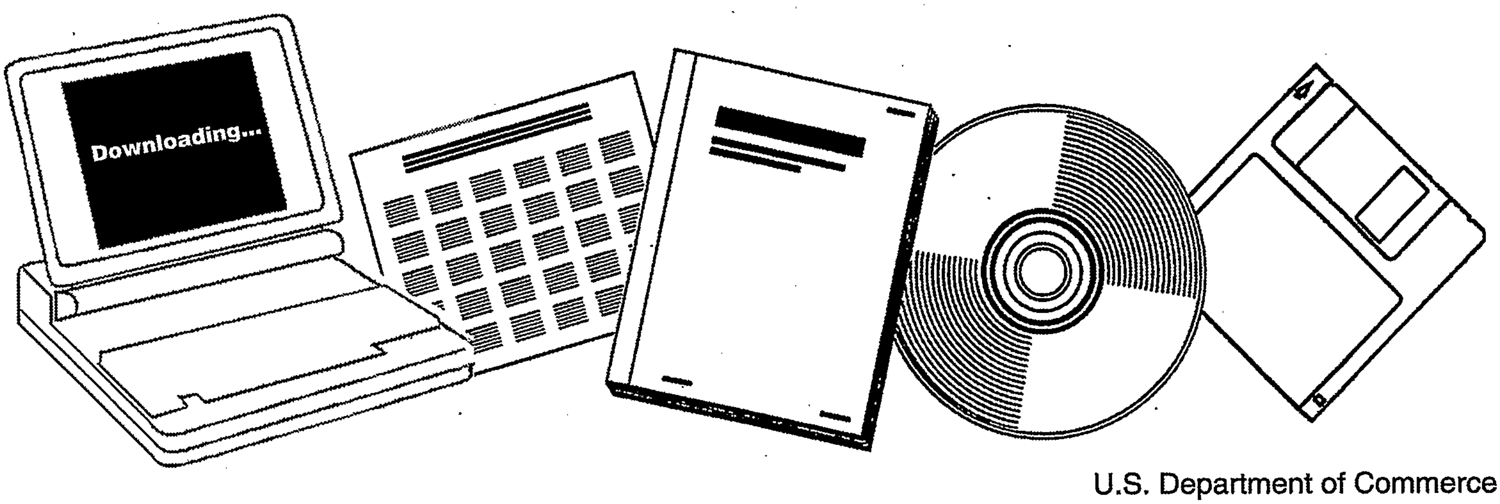
DE84010745

CONFIRMATION TEST FOR GAS/SLURRY FLOW IN SRC-I COAL LIQUEFACTION PROCESS. INTERNAL R AND D FINAL REPORT

INTERNATIONAL COAL REFINING CO.
ALLENTOWN, PA

SEP 1983



DE84010745

DOE/OR/03054-45
(DE84010745)
Distribution Category UC-89

CONFIRMATION TEST FOR GAS/SLURRY FLOW IN SRC-I COAL LIQUEFACTION PROCESS

Internal R&D Final Report

By
Samir F. Moujaes

Work Performed Under Contract No. AC05-78OR03054

International Coal Refining Company
Allentown, Pennsylvania

TABLE OF CONTENTS

ABSTRACT

The overall objective of program 12.11.1 was to provide data needed to confirm the design of the transport system, slurry heat exchangers, and slurry feed manifolds for the SRC-I Demonstration Plant. Because of lack of funds, the program was terminated before most of the work was completed. Two studies related to distribution of two-phase flow in the heat exchanger tubes were finished. A special system was designed to measure slurry concentration and flow rate in different tubes. Results showed that withdrawing slurry samples from the sides of the tubes gives a reasonably accurate measure of the concentration. Flow rate was measured indirectly with a photodiode/digital counter arrangement that measured velocity of a gas slug injected in the tube. A simple linear correlation was found to exist between the average slurry velocity and the gas-slug velocity.

EXECUTIVE SUMMARY

ICRC's program to evaluate gas/slurry flow during coal liquefaction was subdivided into four tasks: (1) determine slurry flow rate and concentration in each heat exchanger tube, in upward and downward flows; (2) study hydrodynamics and flow patterns of gas/liquid and gas/slurry mixtures in the dissolver inlet manifold and transfer lines; (3) evaluate the effect of pressure on gas/liquid two-phase flow in SRC-I transfer lines; (4) assess the effectiveness of the slurry feed pump manifold design. Unfortunately, due to a lack of funds, the project was terminated at the end of October 1982.

At that time, most of the components needed for the heat exchanger tubes for Task 1 were either manufactured or purchased, and were being installed. The objective of the task was to study the detailed distribution of two-phase (liquid/solid) slurry flow, in upward and downward configurations, in the tubes of the shell-and-tube slurry heat exchanger. To do so, a study was initiated to design a special system to measure slurry concentration and flow rate in representative exchanger tubes. Withdrawing slurry samples from the sides of the tubes was found to give a reasonably accurate measure of the concentration. To measure the flow rates, a unique photodiode/digital counter arrangement was developed that could measure the slurry flow rate indirectly by measuring the velocity of a gas slug injected into the tube. A simple linear correlation was found between the average slurry velocity and the gas slug velocity.

All equipment and materials were purchased for Tasks 1 and 2, and construction was 70 and 50% complete, respectively, at the time the program was terminated. Tasks 3 and 4 did not progress beyond the planning stage.

INTRODUCTION

Studies of the physical behavior of two- and three-phase flow systems in transport lines and vessels such as heat exchangers are required to confirm the SRC-I coal-liquefaction plant design. The most important physical behavior of such systems is their flow pattern. All the key design parameters, including heat-transfer rate, pressure drop, liquid holdup, and fluid distribution, depend on the fluid flow pattern.

Although the current SRC-I plant design may not necessarily be changed as a result of studying physical flow behavior, an understanding will certainly improve plant efficiency within the constraints of the design. Because pertinent data to define the behavior of two- and three-phase systems in horizontal, vertical-upward, and vertical-downward flow were lacking, International Coal Refining Company (ICRC) designed a research program to perform a variety of modeling studies on flow behavior. This 21-month experimental program was intended to provide data that are required to confirm the design of the transport systems, slurry heat exchangers, and slurry feed manifolds.

Specific data requirements were divided into four major tasks:

1. Heat Exchanger Study: Determine the effect of fluid viscosity, particle properties, and slurry concentration on slurry distribution and heat exchanger performance, for both upward and downward two-phase solid/liquid slurry flow.

2. Transfer-Line Study: Determine the effect of the piping configuration on flow behavior by studying hydrodynamics and flow patterns of gas/liquid and gas/liquid/solid systems in the transfer lines from the preheater to the dissolver (dissolver inlet manifold) and from the dissolver to the separator.

3. Pressure Study: Determine the effect of pressure on two-phase gas/liquid flow behavior in transfer lines to ensure the application of ambient condition data for demonstration plant scale-up.

4. Slurry Feed Pump Manifold Study: Determine the effect of fluid viscosity, particle size, and flow conditions on performance and distribution in the slurry feed pump manifold, to ascertain the effectiveness of the design.

The overall plan was to obtain data at conditions reflecting start-up as well as normal operation, and to perform experiments in two-phase gas/liquid systems as required to make certain that experimental techniques were adequate for three-phase study.

TASK 1: HEAT EXCHANGER STUDY

A two-phase solid/liquid flow system will exist in the SRC-I plant's slurry heat exchangers. There are 14 heat exchangers in series for each preheater. The flow behavior, heat transfer, and distribution will depend on whether the heat exchanger is downflow or upflow. Of particular concern for the heat exchanger design is the possibility of the tubes plugging as a result of maldistribution. If a substantial portion of the exchanger tubes are not in service, the efficiency of the heat exchanger will be significantly reduced. Furthermore, the reduction of the number of exchanger tubes in service as a result of maldistribution will cause higher flow rates in the remaining tubes, which also reduces heat exchanger efficiency. Task 1 was designed to determine slurry distributions and flow rates from each experimental exchanger tube. A device was designed to sample the flow so that concentration and slurry flow rates could be measured.

Of equal importance is the behavior of solids at the headers. Whether the solids will settle in the header depends upon the design configuration, flow rates, particle density, and dimension. The header design may also significantly influence the slurry distribution in the exchanger tubes. Experiments were planned to visually observe solid behavior in the heater.

In addition, experiments were planned to simulate the restart of the heat exchanger after temporary plant shutdown. Resuspension of solids that have settled because of temporary shutdown will depend upon the flow direction and design of the headers. In a downflow mode, the restart of a heat exchanger will involve pushing settled solids from the bottom header into the exit piping, which is smaller in diameter than the header. This reduction in flow cross-sectional area may plug the exit line. Thus, the solid/liquid two-phase flow behavior during upflow and downflow must be determined in order to specify the flow direction for the heat exchanger.

The current dimensions of the shell-and-tube slurry heat exchanger preceding the preheater are as follows:

- Shell dimensions: 362 in. long; 13½-in. inside diameter (i.d.)
- 78 tubes, 1-in. outside diameter (o.d.); 0.754 in. inside diameter (i.d.)
- 1.25-in. pitch of tubes on a triangular grid

Experimental Plan

A 78-tube bundle, 13.5 in. in diameter by 30-ft tall, was built to simulate the heat exchanger. Plexiglas tubes of 0.75 in. i.d. were used to fabricate the exchanger tube bundles. Figure 1 schematically diagrams the complete experimental apparatus, which has an overall height of 43 ft.

