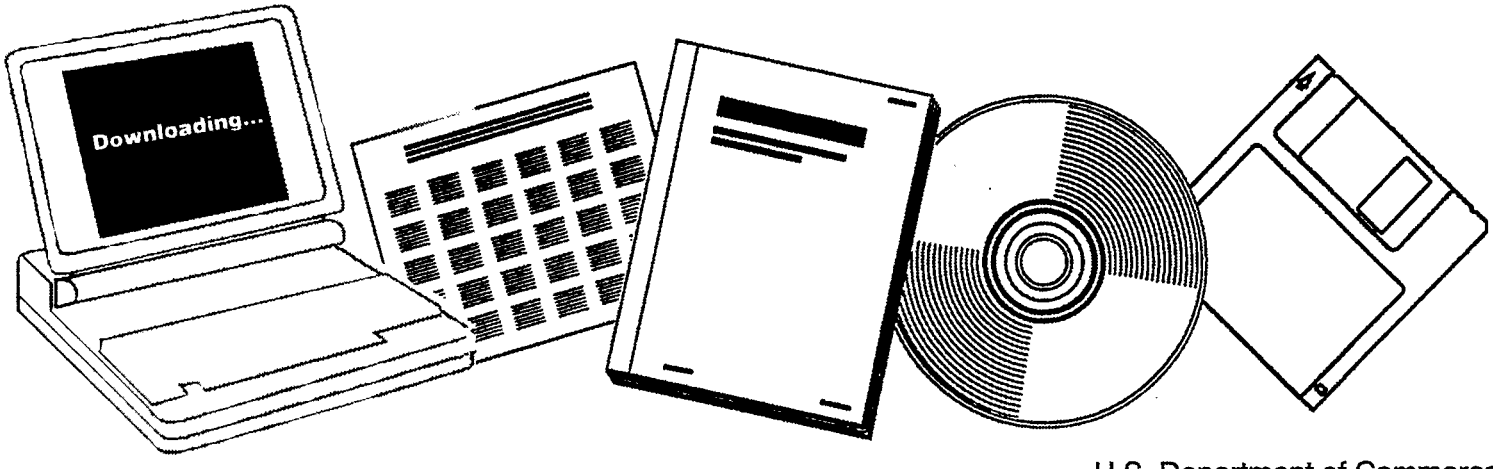
GAS/SLURRY FLOW IN COAL-LIQUEFACTION PROCESSES (FLUID DYNAMICS IN 3-PHASE FLOW COLUMNS). QUARTERLY TECHNICAL PROGRESS REPORT, 1 JULY 1980-30 SEPTEMBER 1980

AIR PRODUCTS AND CHEMICALS, INC.
ALLENTOWN, PA

DEC 1980



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GAS/SLURRY FLOW IN COAL LIQUEFACTION PROCESSES
(Fluid Dynamics in 3-Phase Flow Columns)

Quarterly Technical Progress Report
for Period 1 July 1980-30 September 1980

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SUMMARY

This is the fourth quarterly report under Contract Number DE-AC01-79ET14801 titled, "Gas/Slurry Flow in Coal Liquefaction Processes". This work covers the period 1 July 1980 to 30 September 1980. 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 the 6000 T/D dissolver for the SRC-I demonstration plant which began in July 1978. DOE supported these 3-phase flow studies under the Bridging Task program from 1 July 1979 to 30 September 1979 at the start of the current contract. A background of information developed at Air Products prior to DOE support was included in the first quarterly report.

The 6000 T/D SRC-I demonstration plant will employ vertical tubular reactors feeding slurry and gas concurrently upward through these vessels. In the SRC-I design this 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 either Wilsonville or Ft. Lewis, this program is addressing the need for additional data on behavior of three phase systems in large vessels. Parameters being investigated in this program are being studied at conditions that relate directly to the projected demonstration plant operating conditions. Air/water/sand 3-phase flow system in both a 5-inch diameter and a 12-inch diameter column is used in this cold-flow simulation study program.

The amount of solids retained in the column decreases with increasing column diameter. At the flow conditions designed for the SRC-I demonstration plant, the results show less fine particles (140 mesh minus) accumulated in the 12-inch diameter column than the 5-inch diameter column. This finding agrees qualitatively with the increase in dispersion with increasing column diameter, thereby leading to the conclusion that there will be practically no accumulation of fines in the demonstration plant dissolver. If the retained solids are catalytically active in the coal liquefaction process, the performance of the demo plant dissolver will be lower than projected because of less solids accumulation. In addition, for a wide range of gas (0.05 to 0.33 ft/sec) and slurry velocities (0.01 to 0.05 ft/sec) covering the flow conditions designed for the SRC-I demonstration plant, the amount of retained solids in the 12-inch

diameter column is insensitive to the total opening area in the distributor plate and to gas velocity beyond the critical value. However, the amount of accumulated solids continuously increased with decreasing slurry velocity, thus favoring the dissolvers in parallel operation from the solids accumulation stand point.

The total area of opening in the distributor plate has no effect on gas holdup. For the wide range of gas (0.05 to 0.33 ft/sec) and slurry velocities (0.01 to 0.05 ft/sec) studied in this program, Yoshida and Akita's correlation remains reasonably valid to describe the gas holdup in the air/water/sand system irrespective of the presence of distributor plate with different opening areas or the absence of a distributor. During this quarter, a photographic technique is developed to determine the effect of gas and liquid velocities on bubble size distribution. The results indicate that the fraction of gas holdup occupied by large bubbles increases with increasing gas velocity but is independent of liquid flow rate.

The gas/liquid mass transfer rate, as measured from the oxygen uptake rate in the liquid phase, is independent of column height and the total area of opening in the distributor plate. However, its dependence on gas velocity behaves quite differently as the column diameter varies. In both the 5-inch diameter and 12-inch diameter columns the gas/liquid mass transfer coefficient in 2-phase system increases with 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 constant. In the presence of solids, however, the influence of column diameter becomes less apparent.

The liquid dispersion results have clearly shown that the demonstration plant dissolver is a highly backmixing reactor. The liquid phase dispersion coefficient increases with gas velocity but is independent of liquid flow rate regardless of the type of distributor, the absence of distributor, and the column diameter. Although a small reduction in the dispersion coefficient in the presence of solid has been consistently observed, the dissolver behaves essentially like a well-mixed vessel.

In the next quarter, a solid withdrawal system to remove large settled solid particles from the column bottom will be tested.

1.0 OBJECTIVE

The overall objective of this project is to study the solids accumulation and suspension of various gas/liquid/solid systems in cold-flow tubular columns aimed at providing data for the coal dissolver design in the SRC-I demonstration plant.

The specific objectives are:

1. To check the adequacy of the existing experimental apparatus using a two-phase system (air/water mixture).
2. To study the effects of slurry and gas velocities, solid particle size and concentrations, and liquid viscosity and surface tension on the performance of a cold-flow tubular column.
3. To develop an effective slurry withdrawal technique from the bottom of a tubular column as a means to control the solid concentration in the column.
4. To study the performance of cold-flow tubular column with an improved distributor and in the absence of a distributor.
5. To explore the use of multiple distributors in a tubular column.

2.0 INTRODUCTION

A major element of the coal dissolution section of any liquefaction plant is the dissolver. Although a considerable amount of liquefaction will occur in the preheater, a major amount of necessary chemical change will occur in the dissolver, namely sulfur removal, oil and distillate formation and solvent rehydrogenation.

Vertical tubular reactors are employed in all of the major processes currently under consideration for commercial liquefaction of coal. In all of these processes, SRC, EDS and H-Coal, slurry and gas are concurrently fed upward through these vessels. In the EDS and SRC processes, the reactors are essentially empty vessels, whereas for the H-Coal process a bed of ebullating catalyst is maintained in the reactor. The major differences between the EDS, SRC-I and SRC-II processes in dissolver

operation are the composition of the feed streams and reactants within the dissolver. Other hardware differences such as distributor plates, draft tubes or recycle loops can also cause differences in the behavior of slurries in these vessels. A requirement necessary to any design that will be technically feasible and cost effective is an understanding of the physical behavior of three phase systems in tubular columns.

All of the major processes under development require understanding of backmixed three phase systems. Each process employs at least a portion of its dissolver volume in a backmixed mode. As the design of the 6000 T/D SRC-I 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 represent a considerable scale-up over the Wilsonville and Ft. Lewis dissolvers. To intelligently make good design decisions, more information is needed on the flow properties of three phase systems in large vessels. More important from the standpoint of slurry behavior is the difference in gas and liquid superficial velocities. This difference can have 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 in size. Since considerable evidence points to a definite catalytic effect of the reactor solids, these larger particles will have decreased surface areas exposed and will likely have diminished catalytic activity. Knowing the particle sizes that can accumulate under commercial flow conditions will give us some indication of size of dissolver solids that should be examined for catalytic activity.

Considerable work on the behavior of gas/liquid mixtures flowing through vertical columns has been reported in the literature. Information on three phase (gas/liquid/solid) systems is far less extensive. Detailed

background information was presented in the first quarterly report (1). Under this contract, work is being conducted in a 5-inch diameter by 5-foot tall Plexiglass column and a 12-inch diameter by 25-foot tall glass column which are located at the contractor's site. The physical dimensions, auxiliary equipment, and some of the experimental techniques employed in this study were extensively discussed in the first quarterly report. This report contains experimental results from runs conducted during this reporting period (1 July-30 September 1980).

3.0 TECHNICAL PROGRESS

3.1 Task 3 - Effect of Distributor on Flow

The objectives of this task are:

- To study the entrance effects on the performance of the 12 inch column.
- To study the performance of the column in the absence of a distributor.
- To design, fabricate and install a new distributor and to compare the performance of the different distributors.

During the last quarter, gas holdup and liquid dispersion experiments were completed in the 12-inch diameter column without a distributor. A new distributor plate was designed and fabricated.

During this quarter, gas holdup, liquid dispersion and solids distribution experiments were conducted in the 12-inch diameter column. In addition, mass transfer experiments were conducted with the new distributor (No. 2) and the distributor with bubble caps (No. 1). A description of the two different distributors is given in the experimental section. A photographic technique was used to examine the bubble sizes resulting from both distributors at various gas and liquid velocities. Detailed description of the photographic method used in these experiments is given in the section on experimental procedure.

4.0 EXPERIMENTAL SECTION

4.1 Cold Flow Model Equipment

Both the 5-inch diameter and 12-inch diameter columns used in these cold-flow studies were described in detail in the first quarterly report (1). The distributor used earlier under Task 2 is nearly identical to the distributor in the Wilsonville dissolver. For Task 3, a new distributor was designed and fabricated. This new distributor will be referred to as distributor No. 2 whereas the one for Task 2 is designated as No. 1. A description of the two different distributors is given below.

- Distributor No. 1 (Task 2)

  This distributor's design closely resembles the one in Wilsonville's dissolver. It consists of seven 9/16 inch openings with a 3-3/4 inch center-to-center spacing. A 3-1/2 inch long with 9/16 inch I.D. riser tube was welded to each opening as illustrated in Figure 1. There are two 1/4 inch by 1 inch rectangular slots facing 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 through each hole in the plate protrudes a 1-1/4 inch I.D. by 3-1/2 inch long tube. Each tube was covered by a cap leaving a 5/32 inch spacing between the tube and the cap. The gas/slurry flows up through the riser tube and enters into the upper tube. This gas/slurry continues to climb up through the upper tube and runs down along the spacing between the upper tube and the cap. The material finally flows out of the cap through a 1/4 inch gap between the cap and the distributor plate. The complete assembly is shown in Figure 2.

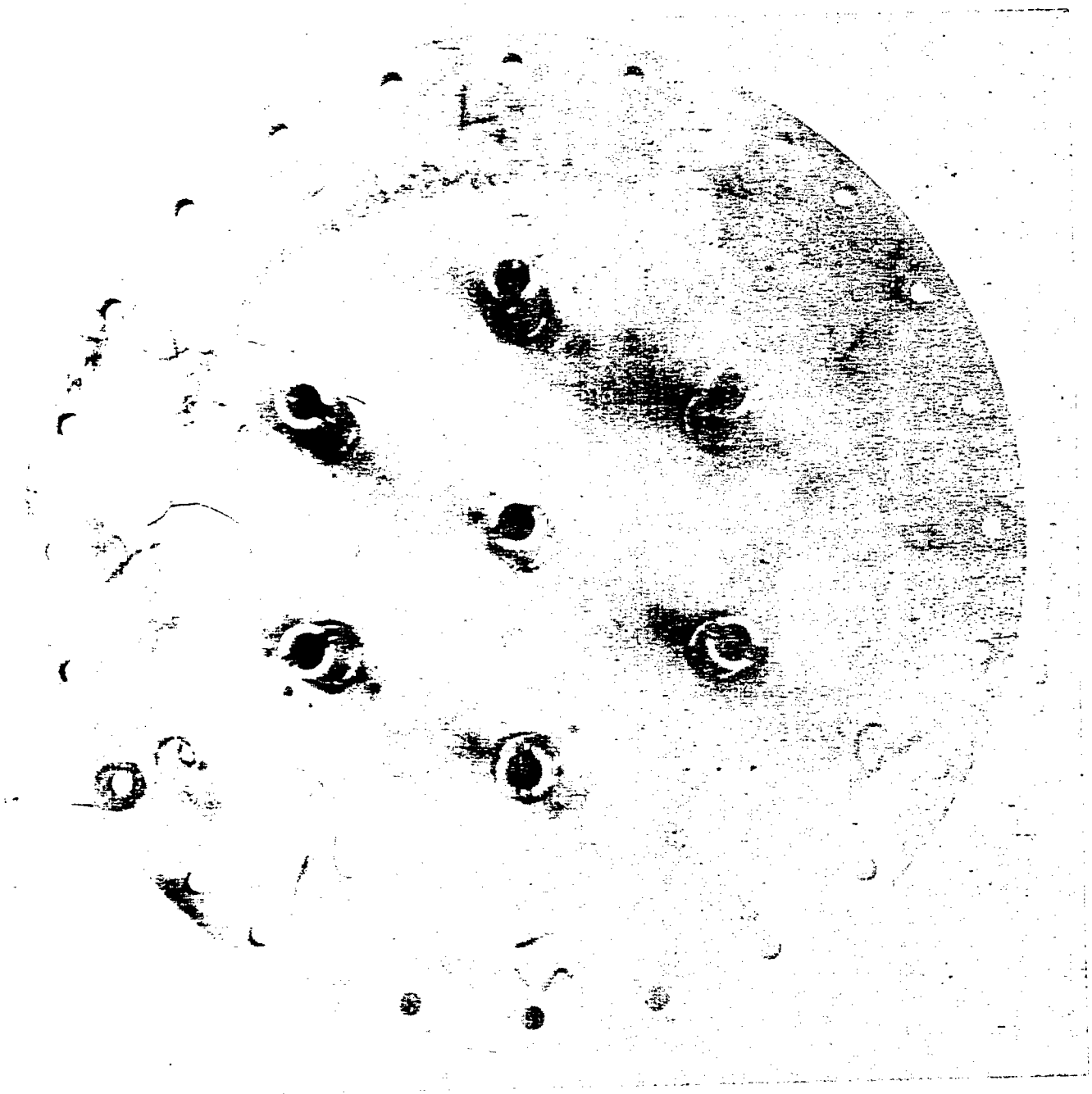


FIGURE 1

Distributor Number 1

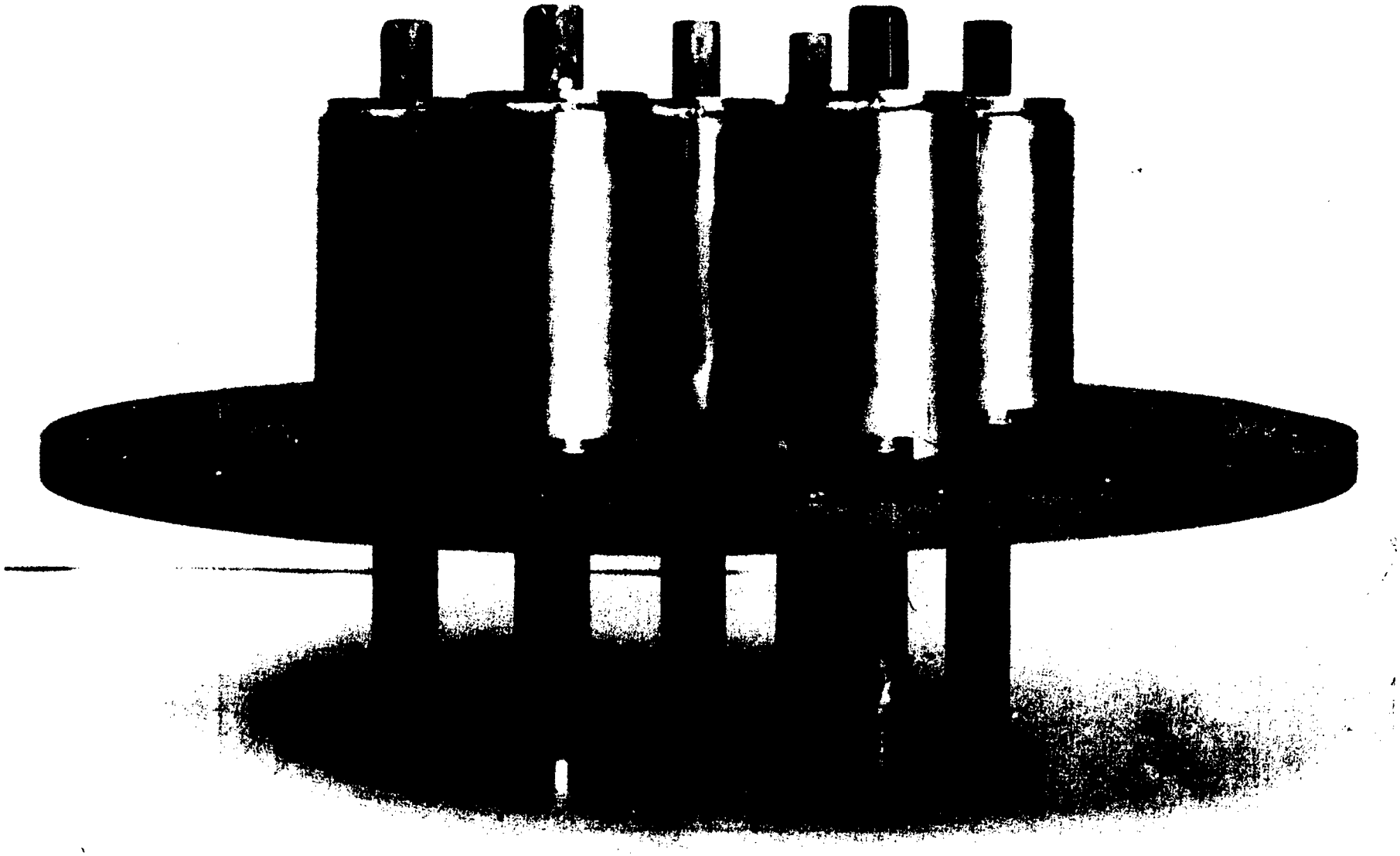


FIGURE 2

Distributor Number 1

- Distributor No. 2 (Task 3)

