
0.043	0.0373	0.1100
Viscosity (cp)	13.08	6.2

Table 28b

Summary of V_p/E_{zp} for Large Particles as a Function of Gas Velocity

Particle Size = 60/80 Mesh

Liquid = Glycol Mixture, 50% by wt.

12" Column		
Gas velocity (ft/sec)	C_s (lb/ft ³)	V_p/E_{zp} 12-in. column
0.05	1.41	
	11.81	0.077
0.20	6.64	0.113
	18.05	0.101
0.40	6.64	0.118
	18.05	0.080

V_p/E_{zp} values for sand and glass beads possibly reflected some intrinsic difference in the distribution of these particles within the 60/70 mesh range and effect of particle shape.

V_p/E_{zp} ratio decreases as the liquid viscosity increases. The results from Table 26-28 are summarized in Table 29 to illustrate the effect of viscosity. Within the range of gas velocity and solids concentration in this study, the V_p/E_{zp} decreases with increasing liquid viscosity. This decrease in V_p/E_{zp} is primarily due to the reduction of solid terminal velocity with increasing viscosity. This finding agrees with the observation of the fine particles.

The straight line plots shown in Figures 59-88 and the good agreement in the V_p/E_{zp} behavior with the wide range of parameters strongly support the validity of this one-dimensional mass-balance model (Equation 4) to describe the behavior of solids in a three-phase upflow column in a batch mode.

5.3.4 Continuous-Flow Solids Concentration

Continuous runs were conducted to observe the effect of gas velocity, liquid velocity, feed slurry concentration, particle size, column diameter, and inlet distributor on the behavior of suspended solids in the dissolver. Tables 30 to 32 list the experimental runs in three liquids (water, tetralin, and 50% glycol-50% water solution). In these experiments, solid/liquid slurry was fed continuously to the column at the bottom. Concentrations were then measured from different sample ports at 30-minute intervals in order to determine when steady state had occurred after which it allowed us to average the data points. Experimental data from a typical run are shown in Figure 89 which plots the concentration distribution as a function of time. Usually steady state was achieved 2 hours after the experiment started.

Two types of continuous operations were conducted: once-through and recycle mode which were described in the experimental section. The once-through mode, which required considerable sand and water to make

Table 29.

Summary of Liquid Property Effect on V_p/E_{zp} for
Large Particles (60/80 Mesh Sand) in a 12-in. Diameter Column

	Water	Tetralin	70% Glycol	90% Glycol
V_p/E_{zp}				
(ft ⁻¹)	0.145-0.195	0.092-0.128	0.069-0.110	0.022-0.037
Viscosity (cp)	1.0	2.7	6.2	13.1

Solids concentration 23-28 lb/ft³

Gas velocity 0.10-0.43 ft/sec

Table 30

List of Continuous Experiments

(Water System)

Column diameter (in.)	Particle mesh size	Liquid velocity ft/sec	Gas velocity ft/sec	Solids conc. lb/ft ³	Distributor type
12	60/80	0.05	0.43	6.17	II
12	60/80	0.05	0.19	7.27	II
12	60/80	0.05	0.29	7.37	II
12	60/80	0.05	0.33	7.81	II
12	60/80	0.03	0.33	6.84	II
12	60/80	0.01	0.33	8.22	II
12	60/80	0.05	0.43	8.35	II
12	60/80	0.01	0.33	31.16	II
12	60/80	0.05	0.33	23.73	II
12	60/80	0.03	0.37	25.49	II
12	-140	0.05	0.43	12.01	II
12	-140	0.05	0.15	14.84	II
12	-140	0.05	0.33	12.11	II
12	-140	0.03	0.33	12.21	II
12	-140	0.01	0.33	13.15	II

Table 30 (Continued)

List of Continuous Experiments

Column diameter (in.)	Particle mesh size	Liquid velocity ft/sec	Gas velocity ft/sec	Solids conc. lb/ft ³	Distributor type
12	60/80	0.05	0.33	23.28	I
12	60/80	0.05	0.15	26.40	I
12	60/80	0.01	0.33	29.42	I
12	60/80	0.03	0.33	29.01	I
12	60/80	0.05	0.43	27.94	I
12	60/80	0.05	0.25	23.35	I
12	-140	0.05	0.43	4.18	I
12	-140	0.05	0.10	4.82	I
12	-140	0.04	0.33	4.44	I
12	-140	0.05	0.19	4.59	I
12	-140	0.02	0.33	4.58	I
5	60/80	0.05	0.40	4.8	
5	60/80	0.05	0.40	4.8	
5	-140	0.05	0.33	4.8	(a)
5	-140	0.05	0.33	4.8	(b)

(a) recycle

(b) continuous feed

Table 31

List of Continuous Experiments

(Tetralin System)

Column diameter (in.)	Particle mesh size	Liquid velocity ft/sec	Gas velocity ft/sec	Solids conc. lb/ft ³	Distributor type
12	60/80	0.02	0.39	19.66	I
12	60/80	0.03	0.39	19.19	I
12	60/80	0.04	0.39	20.14	I
12	60/80	0.05	0.39	19.73	I
12	60/80	0.05	0.31	18.64	I
12	60/80	0.05	0.22	18.66	I
12	60/80	0.05	0.13	18.38	I
12	-140	0.02	0.39	15.09	I
12	-140	0.03	0.39	14.15	I
12	-140	0.04	0.39	14.55	I
12	-140	0.05	0.39	13.23	I
12	-140	0.05	0.31	13.63	I
12	-140	0.05	0.22	13.86	I

Table 32

List of Continuous Experiments

(50% Glycol System)