Methocel solution would be used to simulate the non-Newtonian rheological behavior of the process solvent, with viscosities varied up to 90 cP. Coal, sand, and magnetite were proposed for use as the solid phase.

Slurry distribution was to be determined by flow-rate measurements from each exchanger tube. The heat exchanger was designed to allow slurry flow in either an upward or downward direction, by opening and closing appropriate valves. A centrifugal pump was specially purchased to drive the liquid/slurry flow in the system. This pump has a rubber-lined impeller and casing to minimize the breakup of particles in the system. The flow rate can be varied mechanically in this pump from 250 to 400 gpm.

A special Dynasonics flowmeter with two receivers and amplifiers was also purchased, to measure the overall flow rate in the system. The meter was calibrated in a test run, results of which are illustrated in Figure 2. The test was conducted in a 3/4-in. i.d. tube; the receivers were strapped onto the tube from the outside. The calibration was done for two slurry concentrations of 60/80-mesh sand, over a range of flow velocities that covered the range of velocities expected in the transport piping that leads to the heat exchanger.

As seen from the excellent linear fit graphed in the figure, the instrument requires no calibration. The results are also very important because they substantiate the claim that the Dynasonics meter is not affected by slurry concentration. Note that the straight line plotted in Figure 2 will probably not pass through the origin, because the plane

Figure 1
Layout of Equipment for Task I

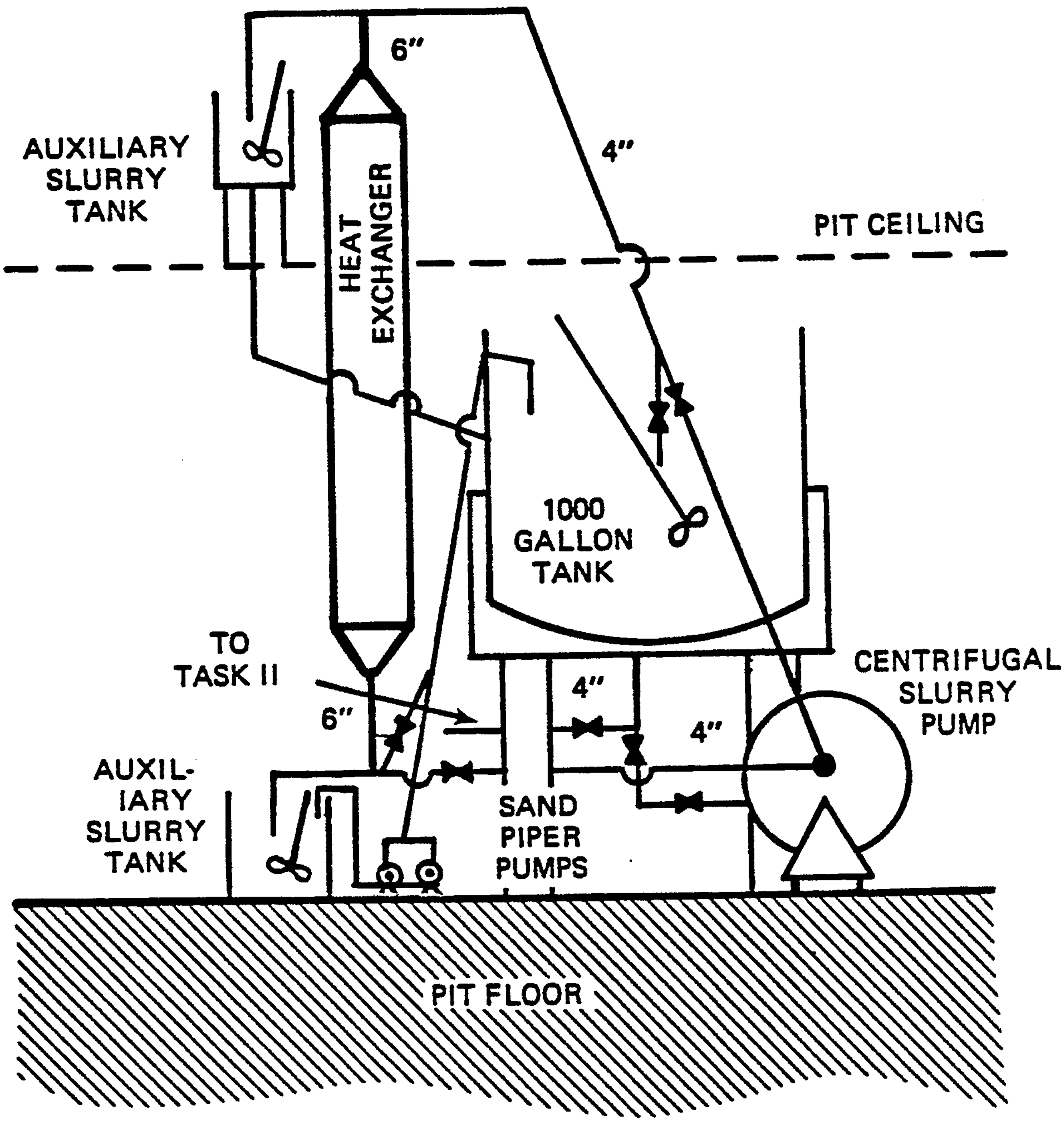
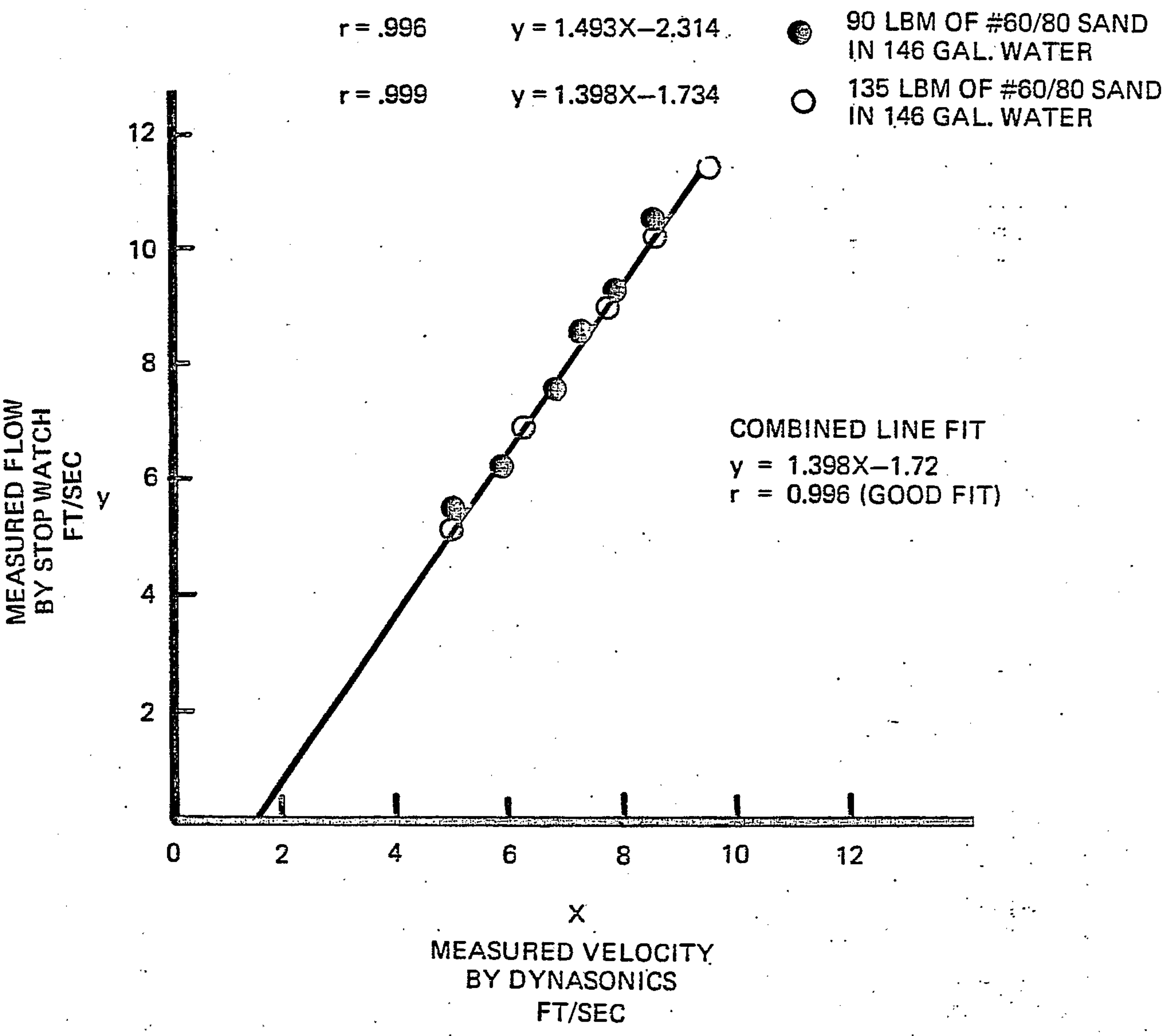


Figure 2
Slurry Concentration at Different Locations



where the two ultrasonic signals meet is basically passing through the axis of the tube. At the axis of the pipe the velocity is expected to be maximum in an upflow situation, whereas the velocity measured directly was calculated by dividing the total flow rate in the tube by the tube's cross-sectional area.