This plate has nineteen 9/16 inch openings with a 2 inch center-to-center spacing as illustrated in Figure 3. Each hole has a riser tube identical to that in distributor No. 1. No bubble cap nor upper tube was attached to the other side of the plate. The distributor is shown in Figure 4.

4.2 Experimental Procedures

4.2.1 Gas Holup

Gas holdup was measured in the 12-inch diameter column with distributor No. 2 in both the absence and the presence of fluid flow. Effects of solid particles on gas holdup were also investigated in both modes of fluid motion.

In the absence of liquid flow, the experiments were performed by completely filling the column with liquid and then passing air through the liquid at specified rates. Excess liquid exited the column at the top through a side opening. A waiting period of 5 minutes was allowed to ensure that steady state was achieved. The bottom valve was then closed to shut off the gas input. The final liquid level was measured, and the difference between the initial and the final levels represented the gas holdup at that particular gas flow rate. Gas flow rates ranging from 0.05 ft/sec to 0.43 ft/sec were studied.

With fluid flow, the liquid and gas passed into the column through a centrally located opening at the bottom. Excess liquid exited the column through a side opening at the top. After 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 above. Liquid velocities ranging from 0.01 ft/sec to 0.05 ft/sec were studied. Both 20/30 mesh and 140 mesh minus sand particles were used to investigate the effect of solids on gas holdup.

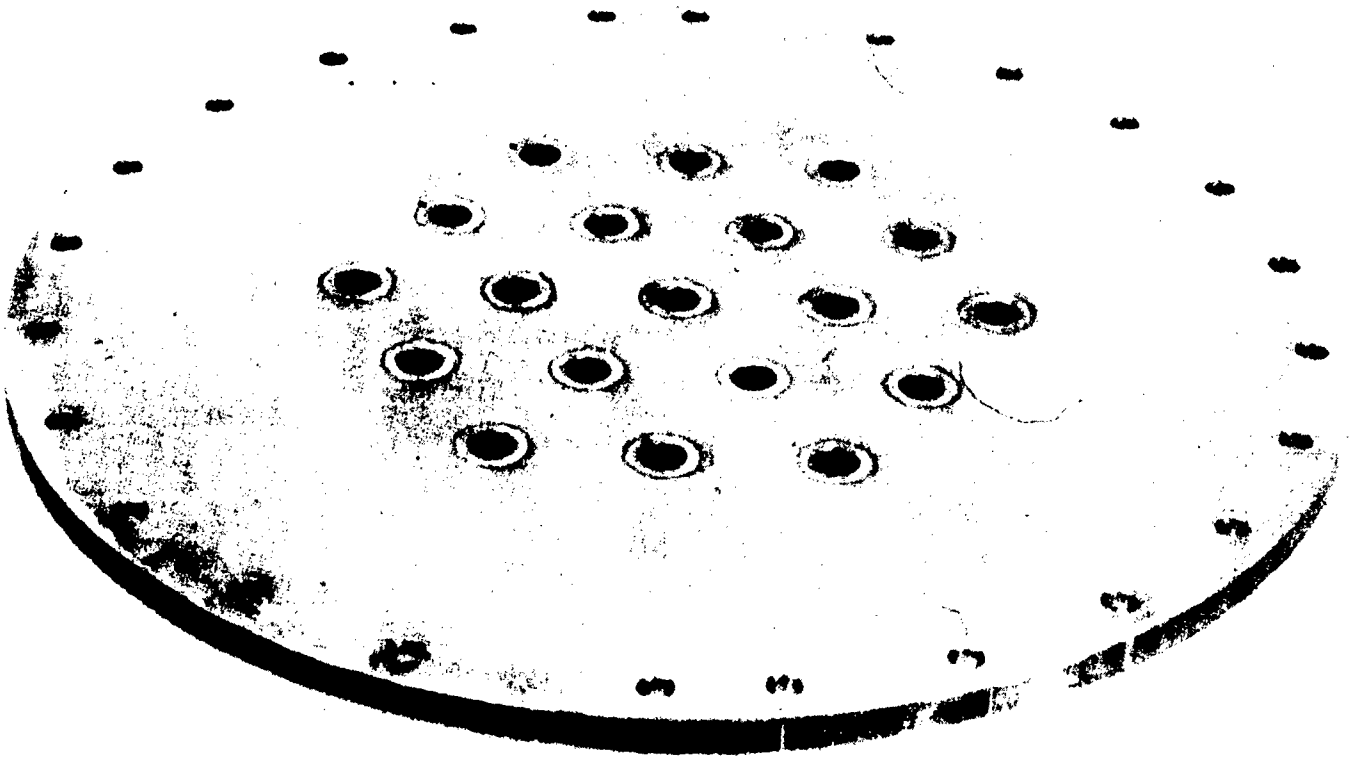


FIGURE 3

Distributor Number 2

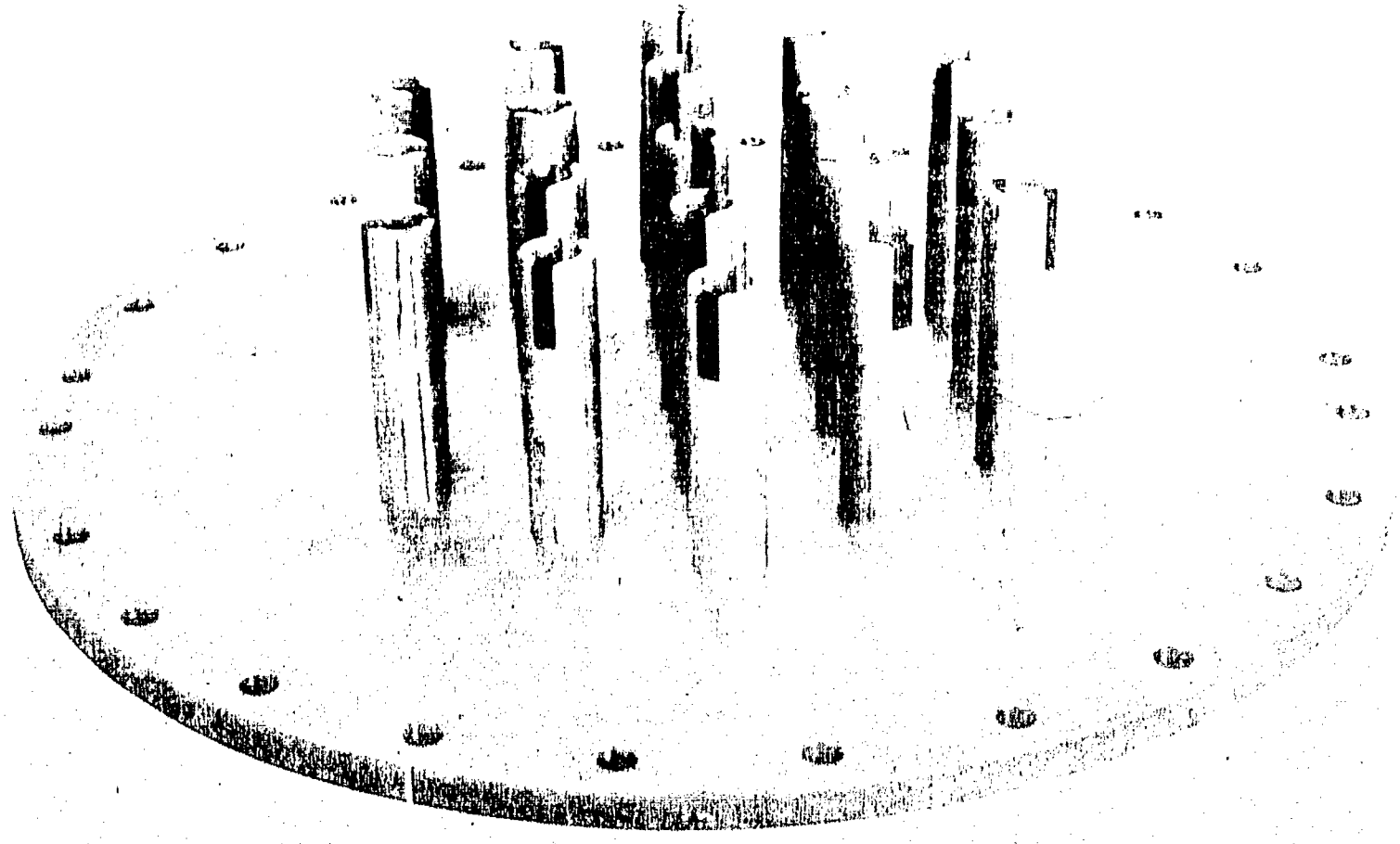


FIGURE 4

Distributor Number 2

Two different solid concentrations in the column were used. For experiments using 20/30 mesh sand, a known amount of sand was initially placed in the column and the experiments proceeded 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 140 mesh minus 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.

4.2.2 Photographic Method

In order to get a qualitative understanding of the bubble sizes that exist in the column, a photographic method was used. A discussion of the reasoning behind the method is presented in the experimental discussion section. Only a description of the experimental procedure is presented here.

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 waiting for a period of ten minutes to establish steady state, 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 leaves the system. This change of aerated liquid level as a function of time was recorded photographically at every one-second interval. This procedure was repeated for various liquid and gas flow rates. A total of 5 experiments were conducted in this quarter with distributor No. 2 in place in the column.

4.2.3 Liquid Dispersion

Liquid phase dispersion in both air/water and air/water/sand systems was investigated in a 12-inch diameter column using a tracer detection method. Sodium chloride was used as the tracer and a conductivity probe mounted at the exit line of the column monitored the ion

conductivity of the solution continuously. Detailed experimental procedures can be found in the first quarterly report (1). During this reporting period, liquid dispersion experiments were conducted with distributor No. 2 in place in the column. Liquid velocities ranging from 0.02 to 0.05 ft/sec and gas velocities ranging from 0.05 to 0.43 ft/sec were employed in these experiments. Two different particle sizes of sand (20/30 mesh and 140 mesh minus) were used in this quarter.

4.2.4 Gas/Liquid Mass Transfer

The effect of the two distributors (No. 1 and 2) on gas/liquid mass transfer was studied both in the presence and absence of solids. Only batch experiments were conducted in this quarter.

The gas/liquid mass transfer rate was measured by the rate of oxygen dissolving in water. A submersible polarographic probe was used to continuously monitor the concentration of dissolved oxygen in the solution with time. Detailed experimental procedures can be found in the third quarterly report (3).

4.2.5 Solids Distribution

The effect of liquid and gas velocities on the axial distribution of 60/80 mesh and 140 mesh minus sand particles were studied in the 12 inch column using the distributor No. 2 in place. All experiments were conducted in a recycle mode which was described in the second quarterly report (2). Briefly, the exit from the column was returned to the feed tank thereby creating a closed loop. The recycle loop was allowed to operate for several hours in order to achieve steady state. Solid concentration profiles were measured from slurry samples withdrawn from sampling ports located along the axis of the column.

5.0 RESULTS AND DISCUSSION

5.1 Gas Holdup

Gas holdup measurements in the 12-inch diameter column with distributor No. 2 in place were made both in the presence and absence of liquid flow and in the presence and absence of solids. Gas velocities varied from 0.05 to 0.43 ft/sec and liquid velocities ranged from 0.01 to 0.05 ft/sec. Two different particle sizes of solids, 20/30 mesh and 140 mesh minus were used. In the case of the larger particle size (20/30 mesh) the sand was placed in the column. Then water and air were allowed to flow through the column. However, in the case of the fine particles (-140 mesh), a sand/water slurry was pumped into the column.

The overall gas holdup is not influenced by entrance effect. A comparison of gas holdup results from experiments with distributor No. 1 and without a distributor indicates that gas holdup was essentially independent of the presence or absence of a distributor and the type of distributor used as shown in Figures 5 through 10. These figures compare gas holdup data obtained from three different distributor configurations (absence of a distributor, distributor with seven bubble caps designated as distributor No. 1 and distributor with nineteen openings but no bubble caps designated as distributor No. 2) for various experimental conditions (presence and absence of liquid flow and solid particles). In these figures (5 through 10) the solid line represents the gas holdup values predicted by the Yoshida and Akita correlation (4) which provides a reasonable fit except at very high gas velocities, particularly when no distributor was present.

At these higher gas velocities extreme surging took place when no distributor was present as explained in the previous quarterly report; hence those values are questionable. Figures 6 through 10 indicate that liquid flow and presence of solids do not significantly affect gas holdup values; the correlation of Yoshida and Akita still predicts gas holdup values reasonably well except at very high gas velocities.