Column diameter (in.)	Particle mesh size	Liquid velocity ft/sec	Gas velocity ft/sec	Solids conc. lb/ft ³	Distributor type
12	60/80	0.18	0.04	21.99	I
12	60/80	0.03	0.05	21.43	I
12	60/80	0.05	0.05	18.12	I
12	60/80	0.01	0.05	15.84	I
12	-140	0.03	0.05	15.32	I
12	-140	0.05	0.05	11.06	I
12	60/80	0.01	0.02	21.33	I
12	60/80	0.03	0.20	19.56	I
12	60/80	0.05	0.19	16.76	I
12	-140	0.01	0.02	14.18	I
12	-140	0.03	0.02	14.57	I
12	-140	0.05	0.02	13.42	I
12	60/80	0.01	0.40	21.24	I
12	60/80	0.03	0.40	19.05	I
12	60/80	0.05	0.40	16.08	I
12	-140	0.01	0.40	13.77	I
12	-140	0.03	0.40	13.44	I
12	-140	0.05	0.40	12.37	I

I - Distributor plate no. 1 with bubble cap.

II - Distributor plate no. 2 with no bubble cap.

Figure 89. TYPICAL EXPERIMENTAL DATA FOR AXIAL SOLIDS DISTRIBUTION

fresh batches of slurry, was very inefficient. The recycle mode, which was used in this work, was more practical. These two modes were compared to determine any appreciable differences.

Figure 90 plots a run from each mode, and Figure 91 plots the non-dimensionalized concentration values, with respect to feed concentration. Although the feed concentrations in both runs are about 10% different, the profile did not seem to vary appreciably. This reinforces the belief that concentration distributions between continuous and recycle runs can be the same if all other run conditions are the same.

The effect of gas velocity on the distribution of fine particles (-140 mesh sand) in all three liquids (water, tetralin, and 50% glycol solution) are displayed in Figures 92-97. Figure 92 shows that at a constant sand/water slurry velocity of 0.05 ft/sec, increasing the gas velocity from 0.10 to 0.43 ft/sec resulted in very little change in solids distribution across the length of the column. A similar result was observed at a higher feed solids concentration as shown in Figure 93. An increase in gas velocity from 0.33 to 0.43 ft/sec resulted in an identical profile. At a gas velocity of 0.15 ft/sec, because of the difference in feed concentration, the profile was different than those at higher gas velocities. But the slopes are identical at all three gas velocities indicating that changes in gas velocity do not appreciably affect the axial solids distribution. Similar results were found in tetralin and 50% glycol solution as shown in Figures 94-97. These results confirm earlier findings from batch experiments (no liquid flow) that gas velocity has practically no effect on the distribution of the fine particles.

The effects of gas velocity on the distribution of large particles (60/80 mesh sand) in all three liquids are shown in Figures 98-104. In general the results are similar to those observed with fine particles; gas velocity has little effect on solid distribution provided that the gas velocity is higher than the critical gas velocity which is defined as the minimum gas velocity necessary to suspend the solid particles. Earlier we had shown that for fine

Figure 90

COMPARISON OF CONTINUOUS FEEDING AND RECYCLE RUNS

FIGURE 91

COMPARISON OF CONTINUOUS FEEDING
AND RECYCLE RUNS
USING NORMALIZED
SOLIDS CONCENTRATION

Figure 92

EFFECT OF GAS VELOCITY ON SOLIDS DISTRIBUTION

Figure 93.

EFFECT OF GAS VELOCITY ON THE DISTRIBUTION OF FINE PARTICLES

Figure 94

Concentration Versus Length in a 12-Inch Column

Figure 95
Solids Distribution Profile

Figure 96

Solids Distribution Profile

System: 50% Glycol

$P_s = 140$ minus mesh

12" Column

$V_L = .03$ ft/sec

□ $V_g = .05$ ft/sec

○ $V_g = .197$ ft/sec

▲ $V_g = .4$ ft/sec

Figure 97
Solids Distribution Profile

Figure 98.

EFFECT OF GAS VELOCITY ON SOLIDS DISTRIBUTION

COLUMN DIAMETER = 5 INCH

PARTICLE SIZE = 60/80 MESH

LIQUID VELOCITY = 0.05 FT/SEC

Figure 99.

EFFECT OF GAS VELOCITY ON THE DISTRIBUTION OF LARGE PARTICLES

Figure 100.

EFFECT OF GAS VELOCITY ON AXIAL SOLIDS DISTRIBUTION

Figure 101

Concentration Versus Length in a 12-Inch Column

Figure 102

Solids Distribution Concentration Versus Axial Length

Figure 103

Solids Distribution Concentration Versus Axial Length

Figure 104

Solids Distribution Concentration Versus Axial Length

particles (-140 mesh) changes in the gas velocity did not significantly affect the solids concentration profiles. All the velocities with the -140 mesh solids were above the critical gas velocities for these solids.

Figure 98 plots the data on 60/80 mesh particles in a 5-in. diameter column. Although increasing gas velocity from 0.05 to 0.194 ft/sec did not change the solids accumulation, increasing up to 0.327 ft/sec resulted in a dramatic decrease. Since the critical gas velocity for 60/80 mesh particles is around 0.2 ft/sec, at the higher 0.327 ft/sec velocity no settled solids will be present in the column thereby sharply decreasing the solids concentration profile.