Slurry Concentration Measurement

Figure 3 diagrams the cross-sectional configuration of the tubes in the proposed heat exchanger. In the actual design, a ¼-in. clearance is found around all tubes. Such minimal clearance hinders direct measurement of slurry concentration or flow rates in all but those tubes found at the periphery. Consequently, special systems to measure concentrations and flow rates had to be devised.

Various configurations were designed to compare different ways to sample the slurry. Figures 4-7 illustrate four preliminary attempts to find mechanisms to measure slurry concentration. For each, a set of 24 tubes was selected to represent different locations of the tubes across the heat exchanger grid. This reduction in the number of pipes to be studied was mainly due to rotational symmetry patterns that exist in this particular heat exchanger pattern. The shaded circles in Figure 8 represent the locations of the tubes to be sampled. The idea was to turn the solid shaft labeled A-A in Figures 4-7 90° clockwise, until the entire flow passing through the tube was shut off. The internal workings of the shaft are shown in Figure 5; a system of "O" rings is included to prevent leakage of the slurry around the shaft. The design included a milled channel on the side of the plate holding shaft A-A, leading to the periphery of that plate. At the periphery, a valve can be installed that can be opened to withdraw the slurry sample. This was an attractive system because the flow was only cut off from a few tubes at one time; hence, pump operation would not be endangered.

Unfortunately, when this mechanism was tested on a bench-scale model it was not found to perform well for a prolonged period. With time, the shaft began to grip its housing, because the very fine slurry particles became trapped between the housing clearance. When this happened, moving the shaft to a fully closed position was very difficult.