FIGURE 5 – EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP IN THE ABSENCE OF LIQUID FLOW AND SOLID PHASE

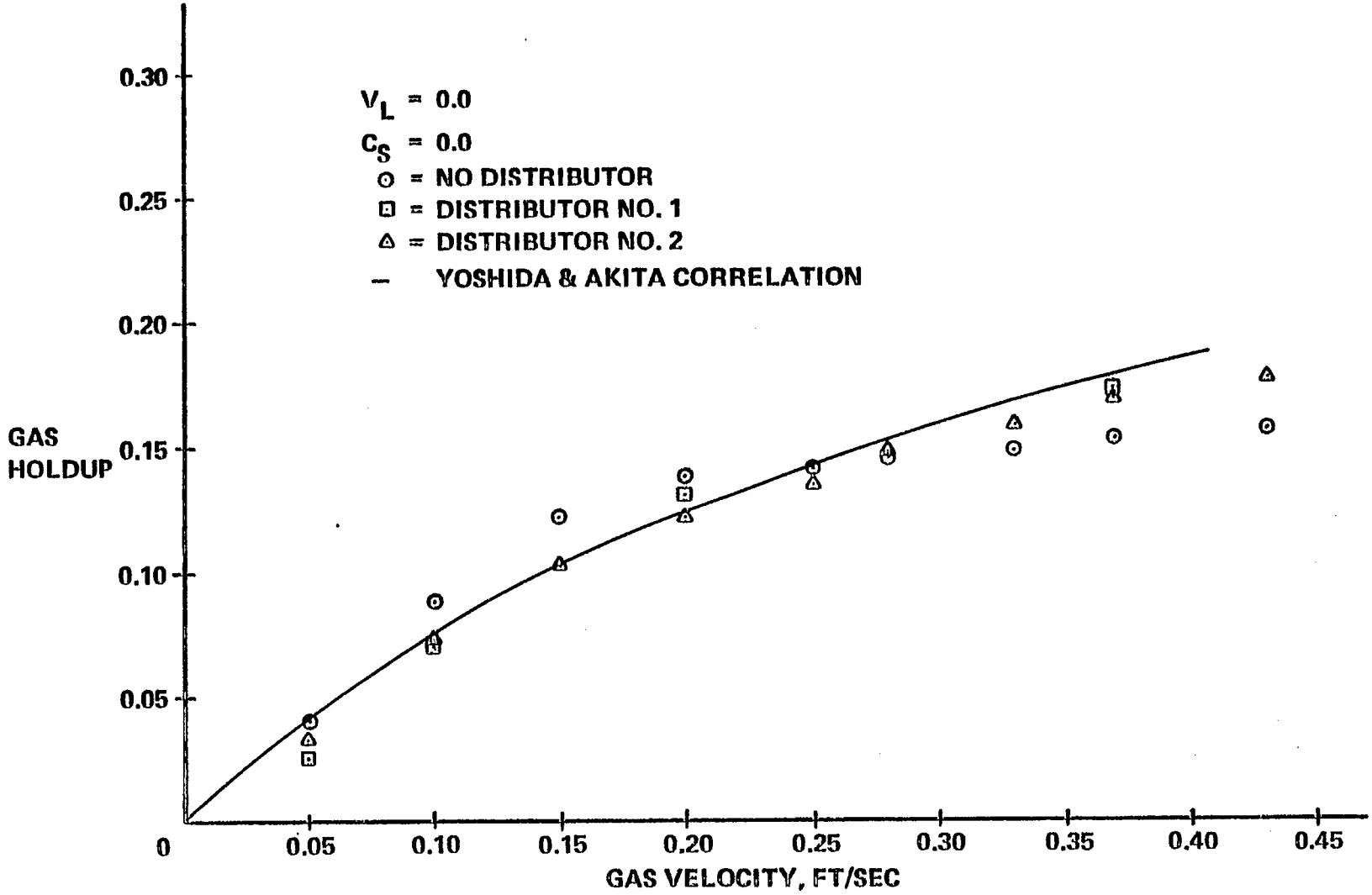


FIGURE 6 – EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT LOW LIQUID VELOCITY WITH NO SOLIDS

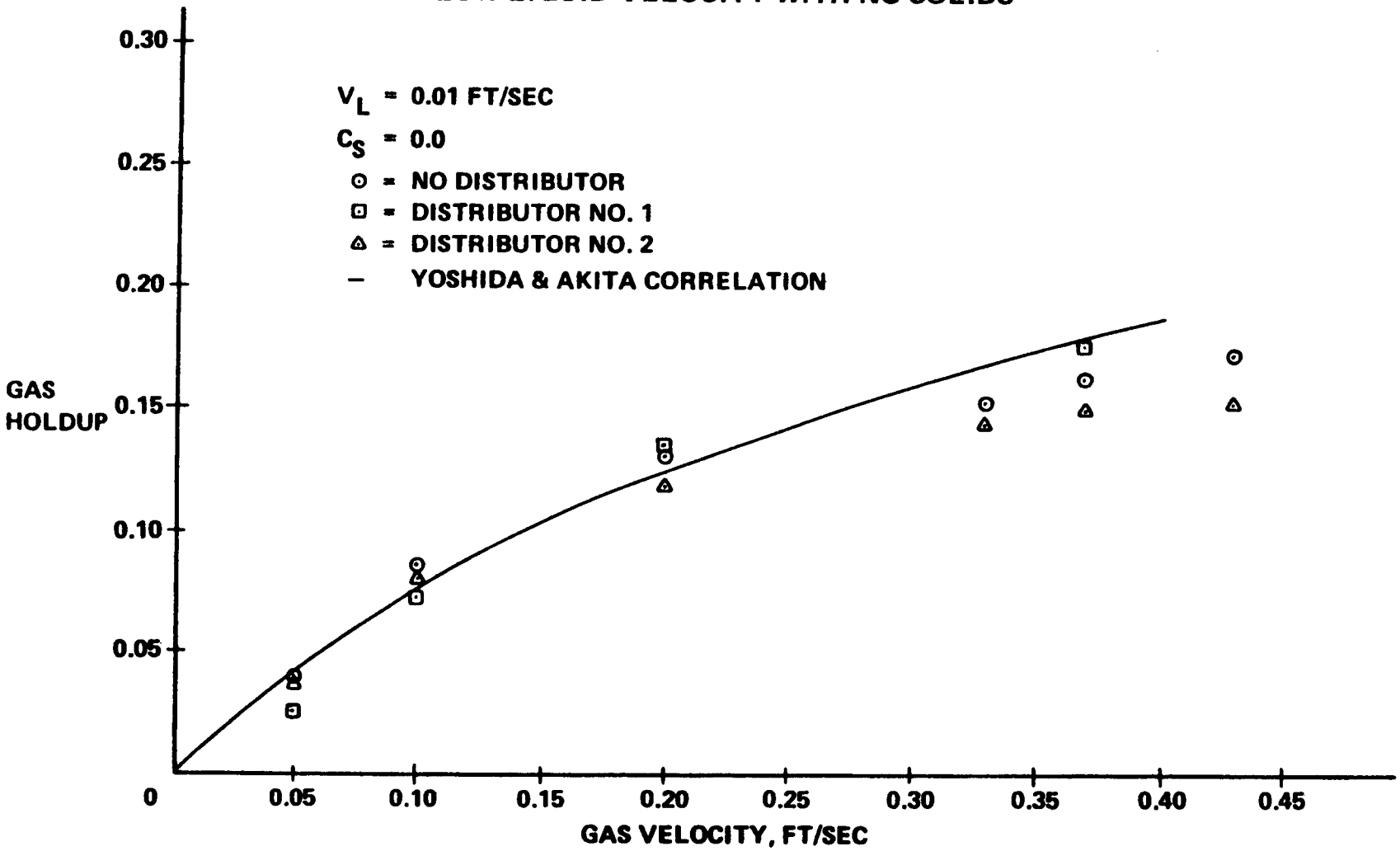


FIGURE 7 – EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT HIGH LIQUID VELOCITY WITH LOW CONCENTRATION OF LARGE SOLID PARTICLES

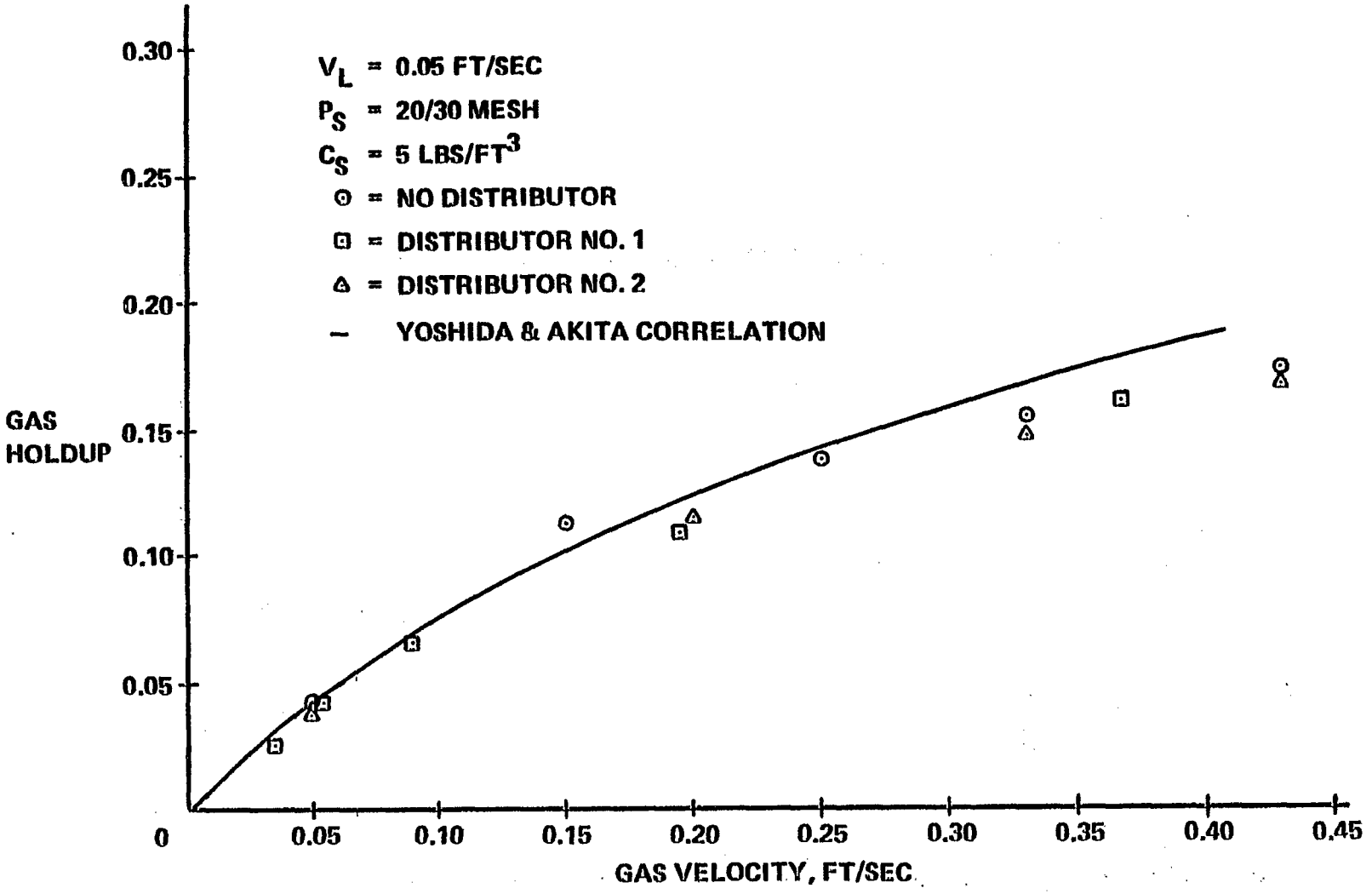


FIGURE 8 – EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT LOW LIQUID VELOCITY WITH HIGH CONCENTRATION OF LARGE SOLID PARTICLES

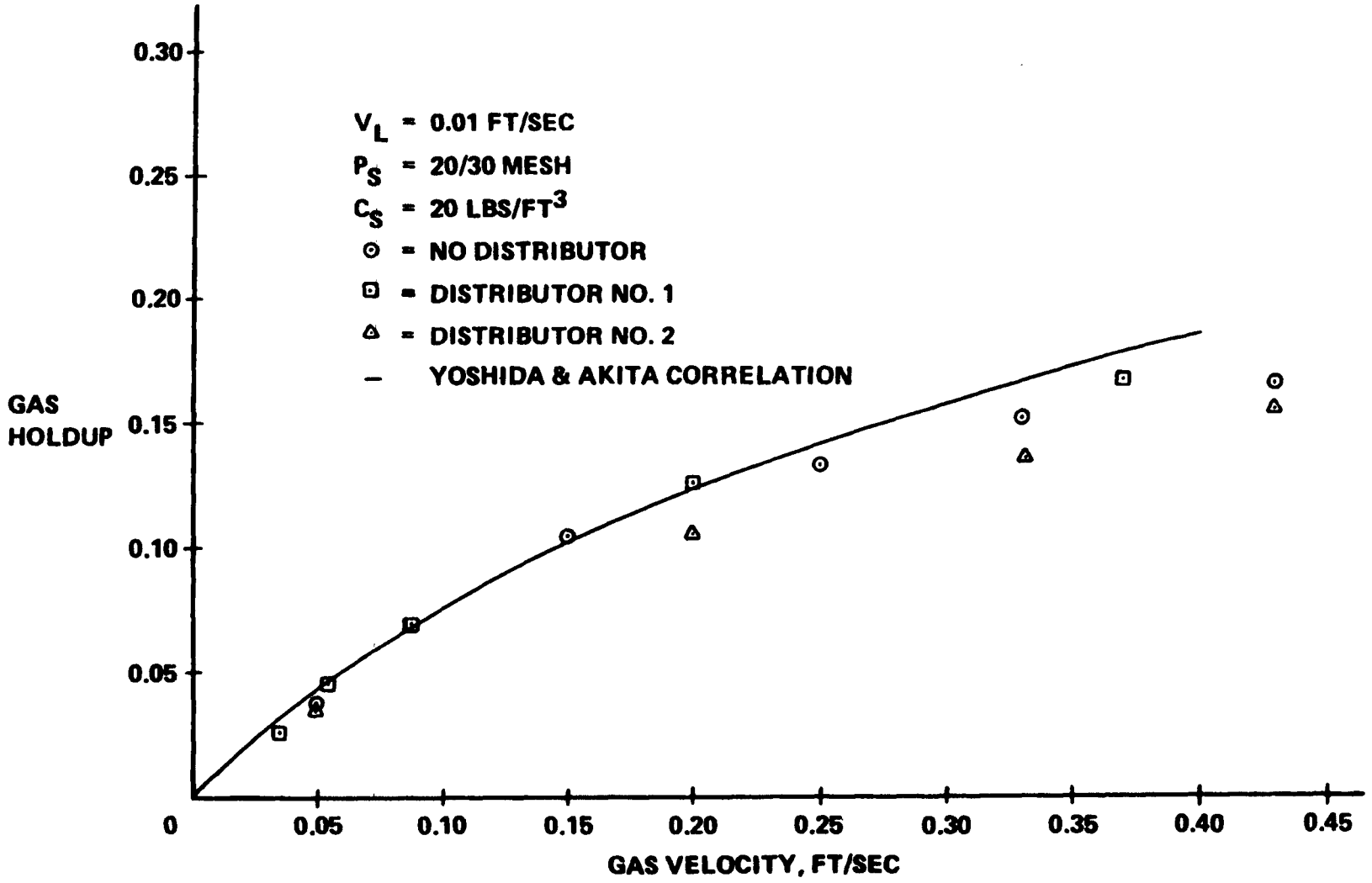


FIGURE 9 – EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT HIGH LIQUID VELOCITY WITH LOW CONCENTRATION OF FINE PARTICLES