The lack of independence on gas velocity can be clearly illustrated in Figure 99 and 100. Figure 99 plots the distribution of the 60/80 mesh particles in a 12-in. diameter column. Although the critical gas velocity of these particles in the 12-in. diameter column was not measured, it is believed to be less than that determined in the 5-in. diameter column (0.2 ft/sec) because of the increased turbulence in the larger diameter column as discussed earlier in the liquid dispersion section. Therefore Figure 99 shows that for 60/80 mesh particles, increasing gas velocity from 0.194 to 0.33 ft/sec did not change the solids concentration in the bulk of the column though a slight difference at the bottom of the column was observed. This slight difference at the column bottom could be due to sampling errors at this very high solids concentration region. Nevertheless, the results shown in Figure 99 illustrate the independence of gas velocity on solid accumulation beyond the critical gas velocity. Likewise gas velocity does not affect the distribution of the 60/80 mesh particles in tetralin and 50% glycol solution as shown in Figures 101-104.

In addition, Figure 105 graphs the ratio of average column solids concentration/feed concentration as a function of gas velocity. The invariance found over a four-fold increase in gas velocity for the fine particles (-140 mesh) strongly indicates the lack of any

Figure 105
Effect of Liquid Velocity on Solids Distribution

dependence on gas velocity. One may note the slightly higher value at 0.4 ft/sec for the 60/80 mesh particles. However, the difference is so small that the effect of gas velocity on the 60/80 mesh particle concentration will be minimal.

Liquid velocity has a more pronounced effect on the solids distribution profiles. The steepness of the profile decreases with an increase in liquid velocity. As liquid velocity increases, the convective force increases and ultimately creates a homogeneous mixture. This also means that net solids accumulation in the column will decrease.

Figures 106-109 show the effect of liquid velocity on the distribution of large particles (60/80 mesh sand). A general trend was observed: more solids were retained in the column as liquid velocity decreased. The concentration profile in the 5-in. diameter column rose dramatically as the liquid velocity was reduced from 0.05 to 0.01 ft/sec as shown in Figure 106. The effect of liquid flow was less dramatic though persistent in the 12-in. diameter column as illustrated in Figures 107-109 because of the enhanced turbulence in a larger diameter column.

It is interesting to note that the exit concentrations (Figures 106-109) varied only slightly with liquid velocity. This slight variation in exit concentration was due to the experimental technique employed in this study. The exit concentration should be identical to feed concentrations at steady-state conditions. For all the recycle experiments a known amount of material was used to make up the slurry inventory which was circulated through the column until steady state was achieved. Hence, the steady state feed concentration would be higher for those experiments with lower solids accumulation in the column. The direction of variation in exit concentration supports the effect of liquid velocity discussed above; the lowest liquid flow rate having the least feed concentration had the highest retained solid concentration which strongly supported the results of increasing solid accumulation with decreasing liquid velocity.

Figure 106.

EFFECT OF LIQUID VELOCITY ON SOLIDS DISTRIBUTION

COLUMN DIAMETER = 5 INCH

PARTICLE SIZE = 60/80 MESH

GAS VELOCITY = 0.327 FT/SEC

Figure 107. **EFFECT OF LIQUID VELOCITY
ON AXIAL SOLIDS DISTRIBUTION**

Figure 108. EFFECT OF LIQUID VELOCITY ON THE DISTRIBUTION OF LARGE PARTICLES

Figure 109

Concentration Versus Length in a 12 Inch Column

Under the conditions tested in this study, a linear relationship exists between liquid velocity and solids accumulation as shown in Figures 110-112. To eliminate the slight variations in exit concentration, Figures 110-112 plot the ratio of average column solids concentration/exit concentration as a function of liquid velocity. The average column solids concentration was determined by integrating the concentration profiles shown in Figures 106-109. These graphs clearly illustrate the effect of liquid velocity on solids accumulation. Less solids accumulation was found in tetralin (Figure 112) than in water primarily because of the viscosity effect. Particle terminal velocity is lower in a more viscous liquid (tetralin vs. water) and can be carried through the column easier by the bulk flow. Therefore, less accumulation was found in tetralin.

A similar effect of liquid velocity was found on the distribution and accumulation of fine particles (-140 mesh) as shown in Figures 113-115. As liquid velocity increases, the accumulation of the fine particles decreased. But the effect was much less dramatic than that observed with the 60/80 mesh particles because the fine particles were more homogeneously mixed. In addition, a linear relationship between liquid velocity and solids accumulation was also observed as shown in Figure 116.

Following the above discussion, the effect of particle size on solids distribution is quite obvious. Figures 117 and 118 graph the comparison of the concentration profiles between -140 mesh and 60/80 mesh sand. With the larger particles at identical flow conditions, more solids accumulate. This effect is expected because the terminal velocity of larger particles would make the profile steeper assuming that the particle size does not have a strong effect on the axial solids dispersion coefficient.

The hydrodynamic effect in the absence of a distributor was evaluated using air/water/-140 mesh sand in a 12-in. diameter column. For fine particles at low concentrations the solids concentration decreases almost linearly with column height as shown in Figure 119 for both

Figure 110. **EFFECT OF LIQUID VELOCITY
ON SOLIDS ACCUMULATION**

Figure 111.

EFFECT OF LIQUID VELOCITY ON SOLIDS RETENTION

Figure 112

Effect of Liquid Velocity on Solids Accumulation

Figure 113. EFFECT OF LIQUID VELOCITY ON THE DISTRIBUTION OF FINE PARTICLES

Figure 114

EFFECT OF LIQUID VELOCITY ON SOLID CONC.

Figure 115

Concentration Versus Length in a 12 Inch Column

Figure 116.

EFFECT OF LIQUID VELOCITY ON SOLIDS ACCUMULATION

Figure 117. **EFFECT OF PARTICLE
SIZE ON SOLIDS DISTRIBUTION**

Figure 118
Effect of Particle Size

Figure 119.