Figure 3
Cross Section of Tubes in Heat Exchanger

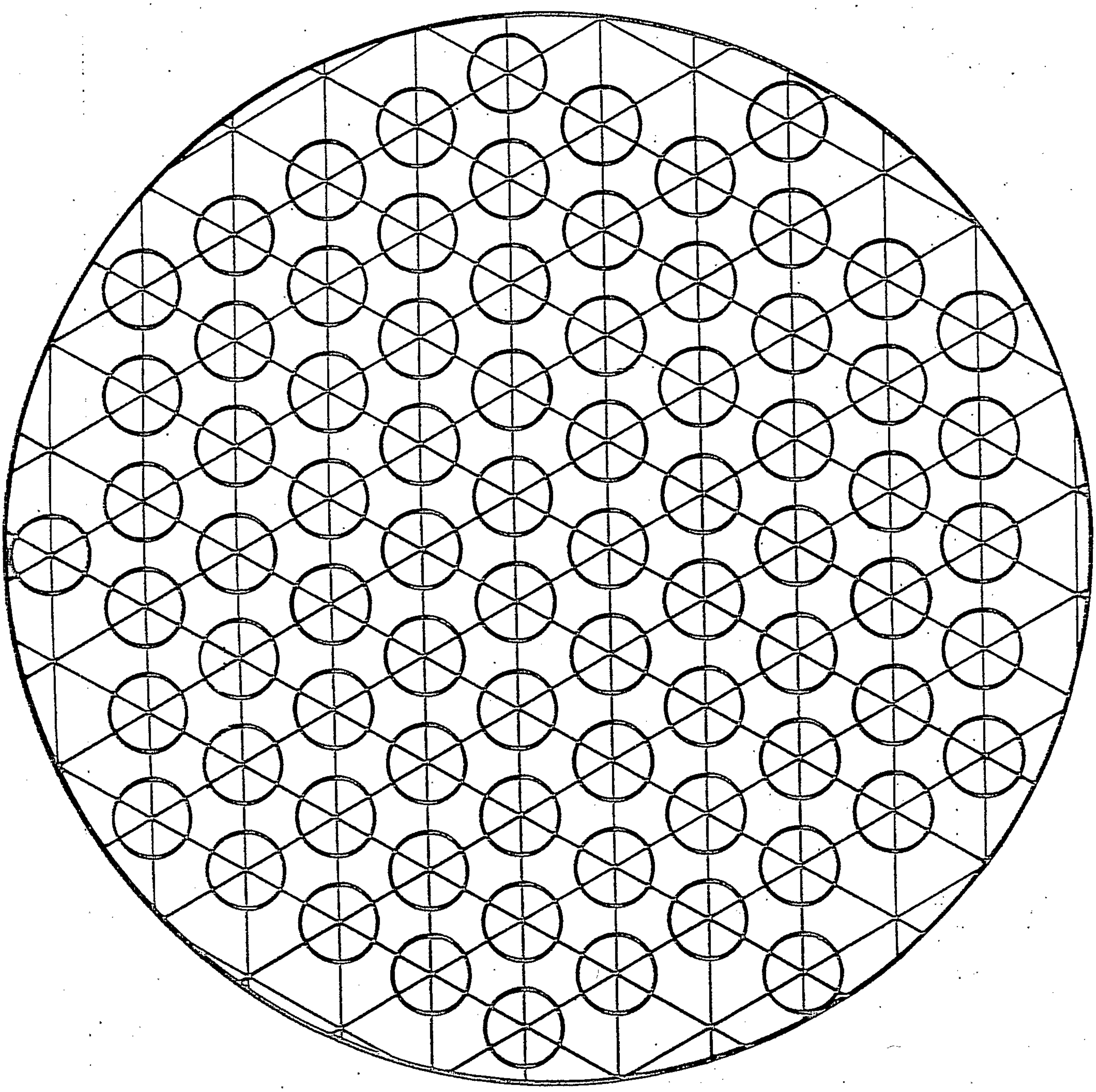


Figure 4
Schematic Diagram Suggesting a
Flow Shut-off Mechanism for Each Tube

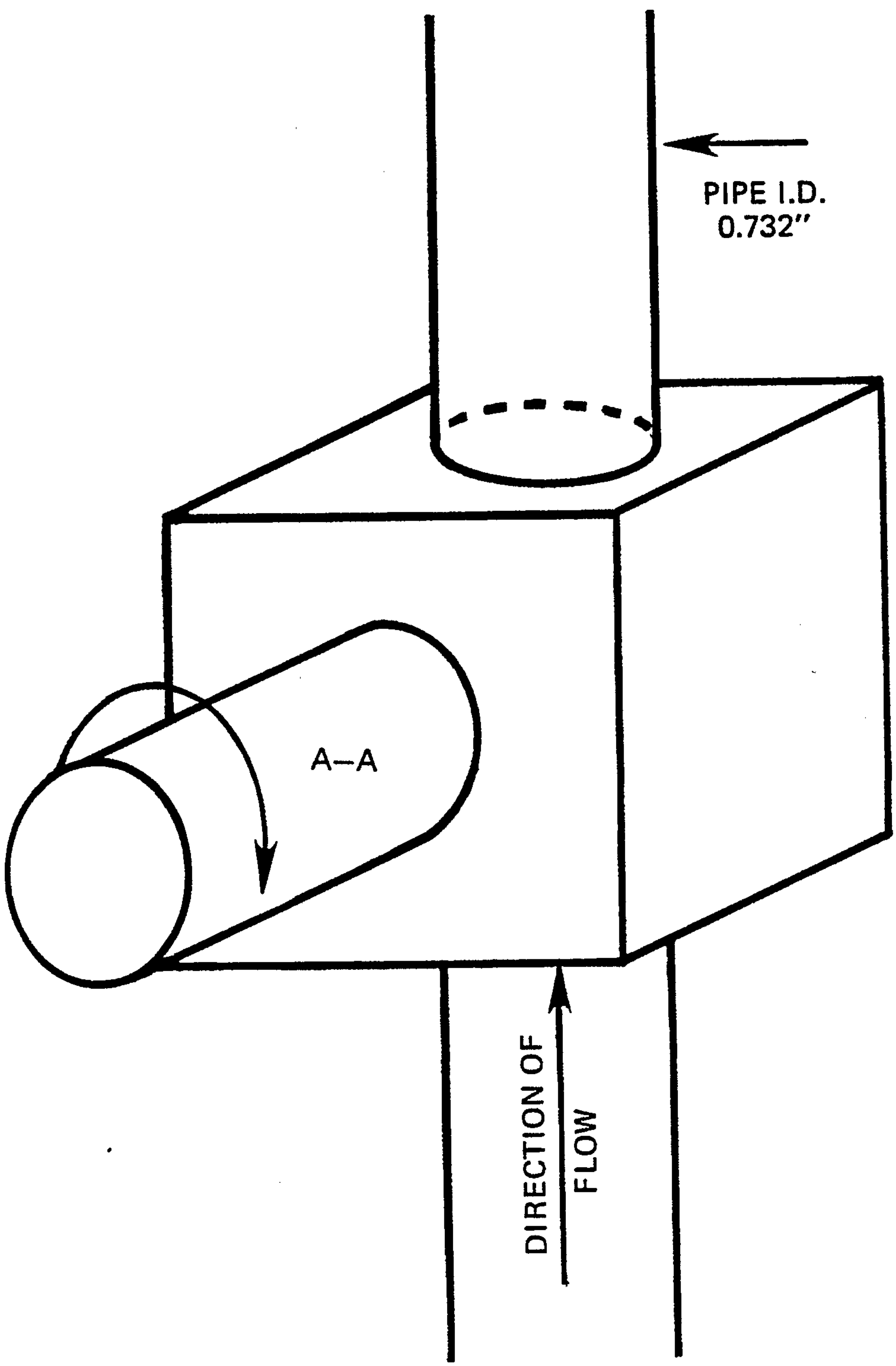


Figure 5
Schematic Diagram Showing Details of Suggested Sealing Arrangements

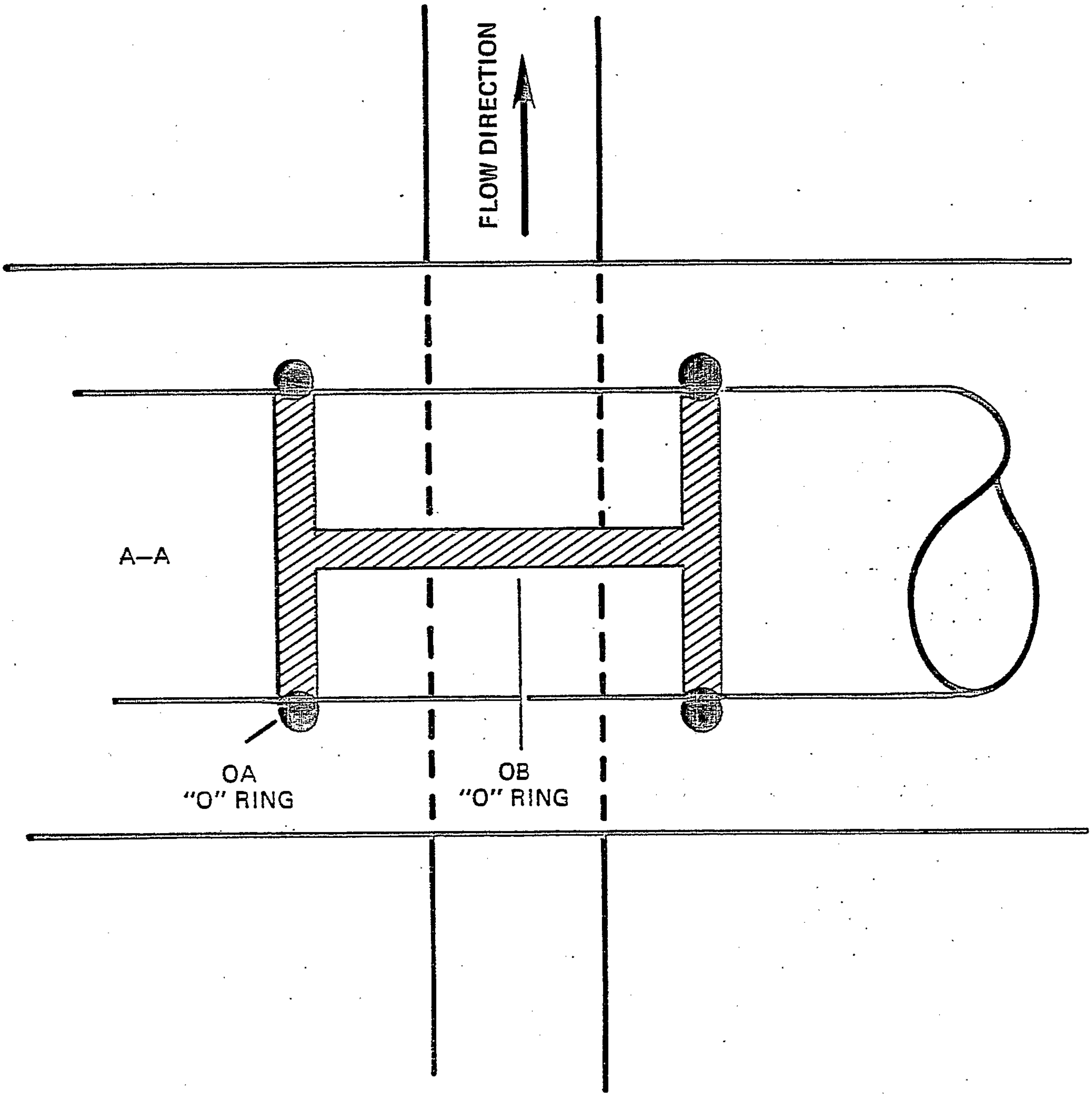


Figure 6
Cross Section of Shaft of Shutoff Mechanism

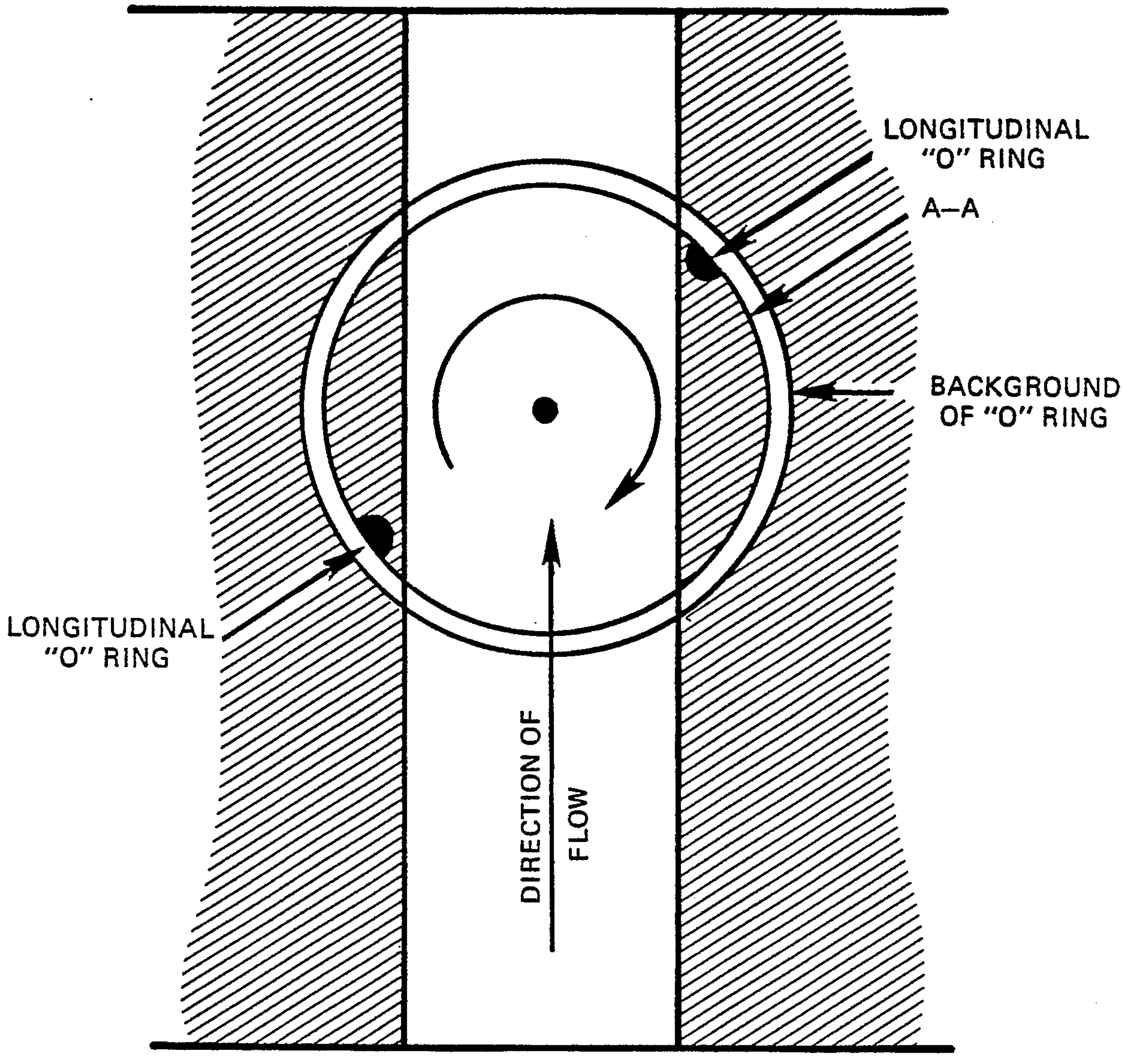


Figure 7
Drain Groove Suggested for Each of the Sampled Pipes

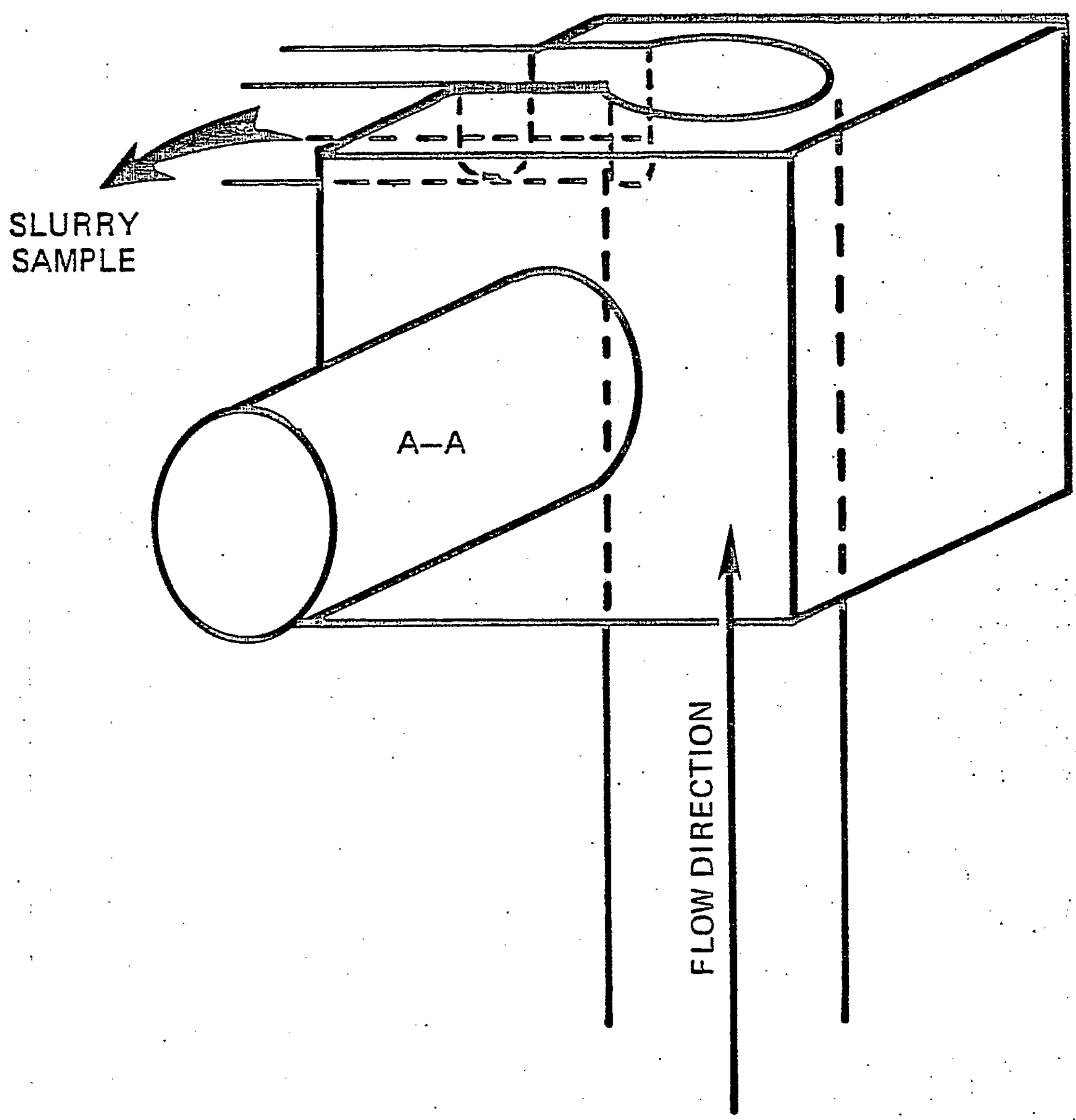
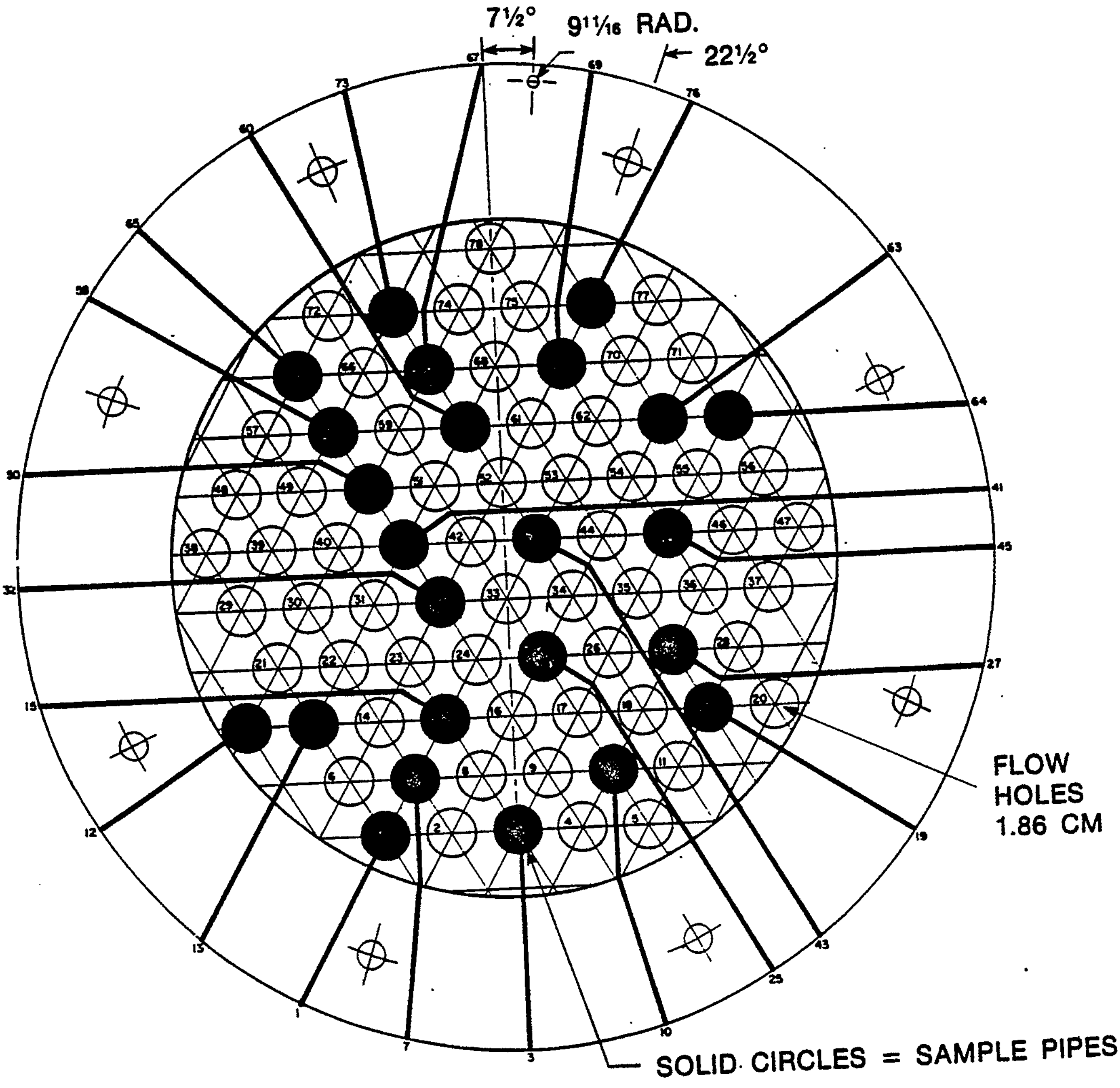


Figure 8
Detailed Cross Section of a Heat Exchanger



Consequently, this sampling technique was abandoned, although the basic idea of having a milled channel on the side of the tube to allow continuous sampling seemed feasible enough to retain.

Thus, the next test run used one 24-ft-tall tube of the same dimensions found in the heat exchanger. Sampling was conducted from two different positions on the side walls of the tube (see Figure 9). The objective was to take samples from the side walls of the tube and compare the slurry concentrations calculated from these samples with the slurry concentration measured in the early tests when the valves were shut off. The latter concentrations are more accurate than those obtained by sampling from the side wall. Thus, comparison with the concentrations measured earlier provides an idea of the accuracy of the side wall sampling technique, which will be used in the actual heat exchanger built for this task.

Figures 10-18 plot the results of calculated concentrations as sampled at various positions along the system. Solids tested were minus 140-mesh and 60/80-mesh sands, at concentrations of 25 lb/ft^3 (379.0 kg/m^3) and 5 lb/ft^3 (72.0 kg/m^3), respectively. These concentrations are close to the expected design concentrations in the heat exchanger. Slurry flow velocities were also varied for these different runs, to observe their effect on the accuracy of the proposed method.

As shown along the abscissae in Figures 10-18, many locations were sampled, including the top and bottom of the mixing tank; the upper half of the pipe (12-24 ft or 3.65-7.31 m); the lower half of the pipe from 0 to 12 ft (0.0-3.65 m); the middle of the lower half of the pipe, by side-wall sampling, i.e., at 6 ft (1.91 m); and, finally, the middle of the upper half of the pipe, i.e., at 18 ft (5.64 m). Also plotted in these figures is a horizontal line representing C_{av}, the average slurry concentration in the system. This is calculated from the weight of solid and the volume of water used at the start of the run.