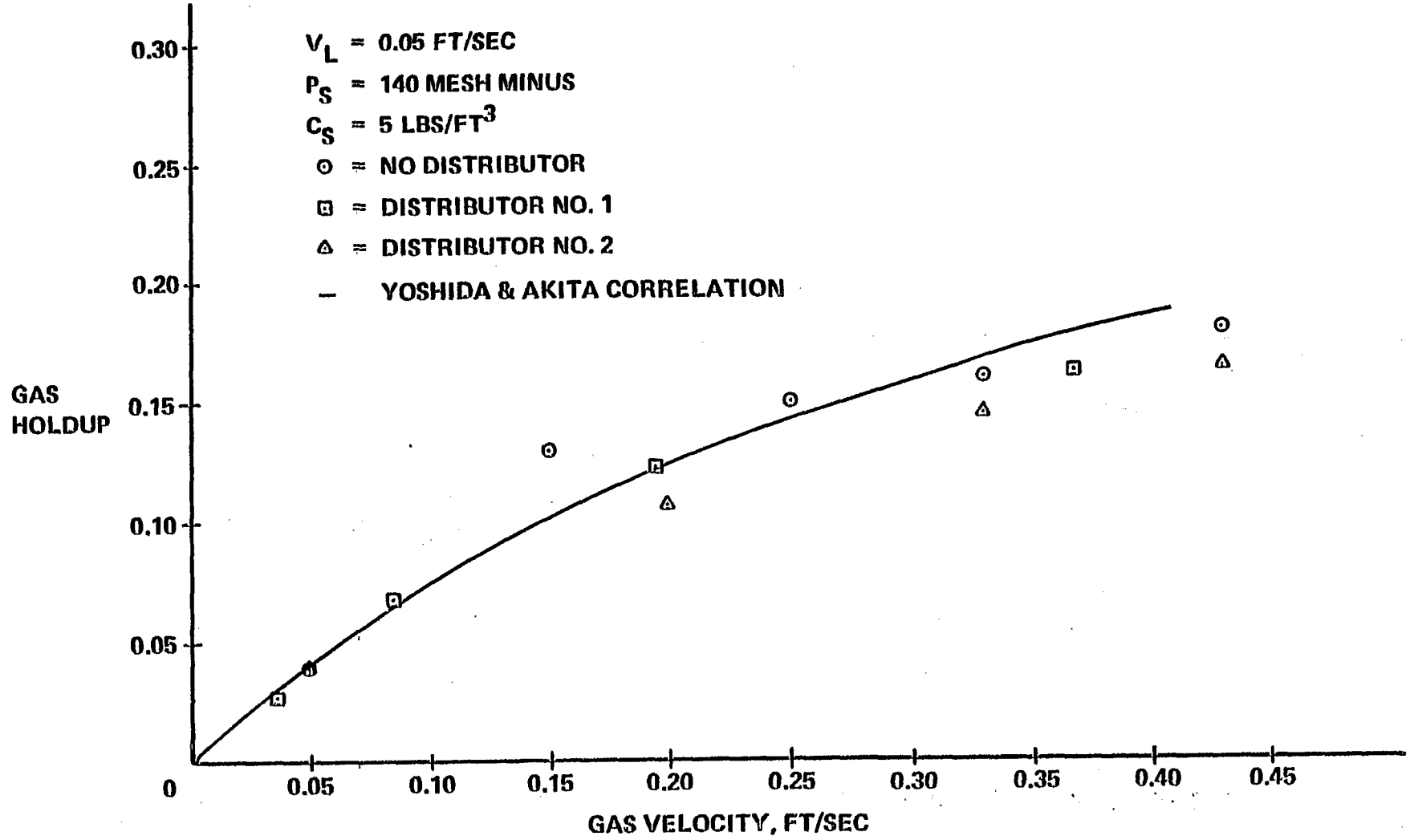
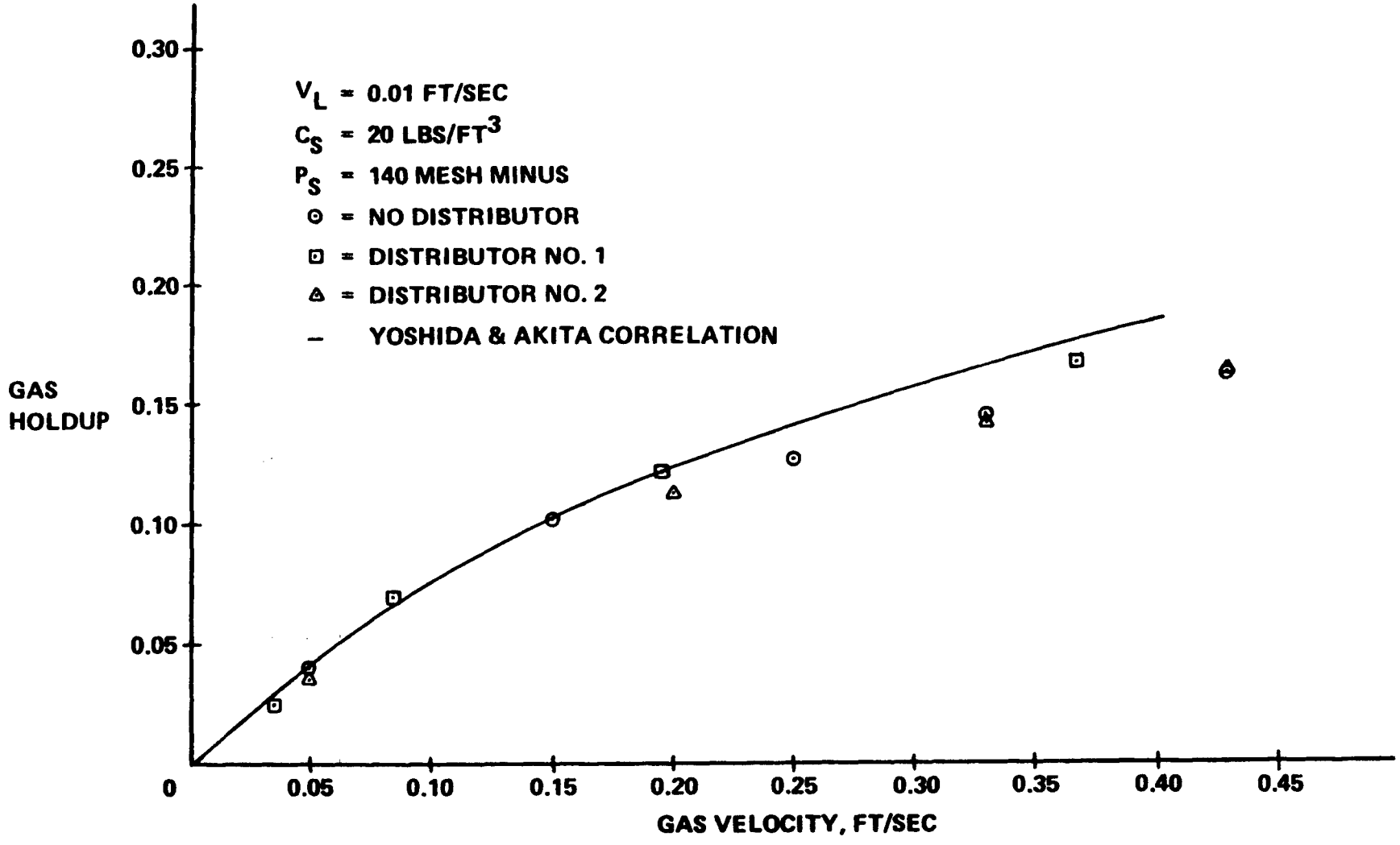


FIGURE 10 – EFFECT OF DISTRIBUTOR PLATE ON GAS HOLDUP AT LOW LIQUID VELOCITY WITH HIGH CONCENTRATION OF FINE PARTICLES



5.2 Photographic Method

A photographic method was developed to identify the fraction of gas void volume occupied by large gas bubbles. Results presented earlier indicate that gas holdup is independent of the type of distributor used. Since gas holdup is inversely proportional to the size of the bubbles, these results indicate that no significant changes occur in bubble sizes between different distributors. However, visual observations had shown that large slugs were formed in the absence of a distributor that resulted in some unusual flow patterns. Still the presence of these large slugs did not result in a decrease in gas holdup. This could lead one to speculate that these large bubbles (slugs) probably occupied a very small fraction of the gas volume. In order to examine this speculation, a photographic method was developed.

A detailed description of the photographic method was given in the section on experimental procedure. The gas void volume is measured by the stop flow method described above. Briefly, the drop in the liquid level at the top of the column is measured every second after the liquid and gas flows had been shut off without the distributor.

Data were taken at various liquid and gas velocities. In interpreting these data we assumed that the large bubbles will rise faster than the smaller bubbles which is logical from a rational standpoint. Realizing the limitations, this method cannot give any quantitative measurements of bubble size. However, one can hope to 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 prior group, rising at a faster rate. Visual observations indicated that the drop in liquid level (after the gas and liquid flows had been shut off) is very unsteady in the beginning and after a certain time becomes smooth indicating the possibility of at least two different rates of drop in liquid level. The objective of this method was to determine 1) the two different rates, if possible, and 2) the fraction of large bubbles present in the column.

The results show that the fraction of large bubbles increases with gas velocity and that gas velocity determines the size distribution of bubbles as shown in Figures 11 through 15. Liquid velocity has no effect on the size distribution of bubbles. In these figures, the bars indicate that the liquid level was not steady and represents the lower and higher values observed at that time from the photograph. Going through a series of photographs taken at one second intervals one can see the change that takes place in the liquid level after a certain time. Figures 16 and 17 are two typical photographs taken at three seconds and twenty-one seconds, respectively, after the valve was closed. In Figure 16 the liquid level was not horizontal but clearly very turbulent. Twenty-one seconds after the valve was closed, the liquid level was still dropping; however, as can be seen from Figure 17, the liquid level was clearly horizontal.

Bubble size distribution, as well as gas holdup, are independent of liquid velocity as illustrated in Figures 12 through 15. A change in liquid velocity from 0.02 to 0.5 ft/sec did not change the rate of liquid level drop nor the final liquid level. This liquid velocity independence behavior was observed at two different gas velocities (0.25 ft/sec and 0.43 ft/sec) suggesting a general behavior rather than coincidence.

The fraction of large bubbles in the column increases with gas velocity as indicated in Figures 11, 12 and 14. At very low gas velocities (0.05 ft/ sec), one could not observe the transition between the two distinct regions due to the obstruction caused by the connecting flanges in the column. However, Figure 11 suggests that the fraction of large bubbles in the column, if any, is very small at a gas velocity of 0.05 ft/sec. Figures 12 and 14 indicate that an increase in gas velocity from 0.25 ft/sec to 0.43 ft/sec resulted in an increase in the fraction of large bubbles. In Figure 14 (and 15) one can clearly see that two