EFFECT OF DISTRIBUTOR ON SOLIDS DISTRIBUTION

cases (presence or absence of a distributor). The differences in solid concentrations between the top and bottom of the column are fairly small in both cases. The normalized solid concentrations with respect to exit concentration, plotted as a function of column height in Figure 120, show that solids accumulation increases slightly in the presence of a distributor. This small difference could possibly be due to the unusual surging behavior in the absence of an inlet distributor (see gas holdup discussion). Nevertheless, the effect of inlet distribution is small.

The average retained solids concentration decreases with increasing column diameter as shown in Figures 121 and 122. Experiments were conducted using -140 mesh sand in a 5-in. and 12-in. diameter column in both the absence and presence of an inlet distributor at identical operating conditions. In order to compare the effect of column diameter, the solids concentration was normalized by dividing by exit concentration whereas the column height was normalized by dividing by the total depth of the column (since the 12-in. diameter column is about 5 times taller than the 5-in. diameter column). Figures 121 and 122 show that solids per unit volume accumulate more in the smaller column than in the larger one. This effect is mainly due to the increased dispersion expected in the larger column, which tends to make the solids distribution more homogeneous. Again this finding agrees with the earlier batch experimental runs which showed that increasing column diameter resulted in an increase in the solid dispersion coefficient leading to a less steep distribution profile.

5.3.5 Solid Dispersion Modelling

The ability to predict solids concentration in the slurry along the axis of the column is important. This will enable us to compute the average concentration in the column, which could be important in determining factors that affect the coal reactions occurring in the dissolver.

Figure 120. **EFFECT OF DISTRIBUTOR ON
NORMALIZED SOLIDS DISTRIBUTION**

Figure 121

EFFECT OF COLUMN DIAMETER ON
SOLIDS CONCENTRATION

Figure 122
Effect of Column Diameter

Based on a one-dimensional axial solids dispersion model that assumes steady-state conditions, our model is essentially a mass balance of the solid particles at any cross section. The equation is as follows:

$$\frac{u_l}{1 - \epsilon_g} (C - C_f) - E_{zp} \frac{dC}{dL} - V_p C = 0 \quad (6)$$

where u_l = superficial liquid velocity (ft/sec); ϵ_g = average gas holdup; C = axial solids concentration in slurry (lbm/ft³); C_f = feed slurry concentration (lbm/ft³); E_{zp} = axial solids dispersion coefficient (ft²/sec); L = axial height (ft); V_p = particle settling velocity (ft/sec).

Rearranging equation (6) by dividing by E_{zp} gives:

$$\frac{u_l}{E_{zp}(1 - \epsilon_g)} (C - C_f) - \frac{dC}{dL} - \frac{V_p}{E_{zp}} C = 0 \quad (7)$$

The values of u_l , E_{zp} , C_f and V_p/E_{zp} are required for this model.

From the earlier section (5.3.3) on solids dispersion we obtained values of V_p/E_{zp} for different gas velocities and different average solids concentrations. Usually, V_p/E_{zp} decreased as the average concentration of solids in the column increased. This was attributed to a decrease in V_p as the concentration increased.

No direct means were used to measure E_{zp} , but we assumed that:

$$E_{zp} = E_{z1} \text{ (liquid dispersion coefficient)} \quad (8)$$

The basis for this assumption is that the particles still follow the movements of the liquid fluctuations to a reasonable degree. Also, another assumption that was made was that $V_p/E_{zp} = V_p/E_{zp}(c)$, i.e. function of local concentration. Hence this ratio will vary with column height. The value of E_{z1} was already obtained in experiments described previously in the liquid dispersion section (5.2).

Due to the possibility of random errors in the measurement of the feed concentration (C_F) the C_F value used in this model was optimized to provide the best agreement between the integrated experimental data and the calculated area of the model profile. We used an iterative method that was usually fast; on the average, about 13 iterations were required to attain a relative accuracy of 10^{-4} ; i.e.;

$$\frac{W_{\text{Calcd}} - W_{\text{act.}}}{W_{\text{act}}} \leq 10^{-4} \quad (9)$$

where $W_{\text{calcd.}}$ and $W_{\text{act.}}$ are amounts of solids retained in the column by calculation and actual measurement, respectively.

Figures D1 to D40 in Appendix D plot the results of 40 runs conducted in water, tetralin, 50% glycol in the 12-in. diameter column and compare these data with the model. All of these data except for two runs with 60/80 mesh sand in 50% glycol showed good agreement between the model and experimental results. In the two runs that did not agree, mixing above the distributor plate was poor due to the low gas velocity employed ($V_g = 0.05$ ft/sec).

Hence, the one-dimensional model shows good agreement with experimental data. It also reveals that a liquid axial dispersion coefficient can be used instead of a solid dispersion coefficient, at least for particles as large as 60/80-mesh sand. In addition, the assumption that exit concentration is reasonably close to the feed concentration seems to be correct as seen by the C_F values predicted by the model.

5.4 Gas/Liquid Mass Transfer

Gas/liquid mass-transfer rates in both air/water and air/water/sand systems were investigated in the 5-in. and 12-in. diameter columns. The mass-transfer coefficient was determined from the rate at which oxygen is transferred from air to water during batch-mode operation (no liquid flow). This method has been used by other investigators and was described in detail in the Experimental Section.

The gas/liquid mass-transfer coefficient ($k_L a$) was calculated by assuming that the system was completely backmixed:

$$k_L a = \frac{1 - \epsilon_g}{t} \ln \frac{C^* - C_i}{C^* - C_f} \quad (10)$$

where $k_L a$ = volumetric liquid-phase mass-transfer coefficient, (sec^{-1}); ϵ_g = gas volume fraction (or gas holdup); t = time (sec); C^* = equilibrium oxygen concentration (mg/l); C_i = initial concentration of dissolved oxygen (mg/l); C_f = final concentration of dissolved oxygen (mg/l).