Figures 10-13 graph results from runs that tested minus 140-mesh sand. Three general conclusions can be derived:

1. The slurry samples tested were homogeneous mixes; no apparent gradients of concentration were noticed in samples taken axially along the pipe. This result was expected, because the

Figure 9
Method of Slurry Concentration Sampling

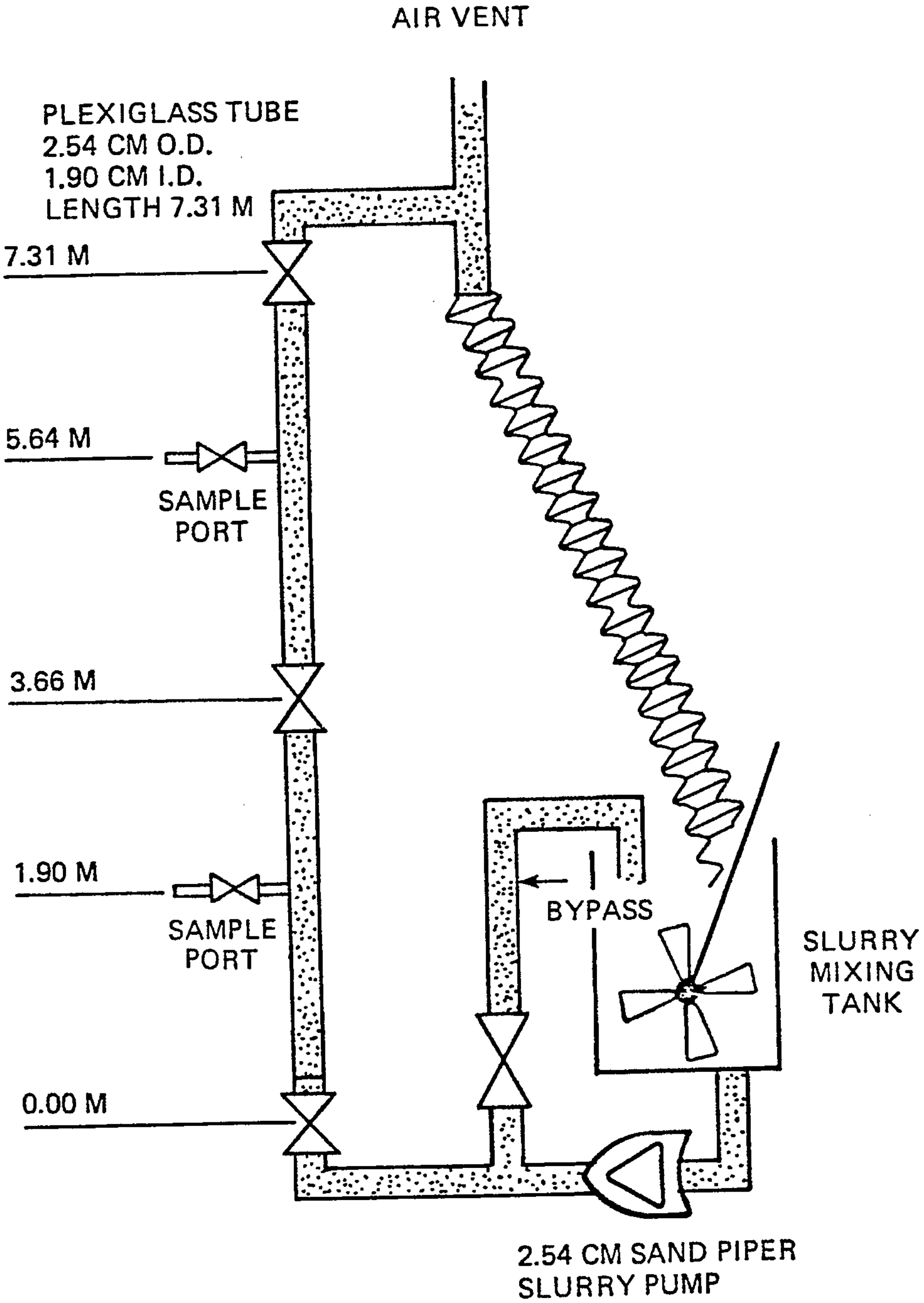


Figure 10
Slurry Concentration at Different Locations
(-140-Mesh Sand; $V_s = 1.30$ m/sec)

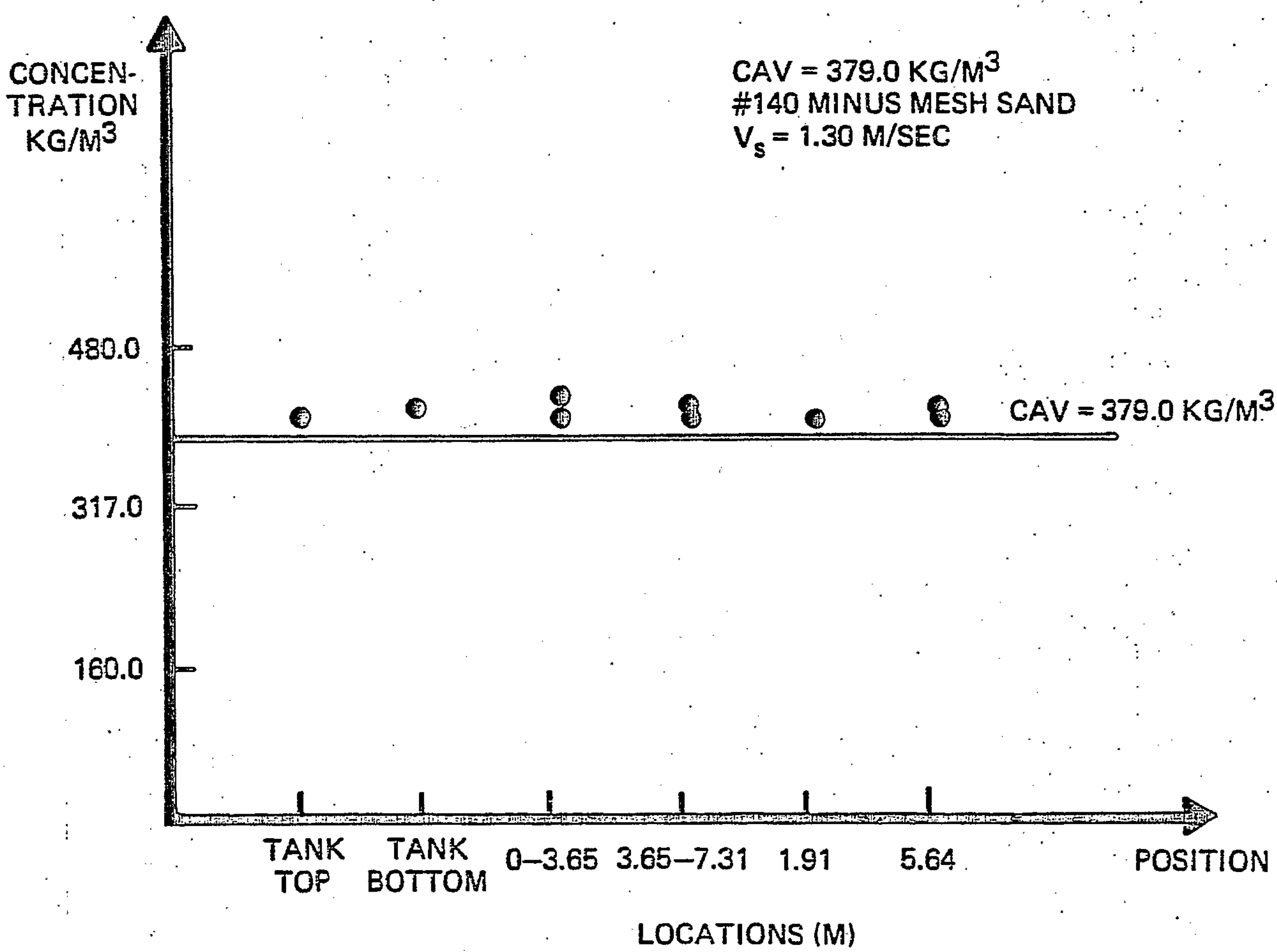


Figure 11
Slurry Concentration at Different Locations
(−140-Mesh Sand; V_s = 1.62 m/sec)