FIGURE 11 – RATE OF THE AERATED LIQUID LEVEL DROP

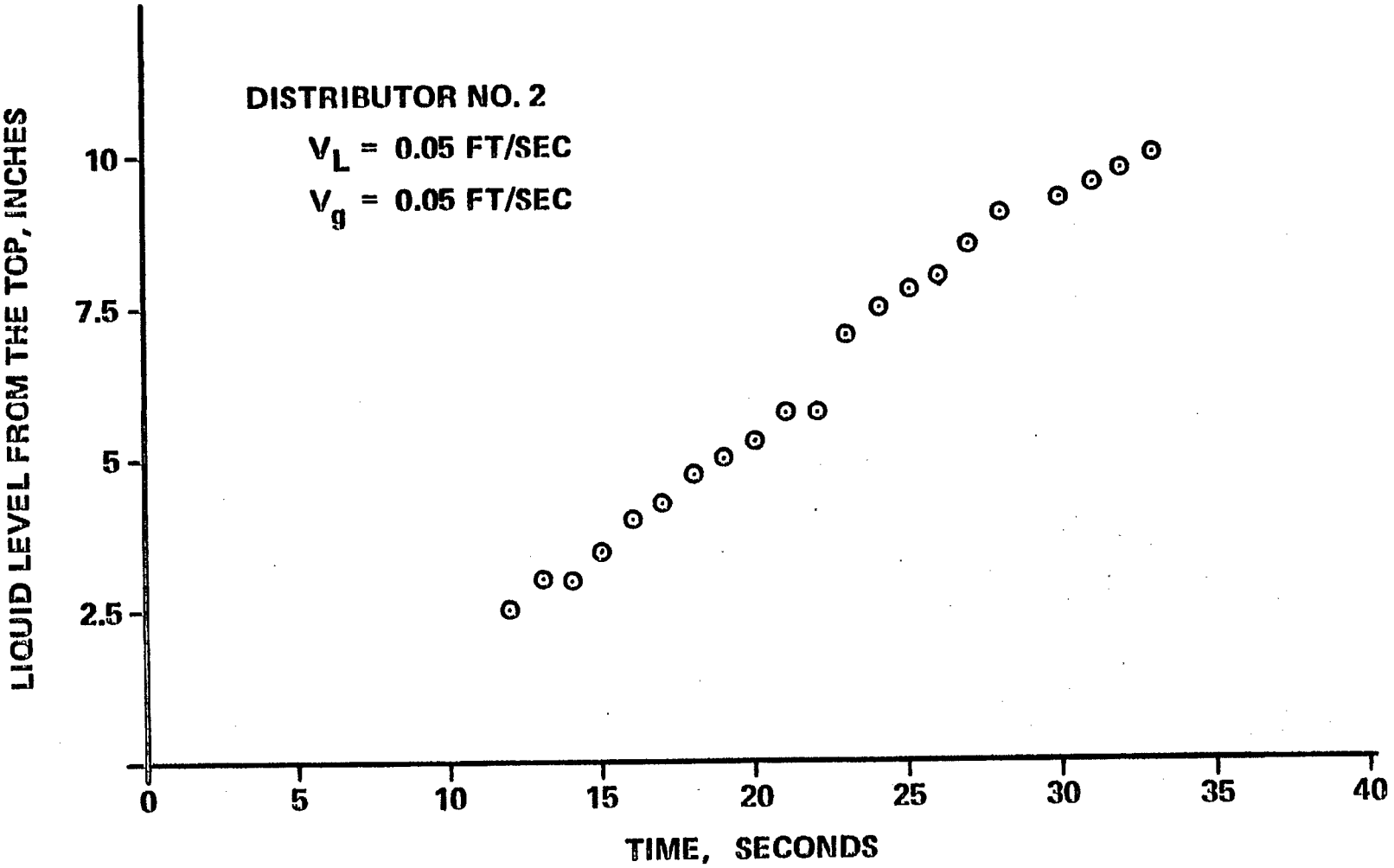


FIGURE 12 – RATE OF THE AERATED LIQUID LEVEL DROP

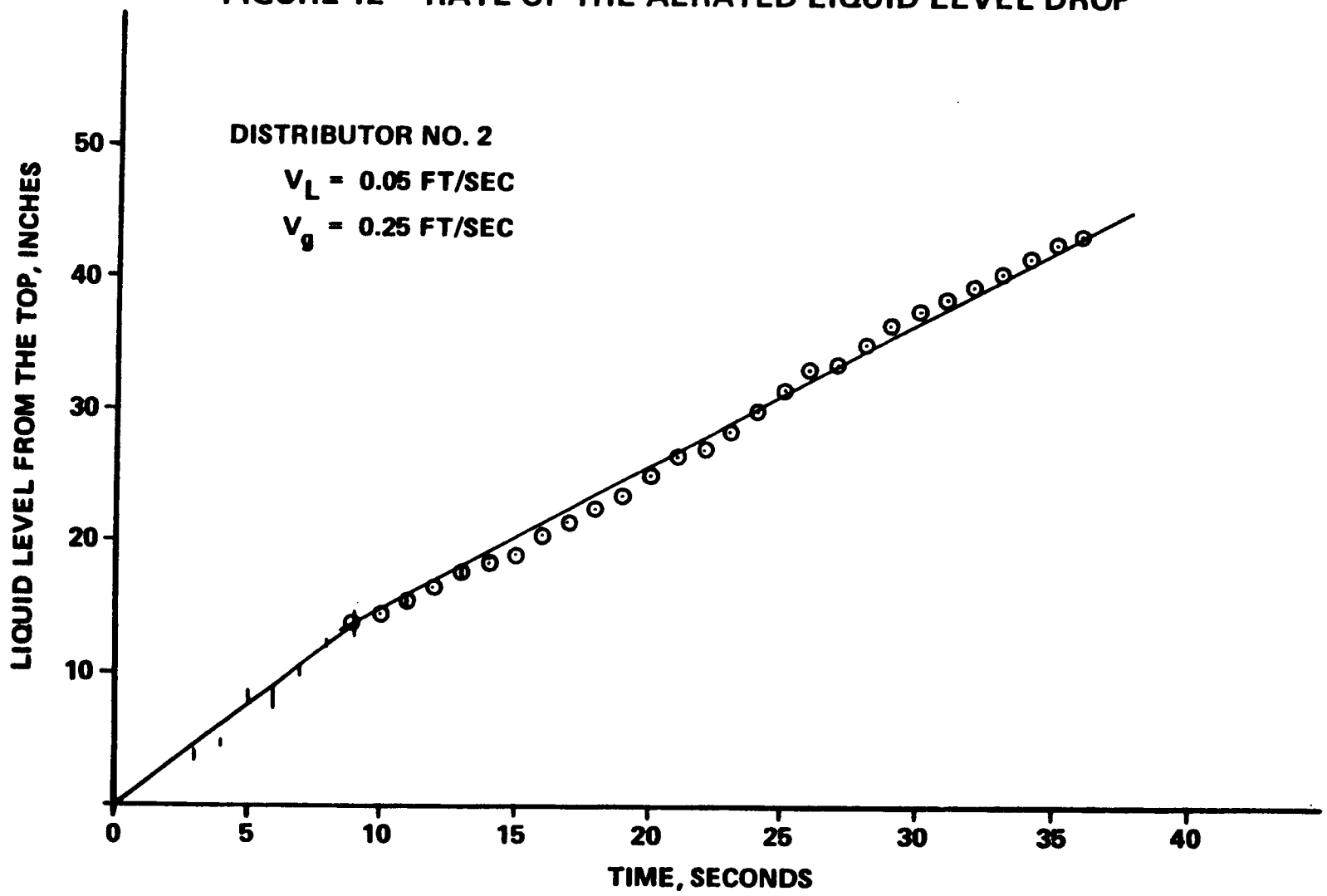


FIGURE 13 – RATE OF THE AERATED LIQUID LEVEL DROP

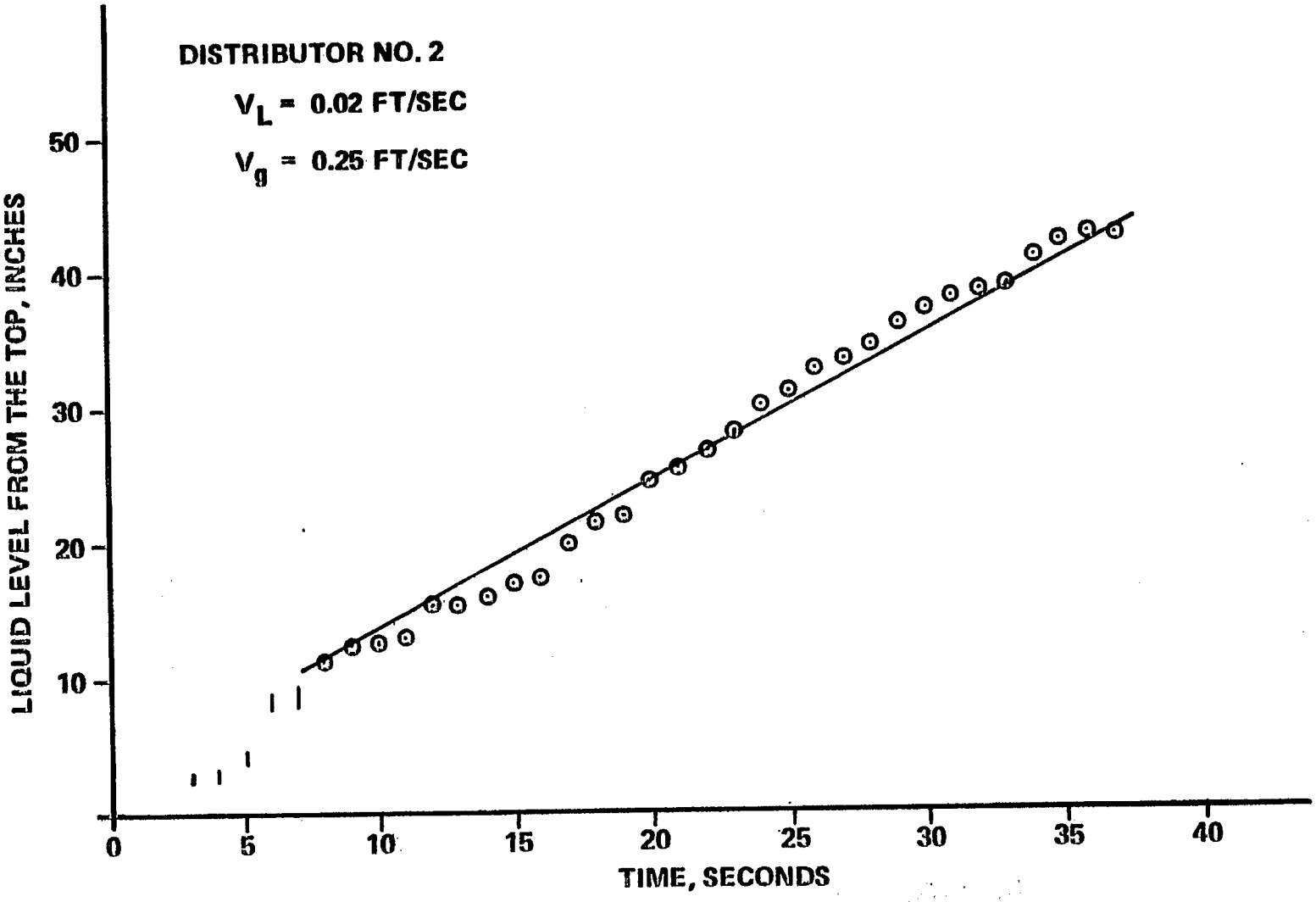


FIGURE 14 – RATE OF THE AERATED LIQUID LEVEL DROP

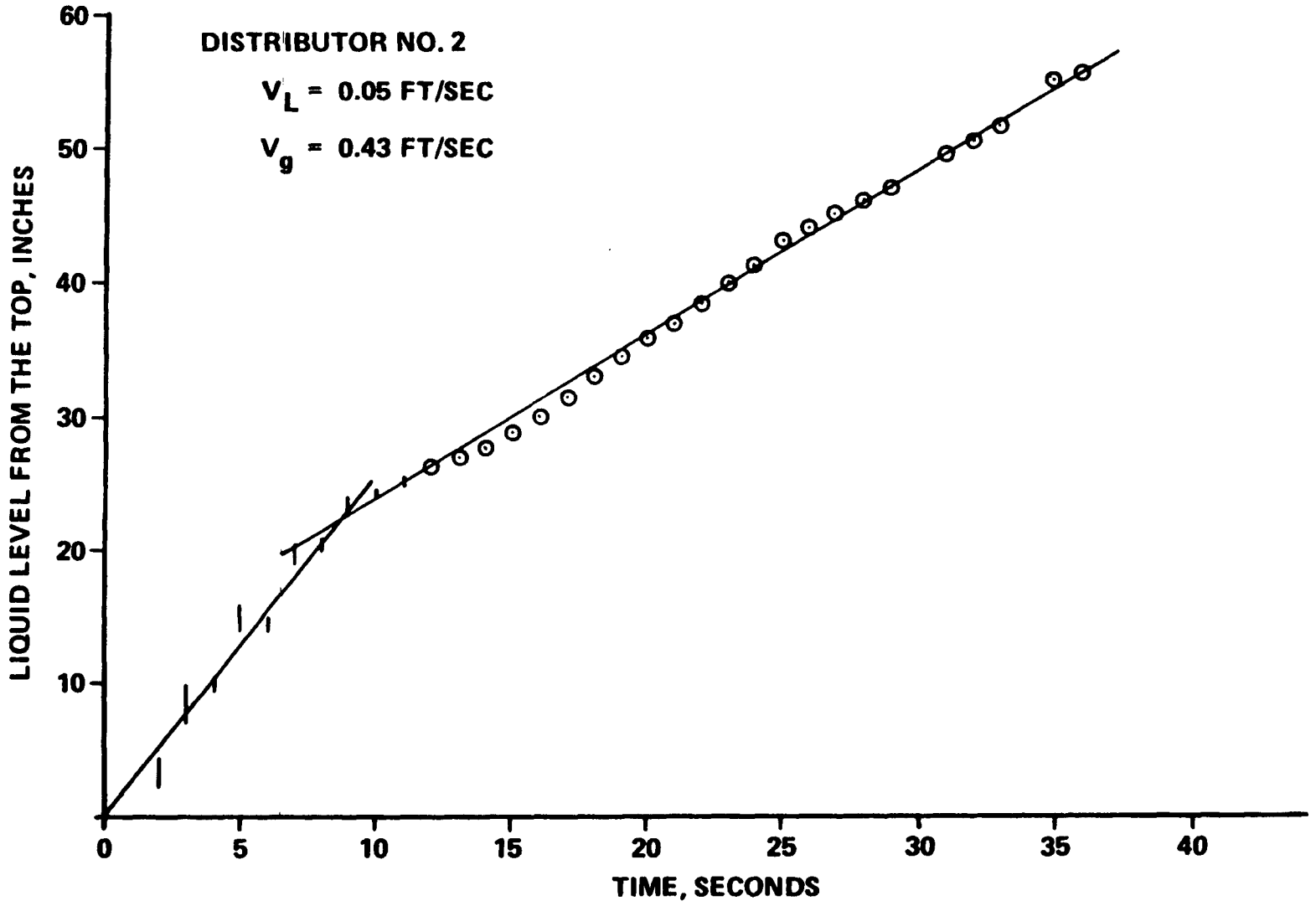
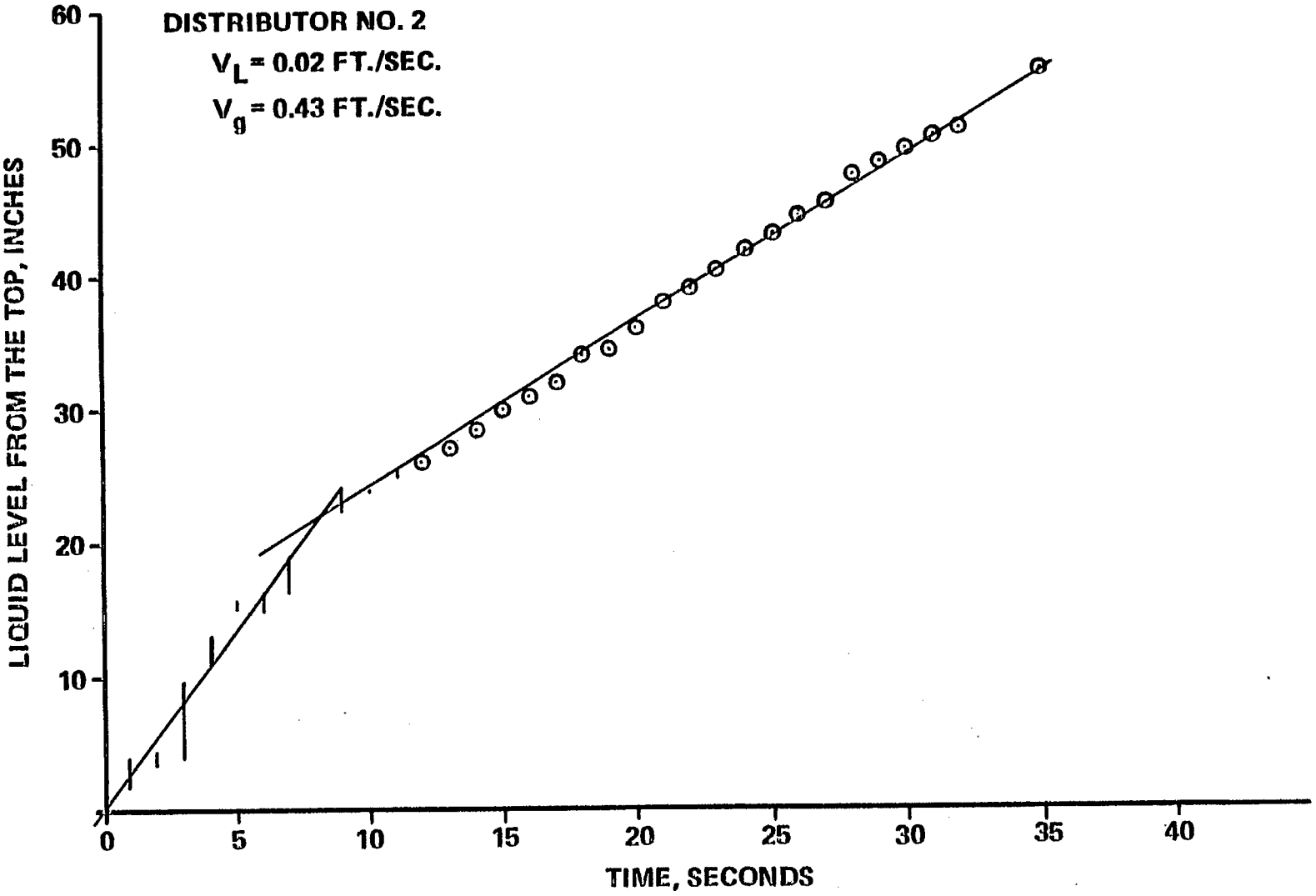


FIGURE 15 – RATE OF THE AERATED LIQUID LEVEL DROP.



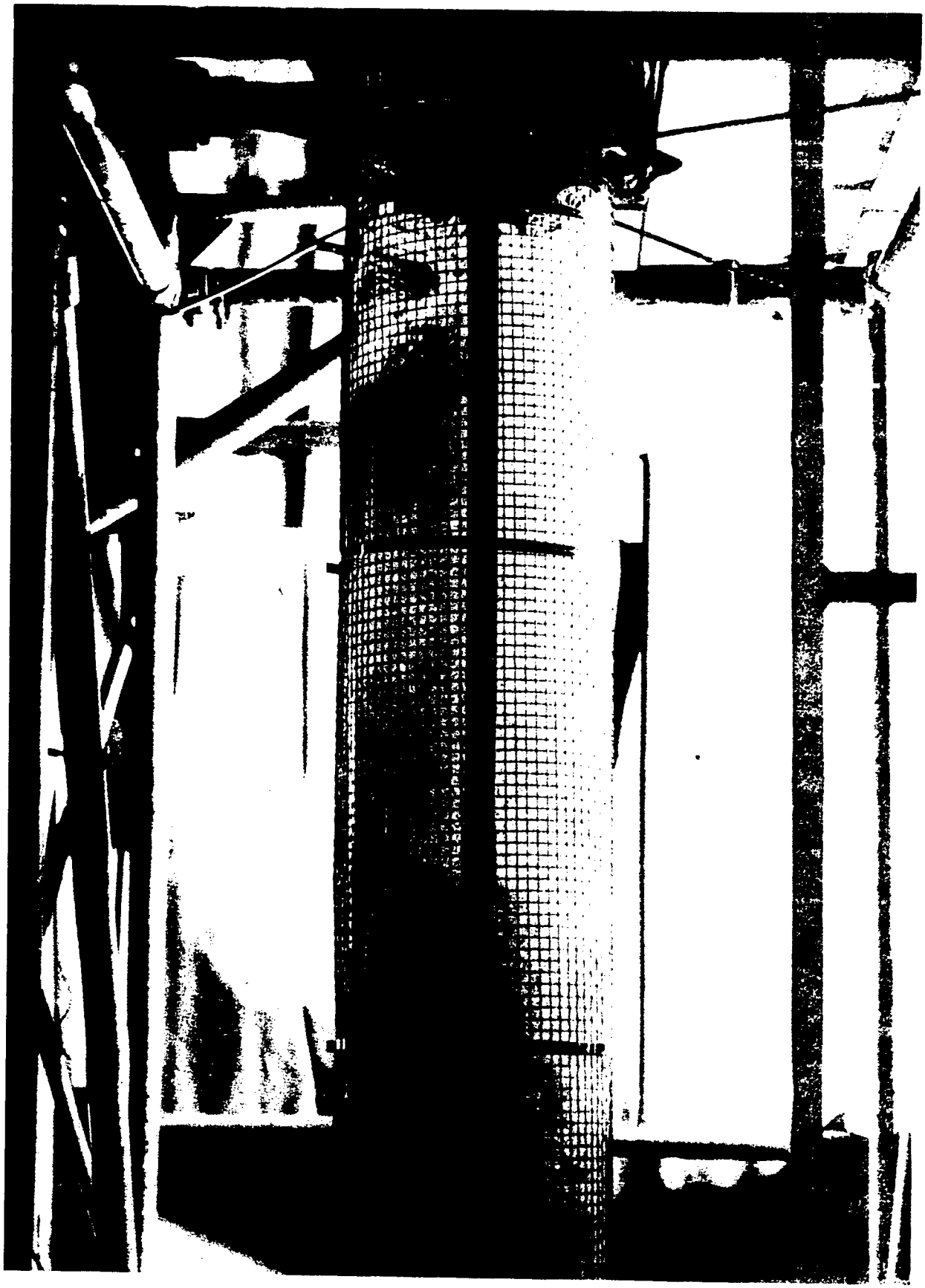


FIGURE 16

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

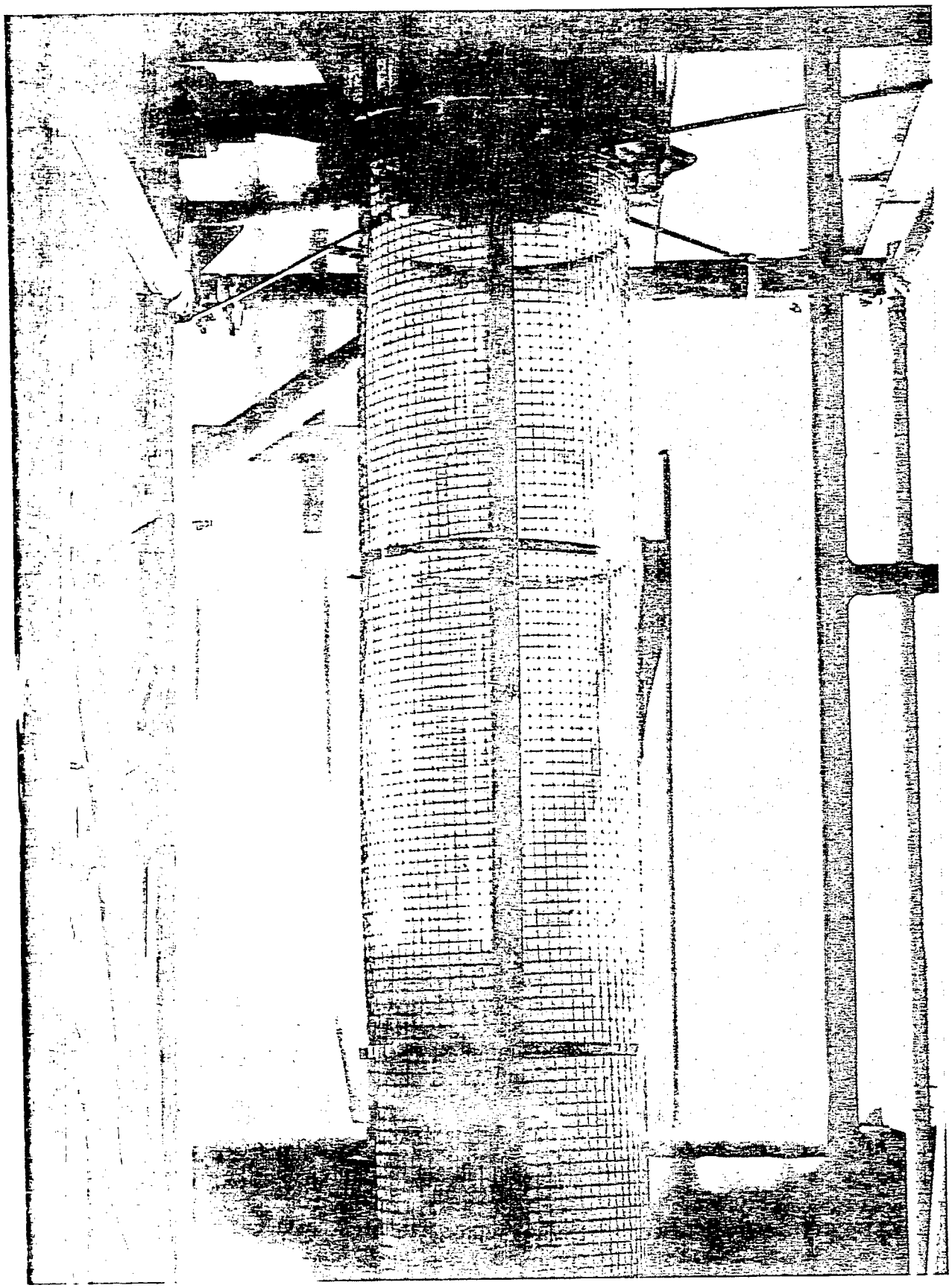


FIGURE 17

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

different rates exist; the large bubbles occupied approximately 40% of the void volume. However, in Figure 12 (and 13) one cannot distinguish between the two rates very clearly. In order to determine the fraction of large bubbles present in the column in this case, one should consider the visual observation of the transition from unsteady to steady liquid level. Using this information, qualitatively, one can say that about 32 percent of the void volume was occupied by larger bubbles.