Our liquid dispersion results show that we can validly assume that the liquid phase is completely backmixed. Comparing our liquid dispersion results with a tank-in-series model indicated that the simulator columns (5-in. and 12-in. diameters) closely resemble 1.1 stirred tank in series, this supports the validity of the simple backmixed assumption used to calculate the gas/liquid mass-transfer coefficient.

5.4.1 Gas/Liquid Two-Phase System

The gas/liquid mass transfer coefficient, $k_L a$, increases with increasing gas velocity. The increase in $k_L a$ is largely due to the increase of gas/liquid interfacial contact as gas velocity increases. The mass-transfer coefficients measured in the 5-in.-diameter column were the

same as predictions from the correlation of Yoshida and Akita (12) (Figure 123). This agreement further supports the experimental technique we used to measure the mass-transfer coefficient. Yoshida and Akita's correlation includes the physical properties of the liquid phase and the diameter of the bubble column in several dimensionless groups, as shown by:

$$N_{Sh}(aD) = C_1 N_{Sc}^{0.5} N_{Bo}^{0.62} N_{Ga}^{0.31} \epsilon_g^{1.1} \quad (11)$$

where N_{Sh} = the Sherwood number = $k_L D / D_L$ (dimensionless); a = specific gas/liquid interfacial area (L^{-1}); D = column diameter (L); C_1 = dimensionless constant; N_{Sc} = Schmidt number = ν_L / D_L (dimensionless); N_{Bo} = Bond Number = $g D^2 \rho_L / \gamma$ (dimensionless); N_{Ga} = Galileo number = $g D^3 / \nu_L^2$ (dimensionless); ϵ_g = gas holdup (dimensionless); k_L = liquid-phase mass transfer coefficient (LT^{-1}); D_L = liquid-phase diffusivity ($L^2 T^{-1}$); ν_L = liquid-phase kinematic viscosity ($L^2 T^{-1}$); g = gravitational constant (LT^{-2}); ρ_L = liquid density (ML^{-3}); γ = surface tension (MT^{-2}).

Hence the agreement shown in Figure 123 also suggests the possible application of this correlation to the SRC-I coal system and scale up to industrial size.

Our results indicate that the gas/liquid mass-transfer rate is independent of column position; the rate is the same at the bottom as at the top of the column. That the rate does not depend on column position is not surprising, because the liquid phase is practically completely backmixed. Since the mass transfer rate is calculated from the oxygen uptake in the bulk liquid, one should not anticipate any effect of column position. Results from experiments with the 12-in. diameter column further substantiate this.

Although the gas/liquid mass-transfer coefficients measured in the 12-in.-diameter column increase as gas velocity increases the values deviate from Yoshida and Akita's correlation at high gas velocities, as shown in Figure 124. The measured values were always lower than

FIGURE 124
GAS/LIQUID MASS TRANSFER IN A
12-INCH DIAMETER COLUMN

those predicted. One possible explanation is the high L/D ratio of the 12-in. column. As observed in the gas holdup study, the fraction of large bubbles in the 12-in. column is substantial and increases with increasing gas velocity. An increased fraction of large bubbles means gas/liquid interfacial contact is reduced. Since the $k_L a$ value increases with increasing gas velocity because of more gas/liquid interfacial contact, the fraction of large gas bubbles observed in the 12-in. diameter column, which has a high L/D ratio, reasonably explains the deviation shown in Figure 124. This finding raises an important point. Because Yoshida and Akita's correlation was developed using a low L/D ratio column, correction is required if a high L/D ratio column is used.

In addition, the distributor plate does not affect the gas/liquid mass-transfer rate. Within experimental accuracy, the $k_L a$ values measured without a distributor plate and with two different distributor designs are the same as shown in Figure 124. The entrance effect is usually limited to a distance of a few diameters above the distributor plate. Therefore, it is not surprising that the mass-transfer rate is independent of entrance effect in the high L/D ratio column.

5.4.2 Gas/Liquid/Solid Three-Phase System

The presence of solids (-140 mesh sand) reduces the gas/liquid mass-transfer rate measured in an air/water system in the 5-in.-diameter column (Figure 123). The average bubble size in any bubble column resulted from the balance of bubble coalescence and break-up rates. The work of Kato et al. (4) showed that solids enhances the bubble coalescence rate, resulting in larger gas bubbles. Thus, gas/liquid interfacial contact will decrease, which qualitatively explains the reduction in the mass-transfer coefficient we observed.

However, the effect of solid particles is less apparent in the 12-in. column. The $k_L a$ values for the same system are indistinguishable from those for the air/water system (see Figure 124). Perhaps the substantial fraction of large gas bubbles that exist in a high L/D

ratio column (as discussed earlier) reduces the effectiveness of the solid particles in enhancing bubble coalescence rate. This qualitative explanation seems logical, but more work is needed to understand mass transfer in a three-phase system. In any event, the lower $k_L a$ values observed in the 5-in. column suggest that Yoshida and Akita's correlation must be modified to account for the effect of solids.

At this stage, our results are insufficient to make any detailed modifications. However, an empirical correction can be obtained by reducing the dimensionless constant (C_1) in equation 1 from 0.60 to 0.36. Figure 125 compares this empirically modified expression with the $k_L a$ values determined for the air/water/sand system in the 5-in. column. The agreement shown in Figure 125 suggests the possible application of this modified correlation to determine the effect of solids. Nonetheless, more work is needed either to test this correlation or to develop an expression for three-phase system.

FIGURE 125
GAS/LIQUID MASS TRANSFER IN A
5-INCH DIAMETER COLUMN

5.5 Solids Removal

Solids removal experiments were also conducted both in the 5-in. and 12-in.-diameter columns, to test the effectiveness of slurry withdrawal from the column bottom to control solids accumulation. Most of the experiments were terminated before a steady-state solid distribution profile was reached. However, some preliminary conclusions can be drawn from the available data.