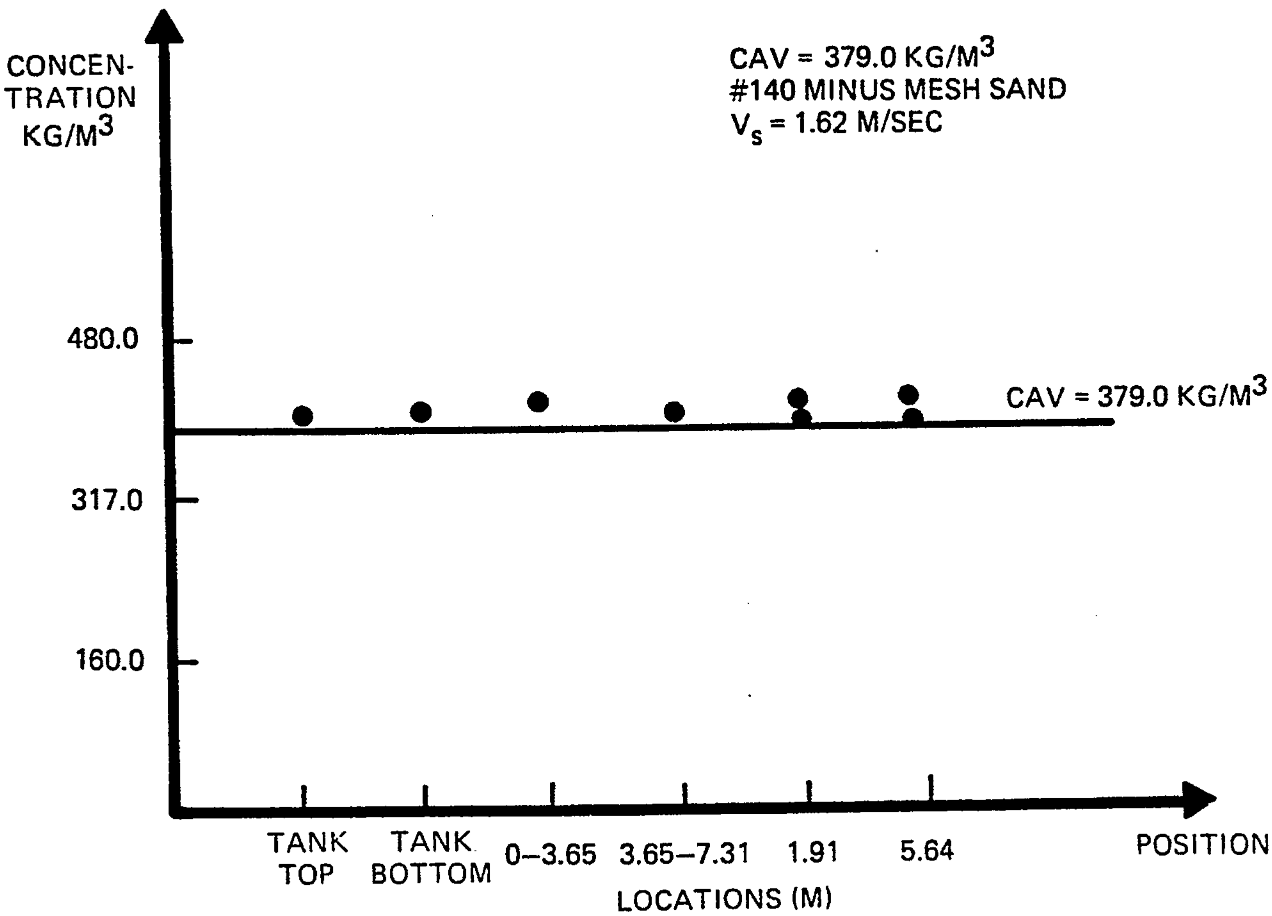


Figure 12
Slurry Concentration at Different Locations
(−140-Mesh Sand; V_s = 1.89 m/sec)

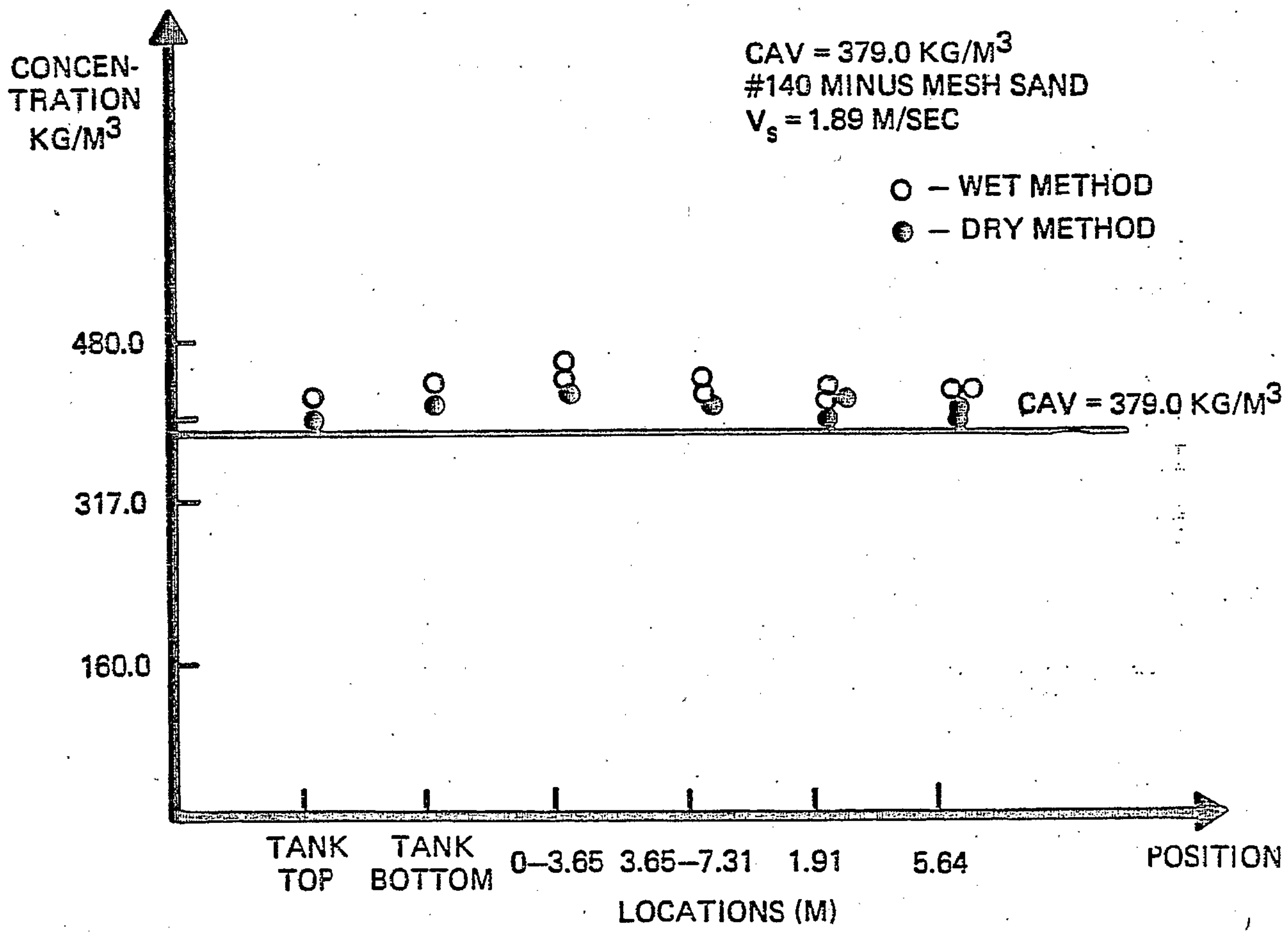
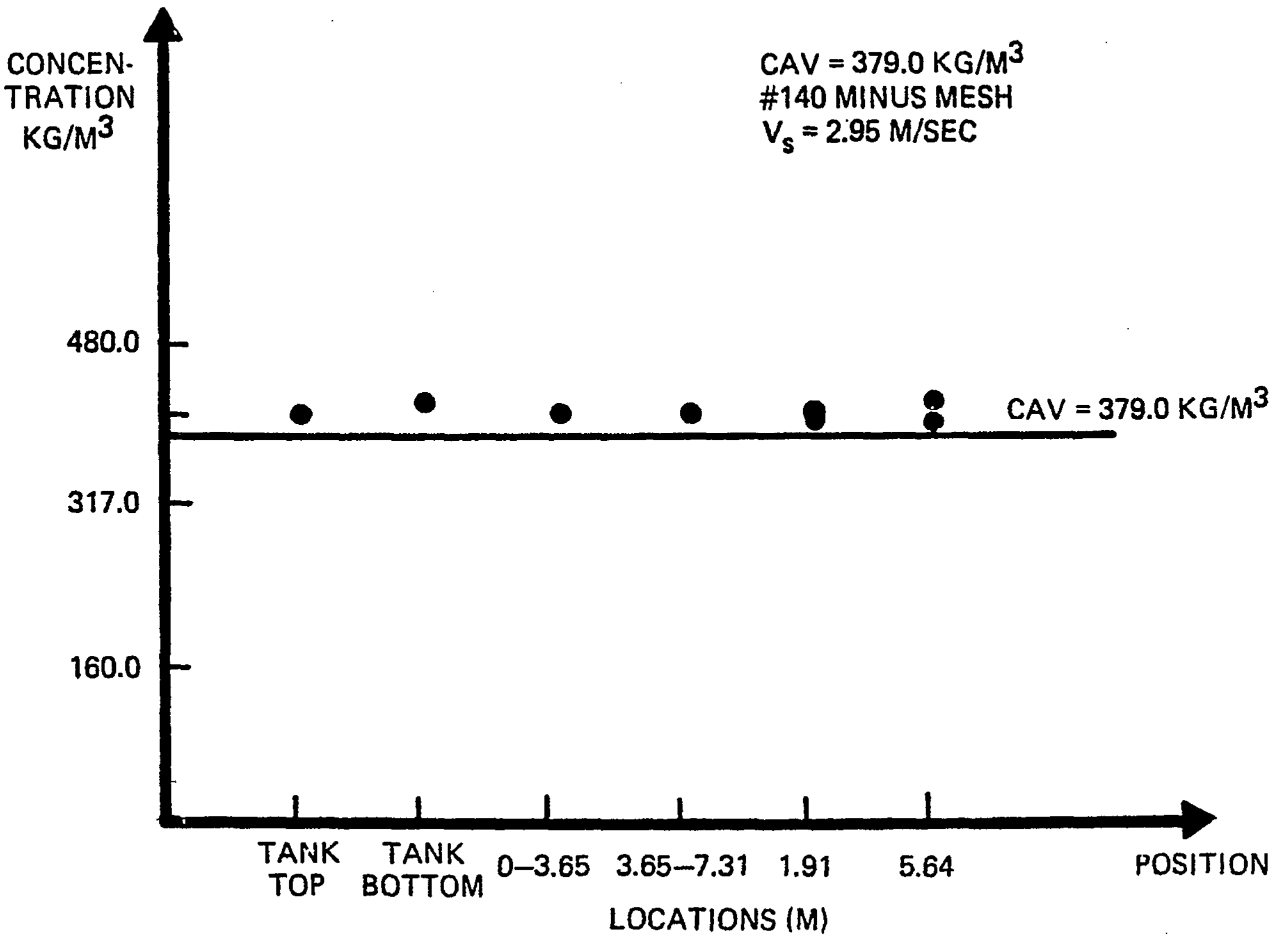


Figure 13
Slurry Concentration at Different Locations
(−140-Mesh Sand; V_s = 2.95 m/sec)



Peclet number (Pe) for this type of flow is expected to be very large, i.e., very little solids axial dispersion.