A comparison of Figures 12 and 14 (or 13 and 15) also indicate that even though the fraction of the large bubbles in the column increase with gas velocity, the size of the smaller bubbles (a group of bubbles of about uniform size) may be unaffected by the increase in gas velocity. When Figure 12 is superimposed on Figure 10, one can see that the two figures are parallel to each other; this indicates that the bubble rise velocities are identical in both cases.

Since these results were very encouraging, this technique was employed with distributor No. 1 in place in order to distinguish between the two different distributors. Results will be presented in the next quarterly report.

5.3 Gas/Liquid Mass Transfer

Batch experiments (zero liquid flow) in the 12-inch diameter column were conducted in this quarter to study oxygen transfer from air to water in air/water and air/water/sand systems with both distributors (No. 1 and 2). Mass transfer experiments in the absence of a distributor were conducted in the last quarter and results were presented in the last quarterly report (3). Detailed description of the method and data analysis can be found in the third quarterly report (3).

The mass transfer coefficient, $k_L a$, is found to be independent of column height for the 12-inch diameter column supporting the earlier data obtained from the 5-inch diameter column. The liquid phase mass transfer coefficient is independent of the presence or absence of a distributor and the type of distributor used as shown in Figures 18 and 19. This observation is similar to gas holdup measurements.

FIGURE 18 -- EFFECT OF DISTRIBUTOR PLATE ON THE MASS TRANSFER COEFFICIENT IN THE ABSENCE OF SOLIDS

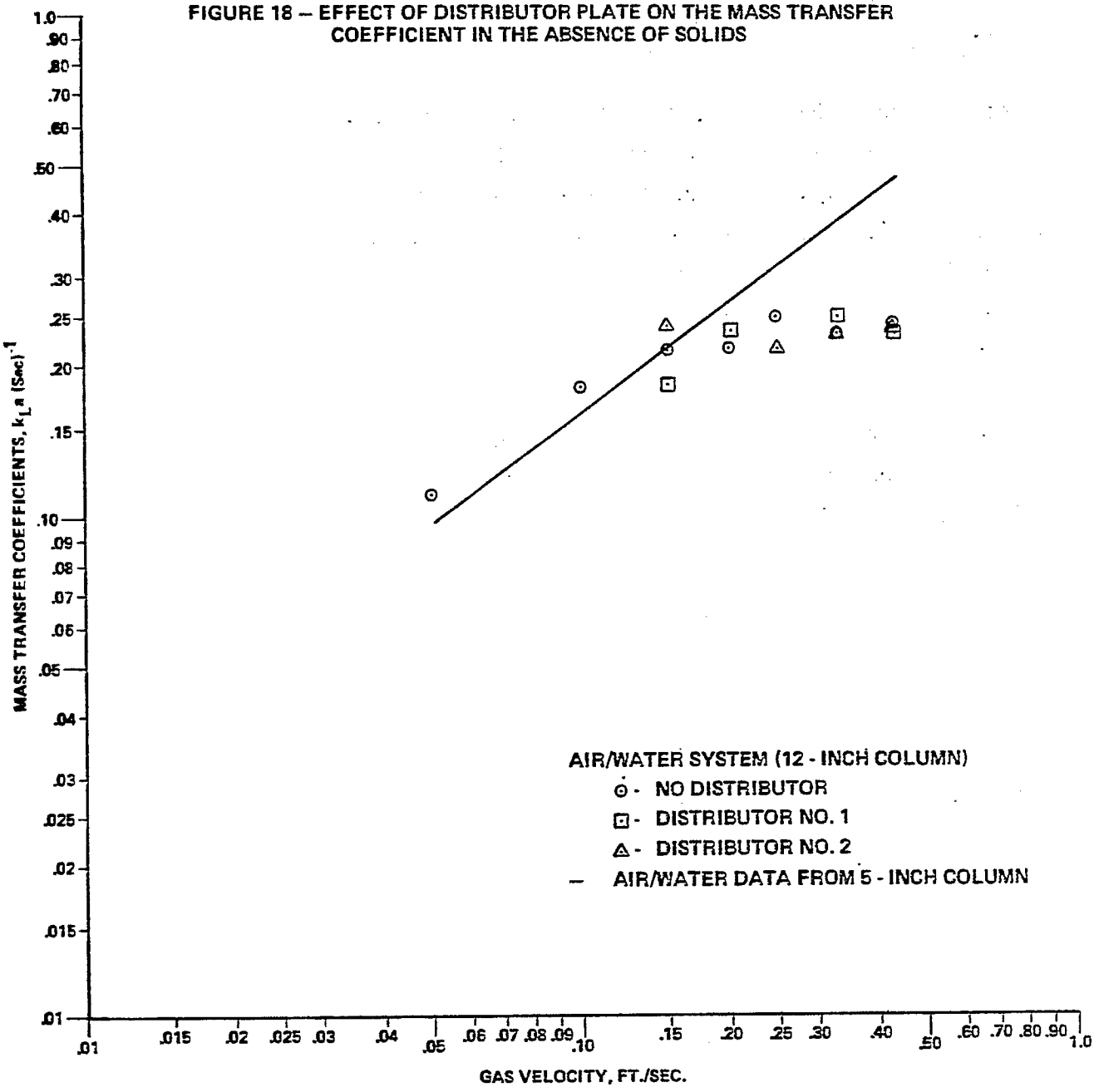
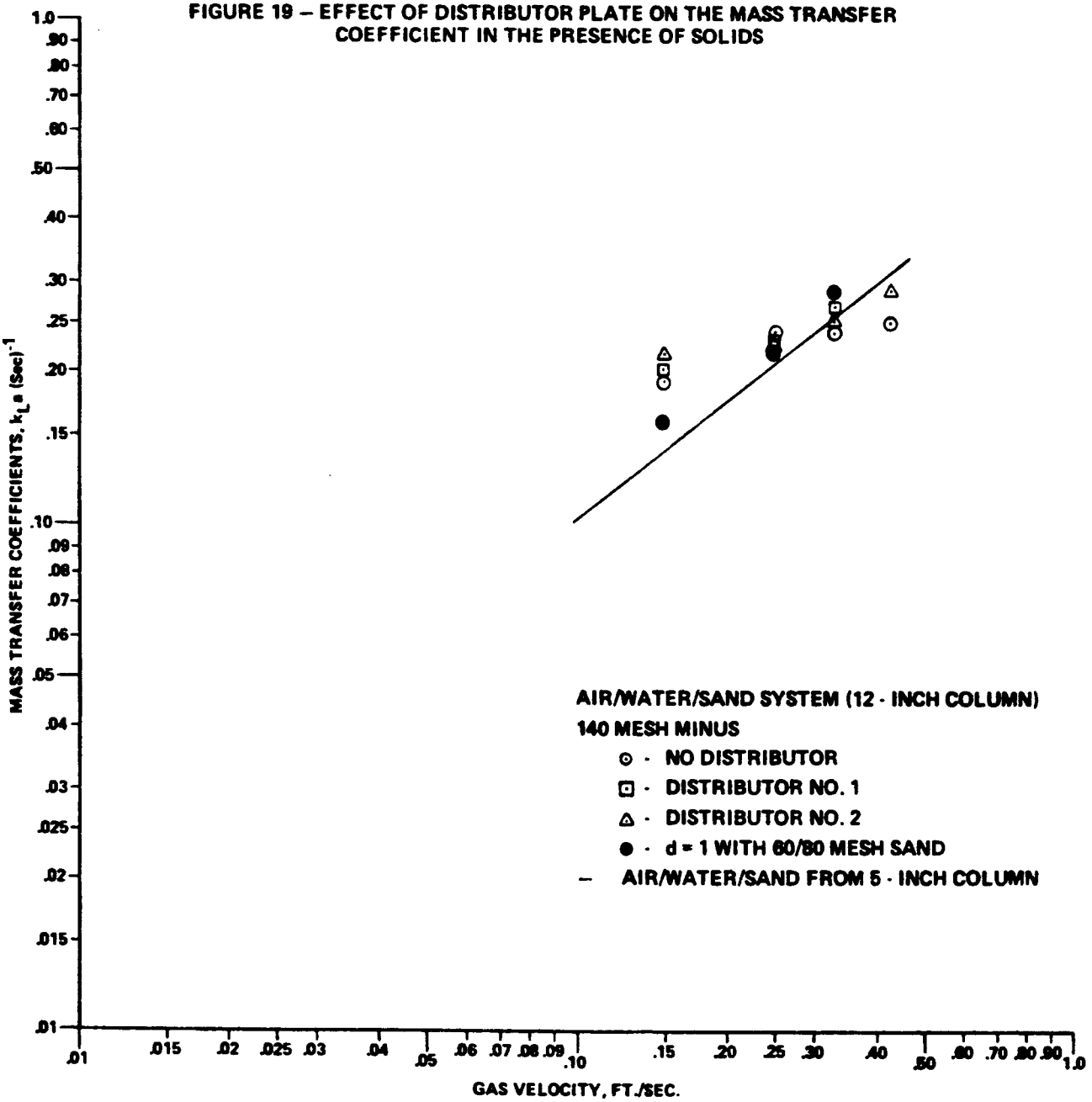


FIGURE 19 – EFFECT OF DISTRIBUTOR PLATE ON THE MASS TRANSFER COEFFICIENT IN THE PRESENCE OF SOLIDS



Increasing the column diameter imposes a dramatic effect on liquid phase mass transfer coefficient. The mass transfer coefficient results obtained from the 12-inch diameter column are distinctly different from that obtained from the 5-inch diameter column. In the 5-inch diameter column, the presence of solids resulted in a definite decrease in the value of the mass transfer coefficients. However, in the 12-inch diameter column, the presence of solids does not seem to affect the mass transfer coefficients. Also, the increase in the values of $k_L a$ with increasing gas velocity is less steep for the 12 inch column than for the 5-inch diameter column.

The leveling off in values of $k_L a$ at high gas velocities was due to the insensitive change of interfacial area with gas flow rate. In order for $k_L a$ to remain constant, the interfacial area should remain constant or very insensitive to increasing gas velocity. Holdup data indicate that gas holdup increases with increasing gas velocity. This should result in an increase in interfacial area. However, this increase is balanced by the larger fraction of big bubbles at higher gas velocities. Therefore, the interfacial area remains constant at high gas velocities, thereby accounting for the leveling off in the values of $k_L a$.

5.4 Liquid Dispersion

The experimental conditions for the liquid dispersion runs conducted in this quarter are listed in Table I. All the runs in the quarter were conducted with distributor No. 2 in place. The fits of all the experimental tracer curves with the theoretical curves from the axial dispersion model are shown in Appendix A. In these curves, the circles represent the experimental tracer curves while the solid line represents the theoretical curve obtained from the axial dispersion model. Detailed description of the procedure can be found in the first quarterly report (1). The dispersion numbers and axial dispersion coefficients for all the experiments conducted in this quarter are listed in Table II.

Table I

Experimental Conditions for Liquid Dispersion Runs

Liquid Velocity - 0.02-0.05 ft/sec

Gas Velocity - 0.05-0.43 ft/sec

Particle Size - 20/30 mesh, 140 mesh minus

Reactor Solids Concentration - 3.7-20 lbs/ft^3

Distributor Used - Number 2

Column Diameter - 12"

Table II

Data from Liquid Dispersion Experiments

With Distributor Number 2

Run No.	V_L ft/sec	V_g ft/sec	P_s Mesh	C_s lbs/ft^3	$\frac{D}{V_1 L}$	^2D ft^2/sec
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
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	0.049	0.03	-140	3.74	0.33	0.408
XXVIII- 1	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
XXIX- 1	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

Results from this quarter (using distributor No. 2) support the earlier conclusion that liquid velocity has no effect on liquid axial dispersion coefficients. Table III shows the effect of liquid velocity on axial dispersion coefficients for both the presence and absence of solid particles. The experimental data obtained in the absence of solids show a large amount of scatter; however, the data obtained in the presence of solids clearly indicate that axial dispersion coefficients were independent of liquid velocity at all conditions.

Liquid axial dispersion coefficients determined with distributor No. 2 in place in the 12 inch column increased with increasing gas velocity and decreased when solids were introduced into the system. These observations were similar to the results obtained from earlier experiments using distributor No. 1. The effect of gas velocity and solids on axial dispersion coefficients from the 12-inch diameter column in the presence of distributor No. 2 is summarized in Table IV. Disregarding some scatter in the data, Table V clearly shows that increasing gas velocity results in an increase in axial dispersion coefficients and the presence of solids results in a decrease in liquid axial dispersion coefficients.

Liquid axial dispersion coefficients in the 12-inch diameter column are independent of entrance effects as shown in Figures 20 through 22. These figures compare the results obtained from the three different distributor configurations used in this study (absence of a distributor, a distributor with bubble caps and a distributor without bubble caps). The data in the absence of a distributor should be treated with some caution. Because of the experimental technique used in injecting the tracer (which was normally injected just above the base plate), the boundary conditions will be different for the "no distributor" case. Therefore, using the same curve matching technique will generate larger uncertainties in the case of no distributor. Figure 20 shows that at low gas velocities, data from the "no distributor" experiments do not fall in line with others. These figures clearly show that dependence of axial dispersion coefficients on linear gas velocities can be described fairly well by a straight line on a log-log plot; this indicates that

Table III

Effect of Liquid Velocity on Axial Dispersion Coefficients

Using Distributor Number 2 in the 12" Column

V_L ft/sec	V_g ft/sec	No Solids	Axial Dispersion Coefficients, ft^2/sec 20/30 Mesh C_s=5 lbs/ft^3	20/30 Mesh C_s=20 lbs/ft^3	140 Mesh Minus C_s=3.7 lbs/ft^3	140 Mesh Minus C_s=14.9 lbs/ft^3
0.02	0.33	0.540	0.429	0.371	0.454	0.339
0.04	0.33	0.609	0.431	0.322	0.418	0.355
0.05	0.33	0.479	0.403	0.363	0.408	0.404

Table IV

Effect of Gas Velocity and Solids on Axial Dispersion Coefficients Using Distributor Number 2 in the 12" Column

V_L ft/sec	V_g ft/sec	Axial Dispersion Coefficients, ft^2/sec				
			20/30 Mesh		140 Mesh Minus	
		No Solids	C_s=5 lbs/ft^3	C_s=20 lbs/ft^3	C_s=3.7 lbs/ft^3	C_s=14.9 lbs/ft^3
0.05	0.05	0.361	0.240	0.230	0.238	0.273
0.05	0.10	0.424	0.401	0.242	0.366	0.433
0.05	0.20	0.485	0.386	0.327	0.443	0.437
0.05	0.33	0.479	0.403	0.363	0.408	0.404
0.05	0.43	0.615	0.558	0.424	0.545	0.408