The results clearly indicate that withdrawing solids at the reactor bottom significantly reduces solids accumulation and changes the concentration profile of the solids retained in the column. Results from the 5-in. column experiments are plotted in Figures 126 and 127. Figure 126 shows the data for 60/80 mesh solids, obtained at various column heights as a function of time. In the beginning of the experiment, a sand/water slurry was allowed to recycle through the column until steady state was obtained. Sampling after 90 and 120 min. of operation clearly indicated that a steady-state condition was reached. At 130 min. the bottom sample port was opened to start the solid withdrawal test; simultaneously, the recycle mode of operation was changed to a once-through mode. The average rate of slurry withdrawal was about 15% of the slurry volumetric feed rate. Slurry samples were obtained along the column at various durations after the start of the withdrawal run.

The results Figure 126 indicate that the solid concentration at every column position decreased after the start of solid withdrawal. Figure 127 presents the solid concentration profile reduced from the data shown in Figure 126. After 36 min. of solids withdrawal, the amount of retained solids decreased by 50%. Using solids concentration profiles shown in Figure 127, we calculated the amount of retained solids in the column. Figure 128, which plots the retained solids as a function of withdrawal time, clearly illustrates that the amount of solids retained decreases with increasing operating time. However, a steady-state value was not reached at the end of the experiment. In addition, the concentration profile changes drastically, as illustrated in Figure 127.

Figure 126.

EFFECT OF SOLIDS WITHDRAWAL ON SOLIDS DISTRIBUTION

Figure 127. **EFFECT OF SOLIDS WITHDRAWAL
ON AXIAL SOLIDS DISTRIBUTION**

Figure 128.

EFFECT OF SOLIDS WITHDRAWAL TIME ON RETAINED SOLIDS

The withdrawal rate has a significant impact on the amount of solids retained in the column. Results from experiments conducted in the 12-in. column using 60/80 mesh particles at different withdrawal rates are graphed in Figures 129 through 131 because precise control of the withdrawal rate was difficult and the rate varies in each experiment. At a low withdrawal rate (varied between 6 and 9% of the volumetric slurry feed rate), solids removal did not seem to affect the solids concentration at all positions along the column axis within the experimental time frame. In this experiment, the solids withdrawal rate was unsteady because the withdrawal line clogged frequently.

However, at higher withdrawal rates, solids removal definitely changed the distribution profile. Figures 130 and 131 show the results at withdrawal rates of 23-29 and 20-22%, respectively. The concentration profile in the column is significantly affected at high withdrawal rates, compared to the results obtained at low rates. Again, steady-state values were not reached in this short experiment. But the sharp difference in the non-steady-state concentration profiles at low and high removal rates definitely suggests the strong impact of the rate of solids withdrawal on solids accumulation.

An experiment with both 20/30 and 60/80 mesh sand showed that the solids withdrawal system located at the bottom of the column preferentially removed larger particle. In this experiment a known amount of 20/30 mesh sand particles was placed in the column, and a slurry containing 60/80 mesh particles was recycled through the column. Two steady-state samples were withdrawn from all the sample ports, and three samples were taken after the start of solids withdrawal. Figure 132 graphs the solids concentration at various axial positions in the column as a function of time, and Figure 133 shows the axial solids distribution before and after solids withdrawal. Both figures indicate clearly that solids distribution changes when withdrawal is employed and fewer solids are retained.

Figure 129. **EFFECT OF SOLIDS WITHDRAWAL ON SOLIDS DISTRIBUTION**

Figure 130.

EFFECT OF SOLIDS WITHDRAWAL ON SOLIDS DISTRIBUTION

Figure 131

EFFECT OF SOLIDS WITHDRAWAL ON SOLIDS DISTRIBUTION

Figure 132. **EFFECT OF SOLIDS WITHDRAWAL
ON AXIAL SOLIDS DISTRIBUTION**

Figure 133

EFFECT OF SOLIDS WITHDRAWAL ON AXIAL SOLIDS DISTRIBUTION (IN THE PRESENCE OF LARGE & FINE PARTICLES)

In addition, solids withdrawal from the bottom of the column depletes larger particles more quickly than small particles. In the experiment that used both 20/30 and 60/80 mesh particles, each sample was also screened to determine concentrations of individual particle sizes. Figure 134 illustrates the solid concentration profile of the individual particles before and 14 min. after the start of withdrawal. This figure clearly illustrates that the concentration of 20/30 mesh particles drops off more rapidly than that of 60/80 mesh particles.

When the column was drained at the end of the experiment, only 0.5 lb of the 88 lb that were retained in the column consisted of large particles. Table 33 which lists the variations in the composition of the withdrawal stream as a function of time, clearly indicates that large particles were drained first. As soon as solids removal started, the concentration of large particles in the withdrawal stream was quite large; this gradually decreased as the concentration of smaller particles increased.

Our results indicate that solids withdrawal can reduce solids accumulation and significantly change the axial solids concentration profile. Our data also indicate that larger particles are removed faster because they settle to the bottom more quickly. However, these results were obtained under non-steady-state operation and should be viewed with caution. Because of time restraints, this entire task was allotted only one month. Undoubtedly, steady-state experiments are needed in the future to determine the optimal withdrawal rate.

5.6 Particle/Particle Interaction

All of our solid distribution experiments were conducted with a narrow range of particle sizes. Individual experiments were conducted for large and small particles. In this section, we discuss results from a special set of experiments to shed some light on the effect of settled particles and particle/particle interactions on solids accumulation in the 12-in. diameter column.