FIGURE 20 – EFFECT OF DISTRIBUTOR PLATE ON AXIAL LIQUID DISPERSION COEFFICIENT IN THE ABSENCE OF SOLIDS

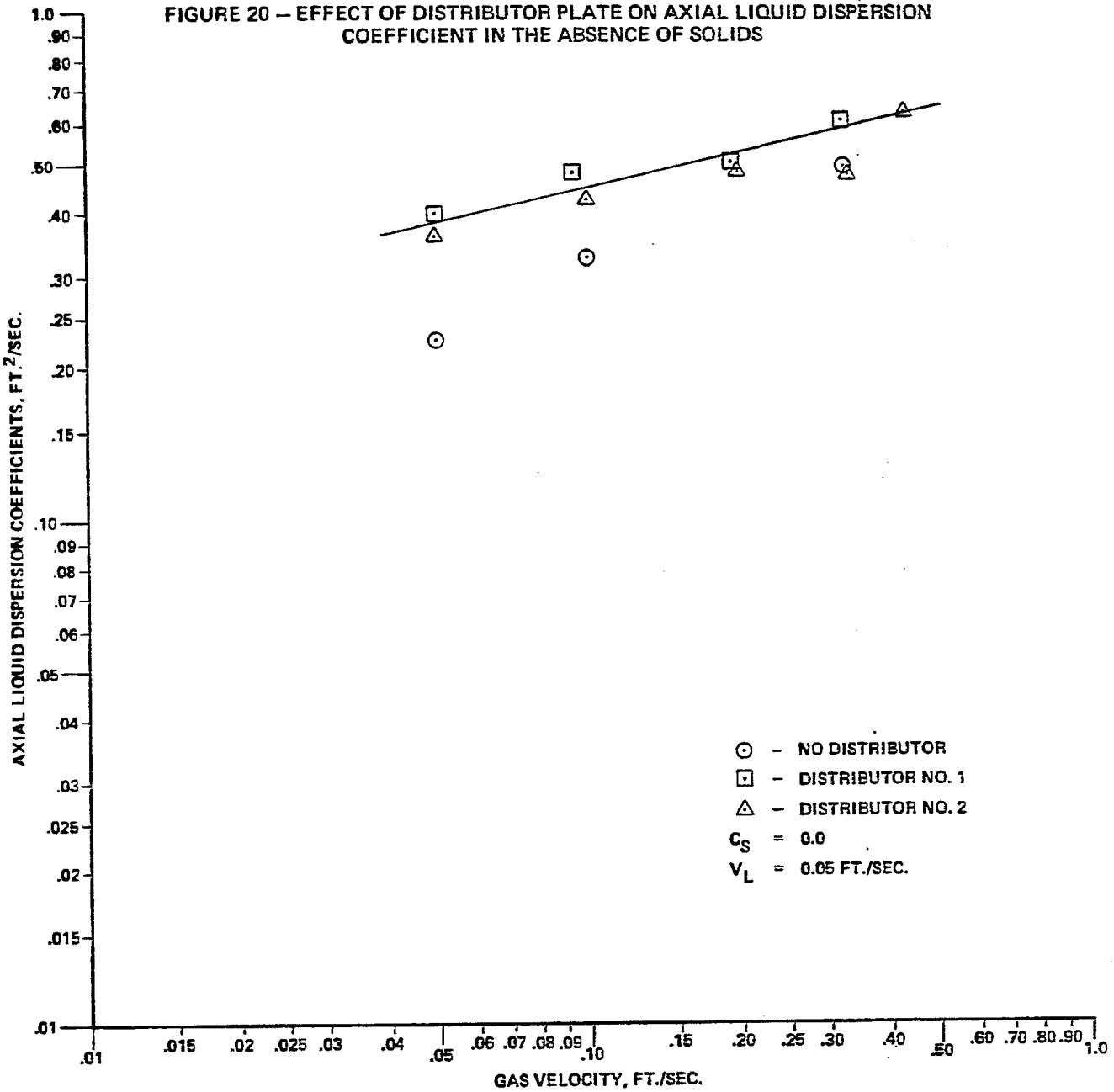


FIGURE 21 – EFFECT OF DISTRIBUTOR PLATE ON AXIAL LIQUID DISPERSION COEFFICIENT IN THE PRESENCE OF LOW CONCENTRATION OF LARGE SOLID PARTICLES

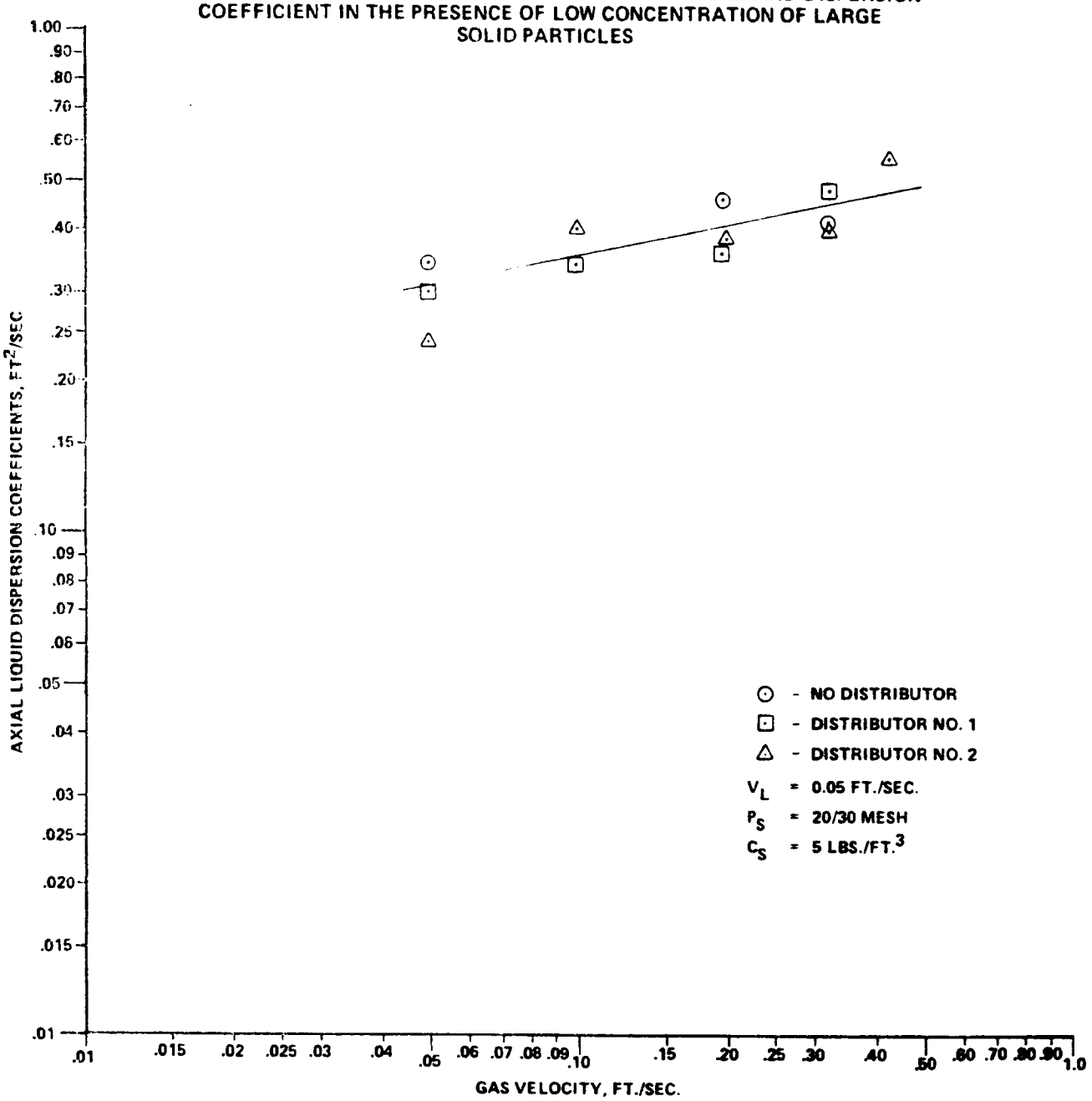
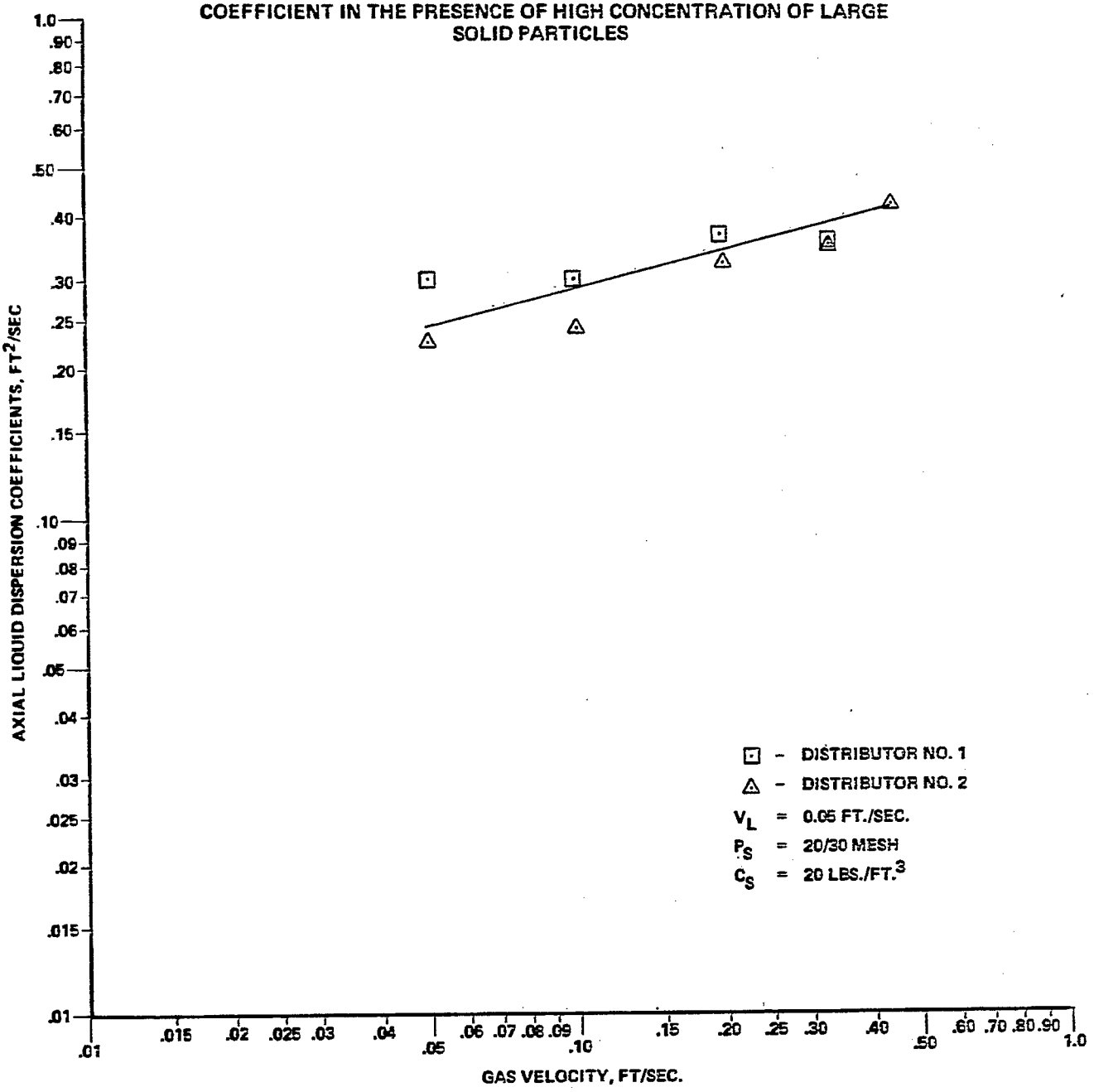


FIGURE 22 – EFFECT OF DISTRIBUTOR PLATE ON AXIAL LIQUID DISPERSION COEFFICIENT IN THE PRESENCE OF HIGH CONCENTRATION OF LARGE SOLID PARTICLES



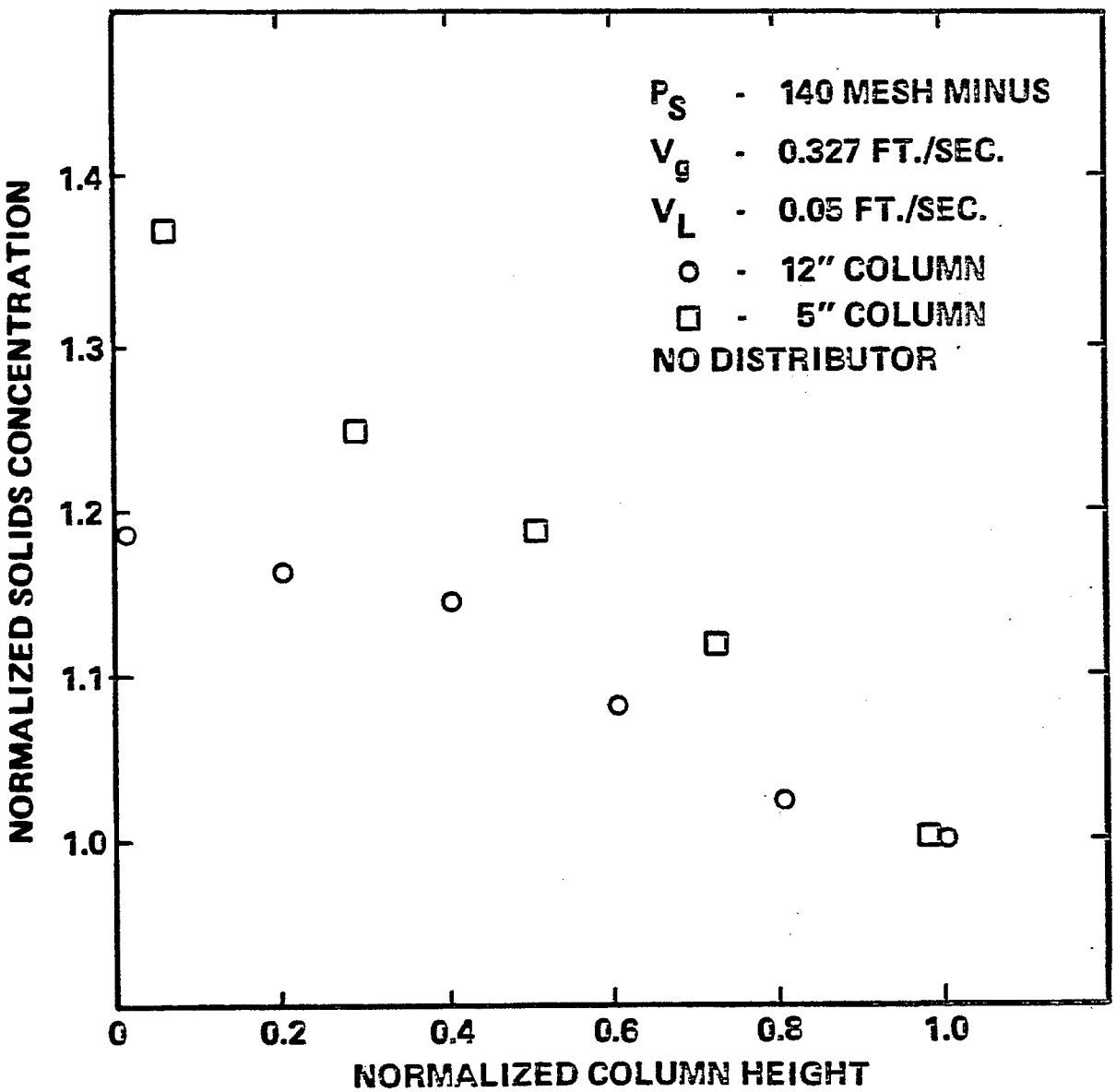
axial dispersion coefficients are proportional to $(V_g)^x$, where V_g is the linear gas velocity and x is an exponent. When one superimposes the three figures (20, 21 and 22) it is clear that the three figures are closely parallel to each other, thereby indicating that the dependence of axial dispersion coefficients on linear gas velocities is not affected by the presence of solids. This superimposition also clearly shows that the axial dispersion coefficients decreases with increasing solids concentration.

5.5 Solids Distribution

Several experiments were made during this quarter using both 140 mesh minus and 60/80 mesh sand to study the effect of liquid and gas velocities on the axial solids distribution in the 12-inch diameter column with distributor No. 2 in place. In addition, data analysis was completed for a single run conducted during last quarter using 140 mesh minus sand in the absence of a distributor. The analysis of the effect of the different distributors on solids distribution is not complete yet. Hence a discussion of the entrance effects on axial solids distribution will be deferred until the next quarterly report.

The average retained solids concentration decreases with increasing column diameter. Experiments were conducted using 140 mesh minus sand in both the 5-inch and 12-inch diameter columns in the absence of a distributor at identical operating conditions. In order to compare the effect of column diameter the normalized solids concentration was plotted against normalized height (since the 12 inch column is about 5 times taller than the 5 inch column). Figure 23 shows that the axial solids distribution profile was less steep for the 12-inch diameter column than the 5-inch diameter column suggesting that the average retained solid concentration decreases with increasing column diameter. This finding is directly in line with the earlier batch experimental results which show that the increase in column diameter may result in an increase in solid dispersion coefficient leading to a less steep solid distribution profile.

FIGURE 23 – EFFECT OF COLUMN DIAMETER ON SOLIDS CONCENTRATION



Results from the 12-inch diameter column experiments using distributor No. 2 indicate that changes in linear gas velocity does not appreciably change the axial solids distribution profile as shown in Figures 23 and 24. This is true in respective of the particle size of the sand used. Figure 23 presents results from experiments using 140 mesh minus size particles. 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 the changes in gas velocity does not appreciably affect the axial solids distribution as well as solid accumulation. Figure 25 shows that for 60/80 mesh particles, increasing gas velocity from 0.194 to 0.33 ft/sec did not change solids concentration in the bulk of the column though 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 25 illustrate the independence of gas velocity on solid accumulation beyond the critical gas velocity.

Changes in liquid velocities, however, has a definite effect on axial solids distribution as shown in Figures 26 and 27. Decreasing liquid velocity results in an increase in the solids concentration at any specific point in the column. The effect is much more profound for the larger particles even though both particle sizes exhibit the same behavior.

6.0 FUTURE WORK

Data analysis will be completed to distinguish the effect of different distributors on axial solids distribution in the column. Data analysis of the photographic technique using distributor No. 1 will be completed to compare the effect of distributors on the bubble sizes present in the column. A series of experiments to investigate the effect of the presence of large particles on the accumulation of fines will be conducted. Solid removal (Task 4) experiments will be designed and conducted.

FIGURE 24 – EFFECT OF GAS VELOCITY ON THE DISTRIBUTION OF FINE PARTICLES

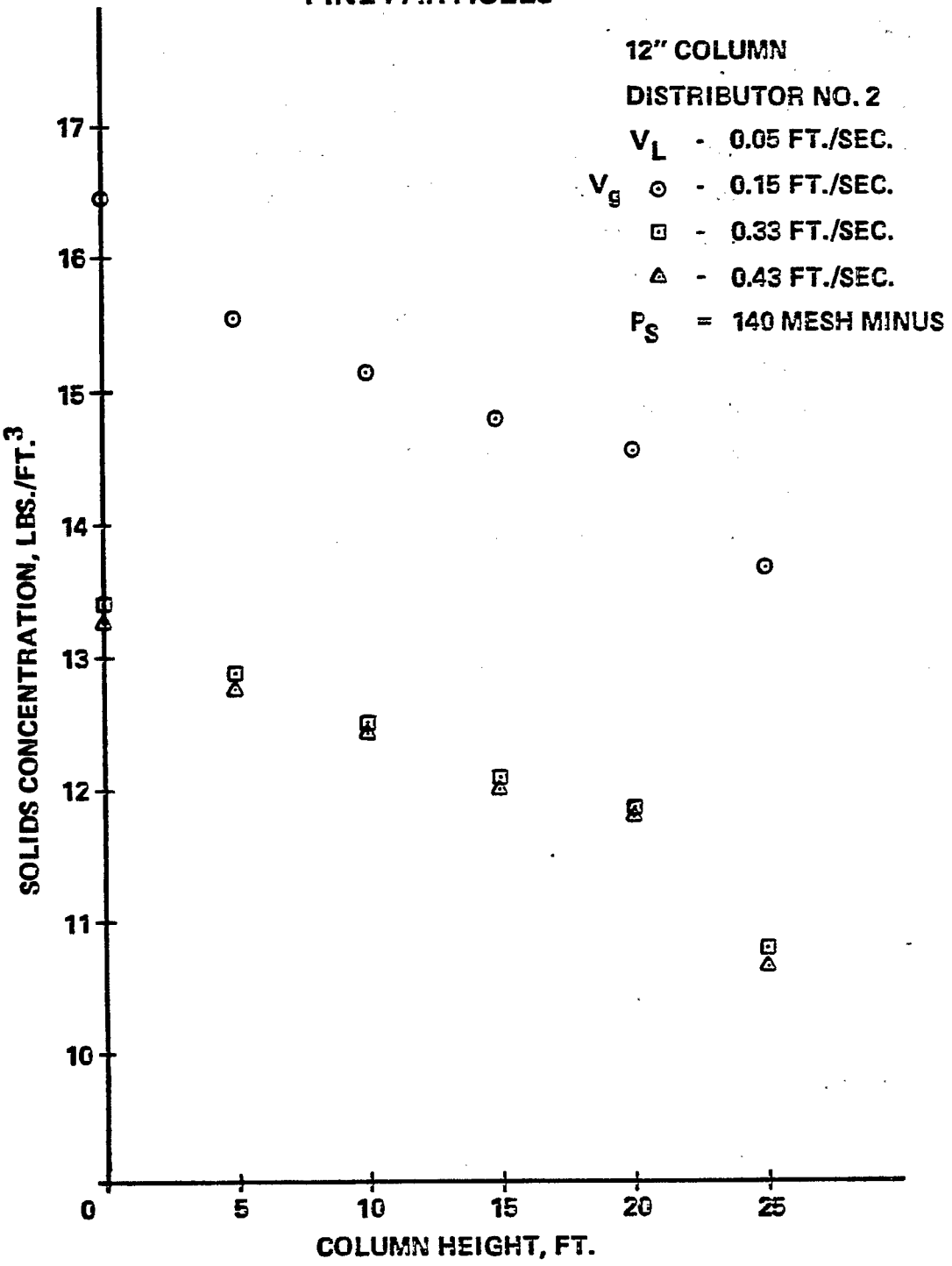


FIGURE 25 – EFFECT OF GAS VELOCITY ON THE DISTRIBUTION OF LARGE PARTICLES

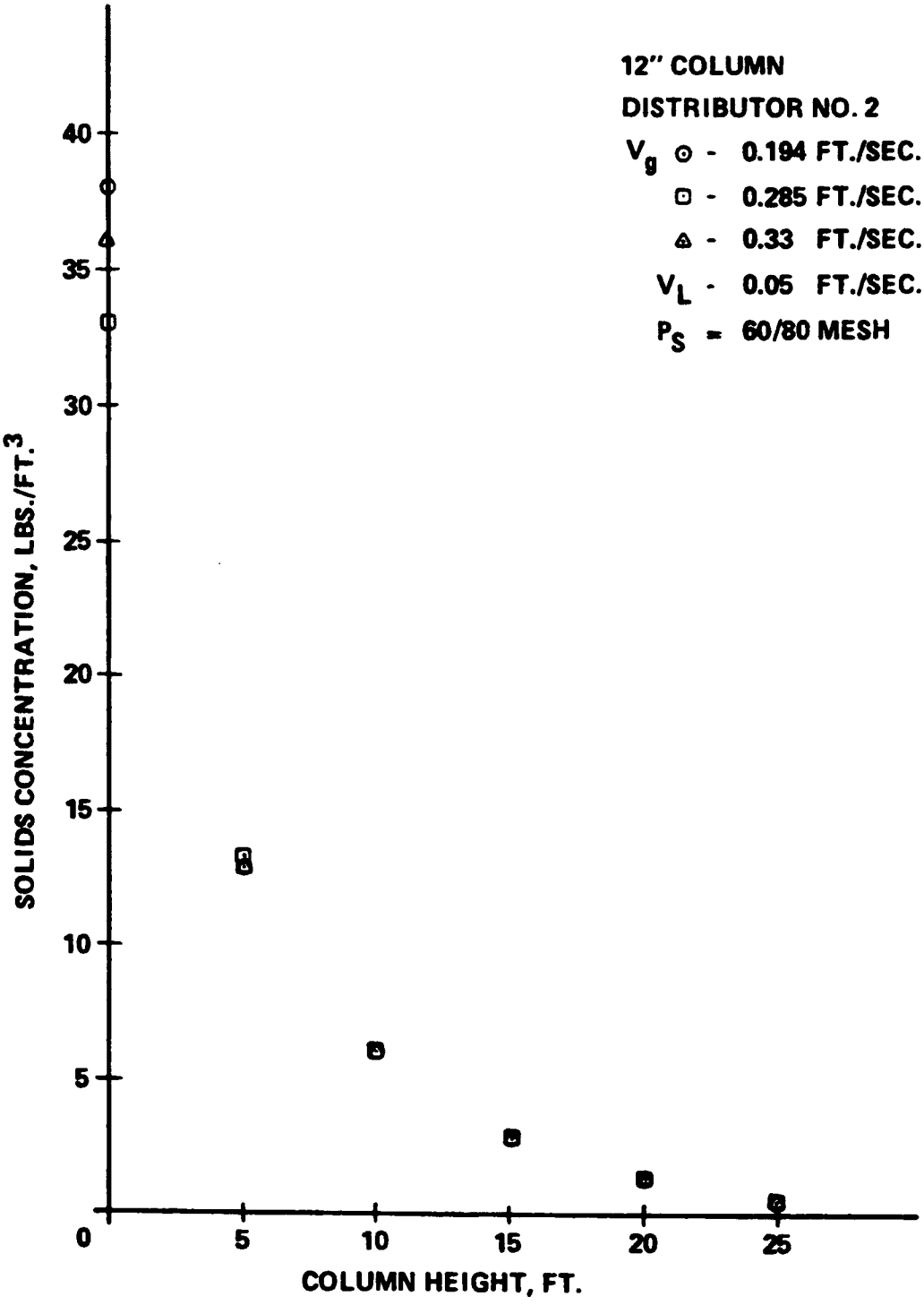


FIGURE 26 – EFFECT OF LIQUID VELOCITY ON THE DISTRIBUTION OF FINE PARTICLES

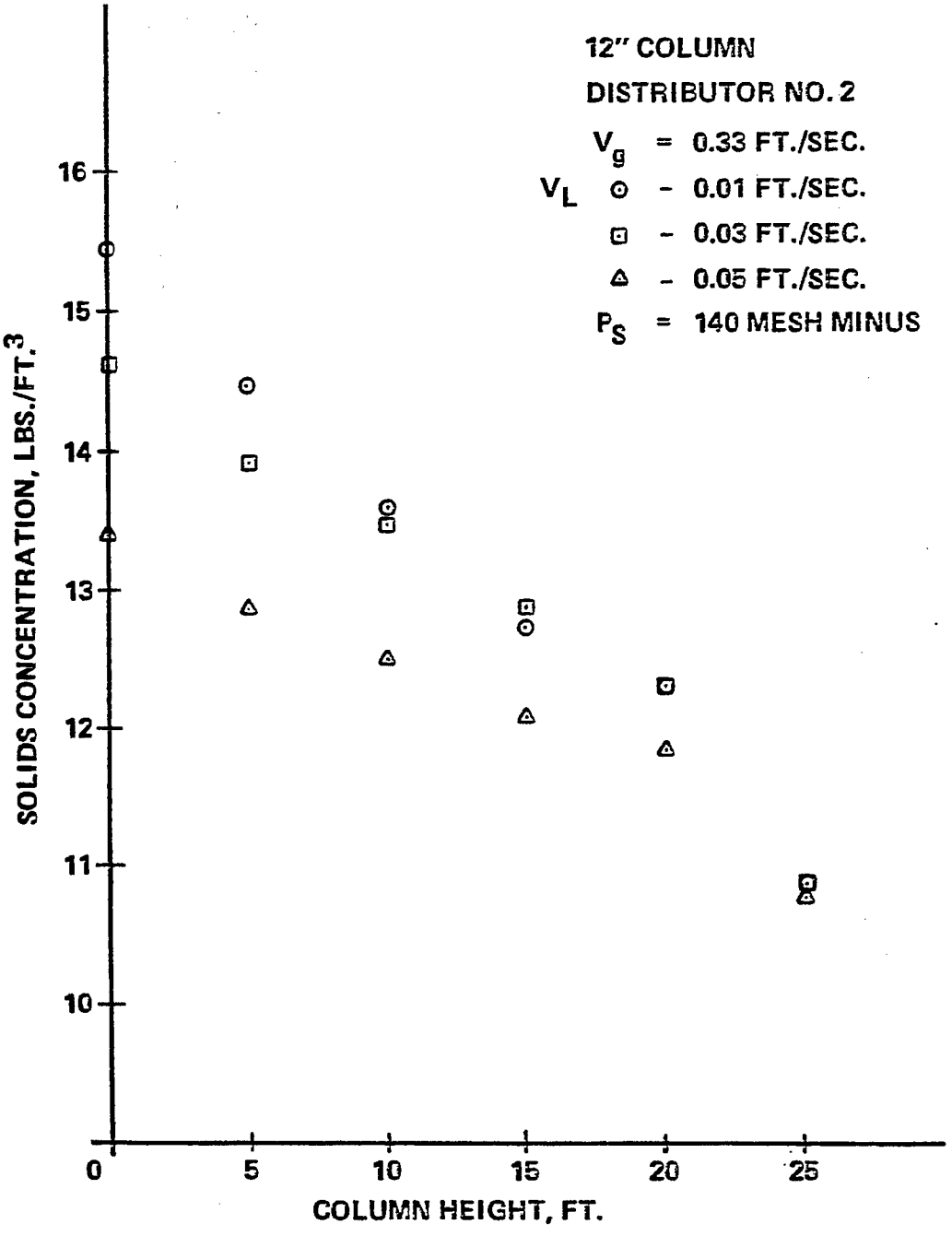
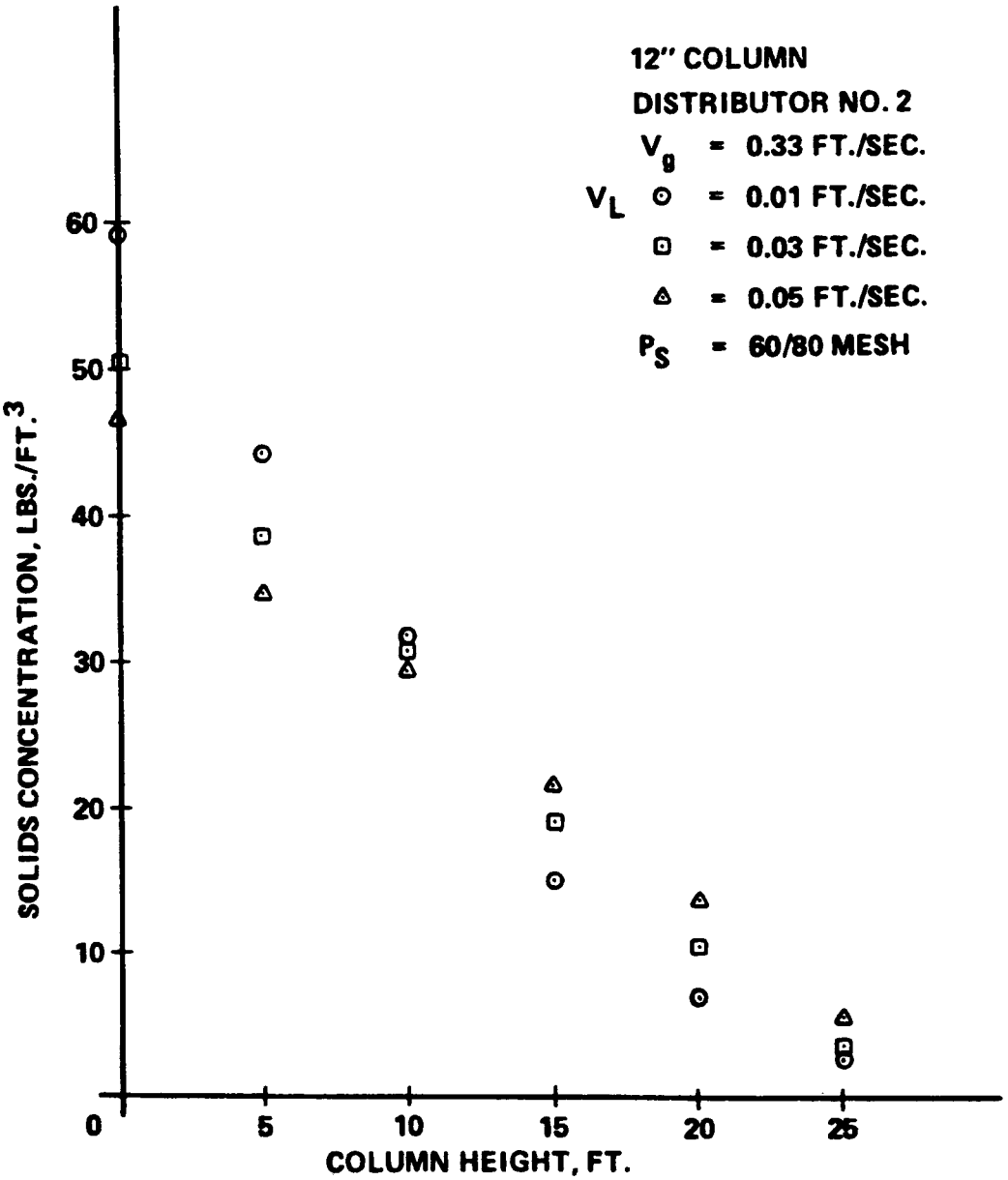


FIGURE 27 – EFFECT OF LIQUID VELOCITY ON THE DISTRIBUTION OF LARGE PARTICLES



Screening experiments using the 5 inch column will start and the construction of the enclosure for the 12-inch diameter column will be completed during the next quarter.

7.0 REFERENCES

1. "Gas/Slurry Flow in Coal Liquefaction Processes", Quarterly Technical Progress Report for period 1 October 1979-31 December 1979, FE-14801-3, January 1980.

2. "Gas/Slurry Flow in Coal Liquefaction Processes", Quarterly Technical Progress Report for period 1 January 1980-31 March 1980, FE-14801-6, May 1980.

3. "Gas/Slurry Flow in Coal Liquefaction Processes", Quarterly Technical Progress Report for period 1 April 1980-30 June 1980, FE-14801-9, August 1980.

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