Figure 134

EFFECT OF THE PRESENCE OF LARGE PARTICLES ON SOLIDS REMOVAL

Table 33

Variations in the Composition of the Withdrawal Stream

	Time after start of solids removal (min.)		
	3	9	14
Total solids concentration (lb/ft ³)	79.7	85.6	68.9
Concentration of 60/80 mesh (lb/ft ³)	33.8	50.7	57.6
Concentration of 20/30 mesh (lbs/ft ³)	45.9	34.9	11.3

The experiments were detailed in the section on experimental procedure. Briefly, they consisted of three parts: (I) only -140 mesh sand was allowed to circulate through the column; (II) a mixture of 60/80 and -140 mesh sand was used to determine the effect of large but suspended particles on the accumulation of fines; (III) 20/30 mesh particles were placed in the column to simulate settled solids, and a slurry of 60/80 and -140 mesh sand was recycled through the column. Samples from parts II and III were screened to obtain the particle-size distribution.

The results show that the presence of large suspended particles does not influence the accumulation or distribution of fine particles. The settled particles also do not affect the distribution of either large suspended or fine particles. Figure 135 summarizes the results from all three parts of the experiment. Samples from parts II and III were screened to -100 mesh, 60/100 or 40/100 and +40 mesh particles. This procedure was adapted so that the samples could be screened more efficiently. Figure 135 plots the solids concentration as a function of column height for the three groups of particles. For simplicity, the particle sizes were distinguished as follows: +40 mesh particles are large, 60/100 or -40/100 mesh particles are medium; and -100 mesh particles are as small. The plot reveals that the presence of large particles does not affect the distribution of medium and small particles; the concentration profiles for particles from parts II and III are similar. For small particles, profile for part I does not fall in the same range as those for parts II and III, which show more fines. This small difference may be due to particle attrition from the medium particles, which were used extensively in this research program. Also note that the experiments in parts II and III classified -100 mesh particles as small, whereas -140 mesh particles were classified as small in part I. Some of the 60/80 mesh particles used in parts II and III could have contained smaller particles, increasing the concentration of small particles (Figure 135). In spite of the slight difference just mentioned, our results show that large particles have a negligible influence on the accumulation of fines. This discovery also implies that superimposing the concentration profiles of individual particle sizes to simulate a solids concentration profile for a wide range of particle sizes is valid.

Figure 135

PARTICLE-PARTICLE INTERACTION

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APPENDIX A

Experimental Tracer Data in the Presence of Distributor #1

Air/Water and Air/Water/Sand Systems
(Figures A1 through A46)

Figure A1

Figure A2

Figure A3

Figure A4

Figure A5

Figure A6

Figure A7

Figure A8

Figure A9

Figure A10

Figure A11

Figure A12

Figure A13

Figure A14

Figure A15

Figure A16

Figure A17

Figure A18

Figure A19

Figure A20

Figure A21

Figure A22

Figure A23

Figure A24

Figure A25

Figure A26

Figure A27

Figure A28

Figure A29

Figure A30

Figure A31

Figure A32

Figure A33

Figure A34

Figure A35

Figure A36

Figure A37

Figure A38

Figure A39

Figure A40

Figure A41

Figure A42

Figure A43

Figure A44

Figure A45

Figure A46

APPENDIX B

Peak Time vs. Dispersion Number Data

AXIAL DISPERSION MODEL

Table B-1

Comparison of Two Different Methods in Estimated
Liquid Dispersion Coefficients

Method I - Matching Experimental and Theoretical Curves

Method II - Determining Dispersion Number from the Peak Time
vs. Dispersion Number Curve

<u>Run No.</u>	Dispersion Number	
	<u>Method I</u>	<u>Method II</u>
XXV-1	0.27	0.25
XXV-2	0.35	0.35
XXV-4	0.38	0.39
XXV-5	0.42	0.43

APPENDIX C

Experimental Tracer Data
in the Absence of Distributor

Air/Water and Air/Water/Sand Systems
(Figures C1 through C15)

Figure C-1

Figure C-2

Figure C-3

Figure C-4

Figure C-5

Figure C-6

Figure C-7

Figure C-8

Figure C-9

Figure C-10

Figure C-11

Figure C-12

Figure C-13

Figure C-14

Figure C-15

APPENDIX D

Solid Distribution Profile Data Fit

Air/Water/Sand

Nitrogen/Tetralin/Sand

Nitrogen/Glycol-Water/Sand

(Figures D1 through D40)

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND; MESH SIZE: 140 GAS VELOCITY (FT/SEC): .33 LIQUID VELOCITY (FT/SEC): .01 EL (FT ² /SEC): .62 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D2

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND; MESH SIZE: 140 GAS VELOCITY (FT/SEC): .33 LIQUID VELOCITY (FT/SEC): .03 EL (FT ² /SEC): .62 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D3

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND; MESH SIZE: 140 GAS VELOCITY (FT/SEC): .33 LIQUID VELOCITY (FT/SEC): .05 EL (FT/SEC): .62 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D4

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND, MESH SIZE: 60/80 GAS VELOCITY (FT/SEC): .33 LIQUID VELOCITY (FT/SEC): .01 EL (FT/SEC): .62 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D5

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND; MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .93 LIQUID VELOCITY (FT/SEC): .03 EL (FT=2/SEC): .52 COLUMN DIAMETER: 12 INCHES	
□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND; MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .99 LIQUID VELOCITY (FT/SEC): .05 EL (FT=2/SEC): .62 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND; MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .33 LIQUID VELOCITY (FT/SEC): .01 EL (FT/SEC): .62 COLUMN DIAMETER: 12 INCHES	
□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D8

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND; MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .93 LIQUID VELOCITY (FT/SEC): .03 EL (FT ² /SEC): .62 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: WATER SOLID: SAND; MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .33 LIQUID VELOCITY (FT/SEC): .05 EL (FT ² /SEC): .82 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND, MESH SIZE: 100 GAS VELOCITY (FT/SEC): .218 LIQUID VELOCITY (FT/SEC): .06 EL (FT/SEC): .85 COLUMN DIAMETER: 12 INCHES	
	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .308 LIQUID VELOCITY (FT/SEC): .06 EL (FT=2/SEC): .70 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D12

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 180 GAS VELOCITY (FT/SEC): .392 LIQUID VELOCITY (FT/SEC): .02 EL (FT/SEC): .85 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 140 GAS VELOCITY (FT/SEC): .392 LIQUID VELOCITY (FT/SEC): .03 EL (FT**2/SEC): .85 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D14

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .392 LIQUID VELOCITY (FT/SEC): .05 EL (FT ² /SEC): .95 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 140 GAS VELOCITY (FT/SEC): .392 LIQUID VELOCITY (FT/SEC): .05 EL (FT ² /SEC): .85 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D16

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 8080 GAS VELOCITY (FT/SEC): .133 LIQUID VELOCITY (FT/SEC): .05 EL (FT²/SEC): .85 COLUMN DIAMETER: 12 INCHES	
□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D17

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .216 LIQUID VELOCITY (FT/SEC): .05 EL (FT²/SEC): .65 COLUMN DIAMETER: 12 INCHES	
	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D18

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .308 LIQUID VELOCITY (FT/SEC): .05 EL (FT/SEC): .70 COLUMN DIAMETER: 12 INCHES	
	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D19

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .392 LIQUID VELOCITY (FT/SEC): .02 EL (FT ² /SEC): .85 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D20

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND, MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .382 LIQUID VELOCITY (FT/SEC): .03 EL (FT/SEC ²): .86 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .992 LIQUID VELOCITY (FT/SEC): .04 EL (FT/SEC): .85 COLUMN DIAMETER: 12 INCHES	
	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: TETRALIN SOLID: SAND; MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .392 LIQUID VELOCITY (FT/SEC): .05 EL (FT=2/SEC): .85	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .05 LIQUID VELOCITY (FT/SEC): .01 EL (FT**2/SEC): .430 COLUMN DIAMETER: 12 INCHES	
□	EXPERIMENTAL POINTS
—	MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND; MESH SIZE: 140 GAS VELOCITY (FT/SEC): .05 LIQUID VELOCITY (FT/SEC): .03 EL (FT=2/SEC): .438 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .05 LIQUID VELOCITY (FT/SEC): .05 EL. OPT. (2/SEC): .438 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D26

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .200 LIQUID VELOCITY (FT/SEC): .01 EL (FT ² /SEC): .688 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND; MESH SIZE: 140 GAS VELOCITY (FT/SEC): .197 LIQUID VELOCITY (FT/SEC): .03 EL (FT/IN ² /SEC): .588 PIPE DIAMETER: 1.5 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D28

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .200 LIQUID VELOCITY (FT/SEC): .05 FL (FT²/SEC): .688 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D29

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .400 LIQUID VELOCITY (FT/SEC): .01 EL (FT/SEC): .803 PIPE DIAMETER: 12 INCHES	
□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .400 LIQUID VELOCITY (FT/SEC): .03 EL (FT/SEC): .803 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 140 GAS VELOCITY (FT/SEC): .400 LIQUID VELOCITY (FT/SEC): .05 EL (FT/SEC): .003 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE, 6080 GAS VELOCITY (FT/SEC): .043 LIQUID VELOCITY (FT/SEC): .01 EL (FT/SEC): .438 -----	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .05 LIQUID VELOCITY (FT/SEC): .03 EL (FT=2/SEC): .430	
□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D34

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .05 LIQUID VELOCITY (FT/SEC): .05 EL (FT*2/SEC): .439 COLUMN DIAMETER: 12 INCHES	
<div> <div></div> EXPERIMENTAL POINTS </div>	<div> <div></div> MODEL CURVE </div>

FIGURE D35

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND; MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .2 LIQUID VELOCITY (FT/SEC): .01 EL (FT/SEC): .500 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .2 LIQUID VELOCITY (FT/SEC): .03 EL (FT ² /SEC): .588 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D37

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND; MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .197 LIQUID VELOCITY (FT/SEC): .05 EL (FT/SEC²): .588 COLUMN DIAMETER: 12 INCHES	
<div style="display: inline-block; width: 10px; height: 10px; border: 1px solid black; margin-right: 5px;"></div> EXPERIMENTAL POINTS	<div style="display: inline-block; width: 10px; border-bottom: 1px solid black; margin-right: 5px;"></div> MODEL CURVE

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND; MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .400 LIQUID VELOCITY (FT/SEC): .01 EL (FT/SEC): .903 COLUMN DIAMETER: 12 INCHES	
□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D39

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: SO2 GLYCOL SOLID: SAND; MESH SIZE: 6080 GAS VELOCITY (FT/SEC): .400 LIQUID VELOCITY (FT/SEC): .03 EL (FT/SEC): .903 COLUMN DIAMETER: 12 INCHES	□ EXPERIMENTAL POINTS — MODEL CURVE

FIGURE D40

CONC. VS. L	
SLURRY CONCENTRATION VS. AXIAL LENGTH LIQUID: 50% GLYCOL SOLID: SAND, MESH SIZE: 6000 GAS VELOCITY (FT/SEC): .400 LIQUID VELOCITY (FT/SEC): .05 Q (FT³/SEC): .803 COLUMN DIAMETER: 12 INCHES	
□ EXPERIMENTAL POINTS — MODEL CURVE	