2. Slurry concentrations calculated for samples collected using the side-wall technique were very similar to actual slurry concentrations in the tube at these locations. This result shows the validity of using such a technique to obtain a reliable quantitative measure of the solids concentration in such tubes.

3. The effect of varying average slurry velocity on slurry concentration was negligible, as measured by this method.

Figures 14-16 show that the results of using this sampling technique on 60/80-mesh sand are similar:

1. There is still a homogeneous mixture axially along the pipe, even with a much larger particle size, as reflected by the average concentrations for the sections from 0 to 3.65 m and from 3.65 to 7.41 m.

2. The calculated slurry concentrations obtained from the side-wall technique are lower than the actual measured concentration, by about 15-20%. Although this seems like a large relative difference, it must be kept in mind that the average concentration is five times smaller than that of the minus 140-mesh runs. The repeatability of the method is also shown by Figures 14-16; repeated measurements have been made at the same location with very similar results.

During actual operation, the concentration of the large particles will probably be much smaller than that of the fines tested here. This will improve the overall accuracy of the method even further.

One can assume that, on the average, the flow of minus 140-mesh sand at a concentration of 25 lb/ft^3 will resemble that of 60/80-mesh sand at a concentration of 5 lb/ft^3. Based on this assumption and a

Figure 14
Slurry Concentration at Different Locations
(60/80-Mesh Sand; V_s = 1.12 m/sec)

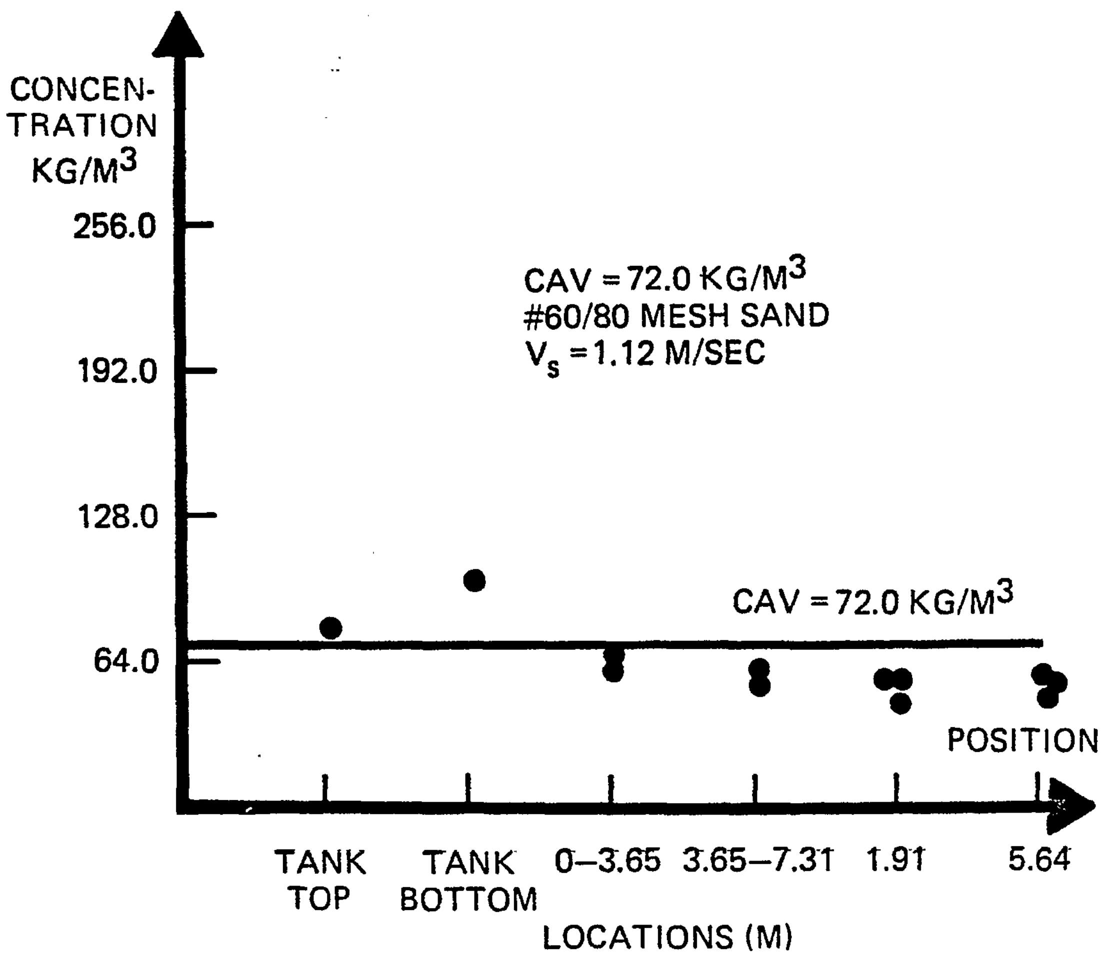


Figure 15
Slurry Concentration at Different Locations
(60/80-Mesh Sand; V_s = 1.72 m/sec)

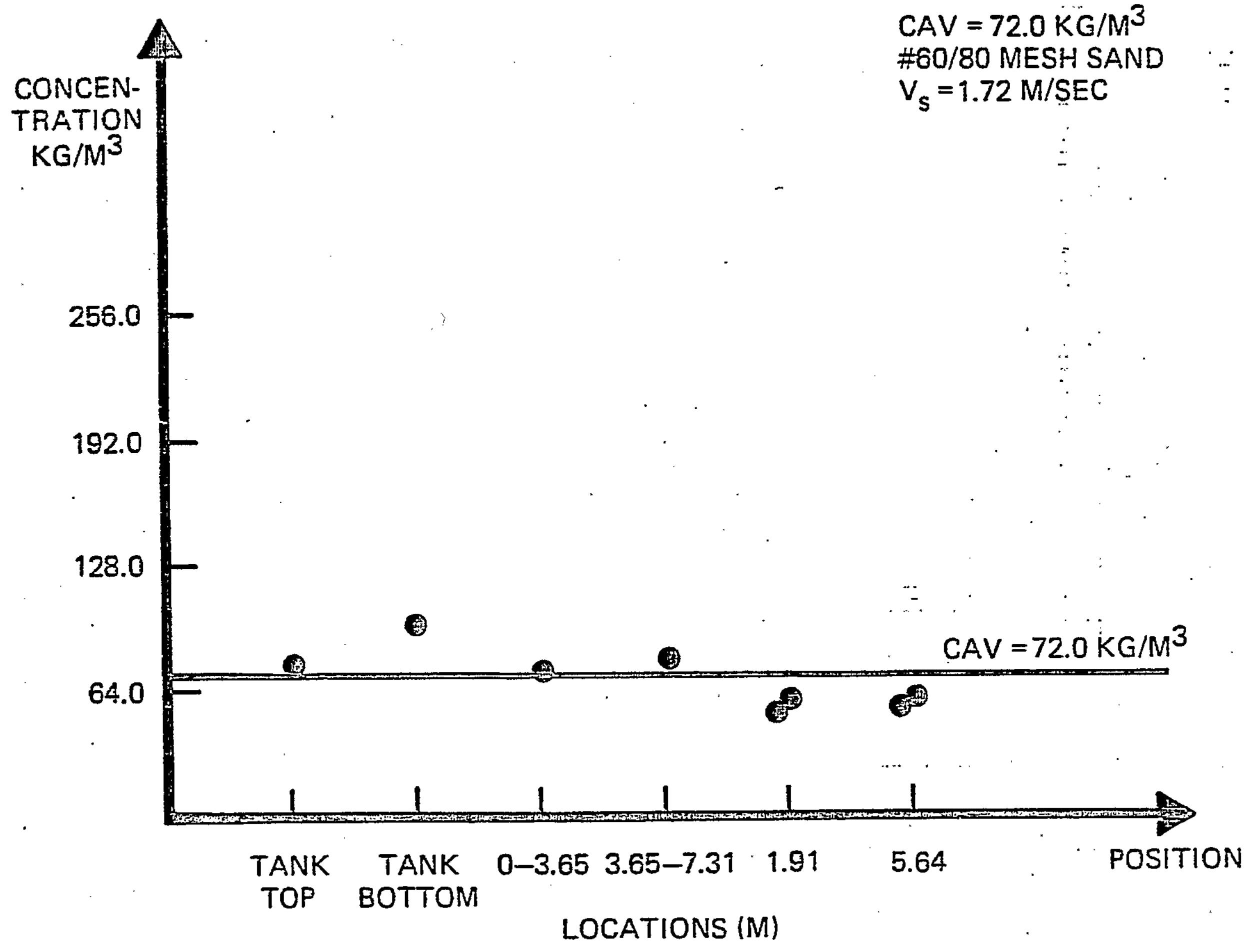
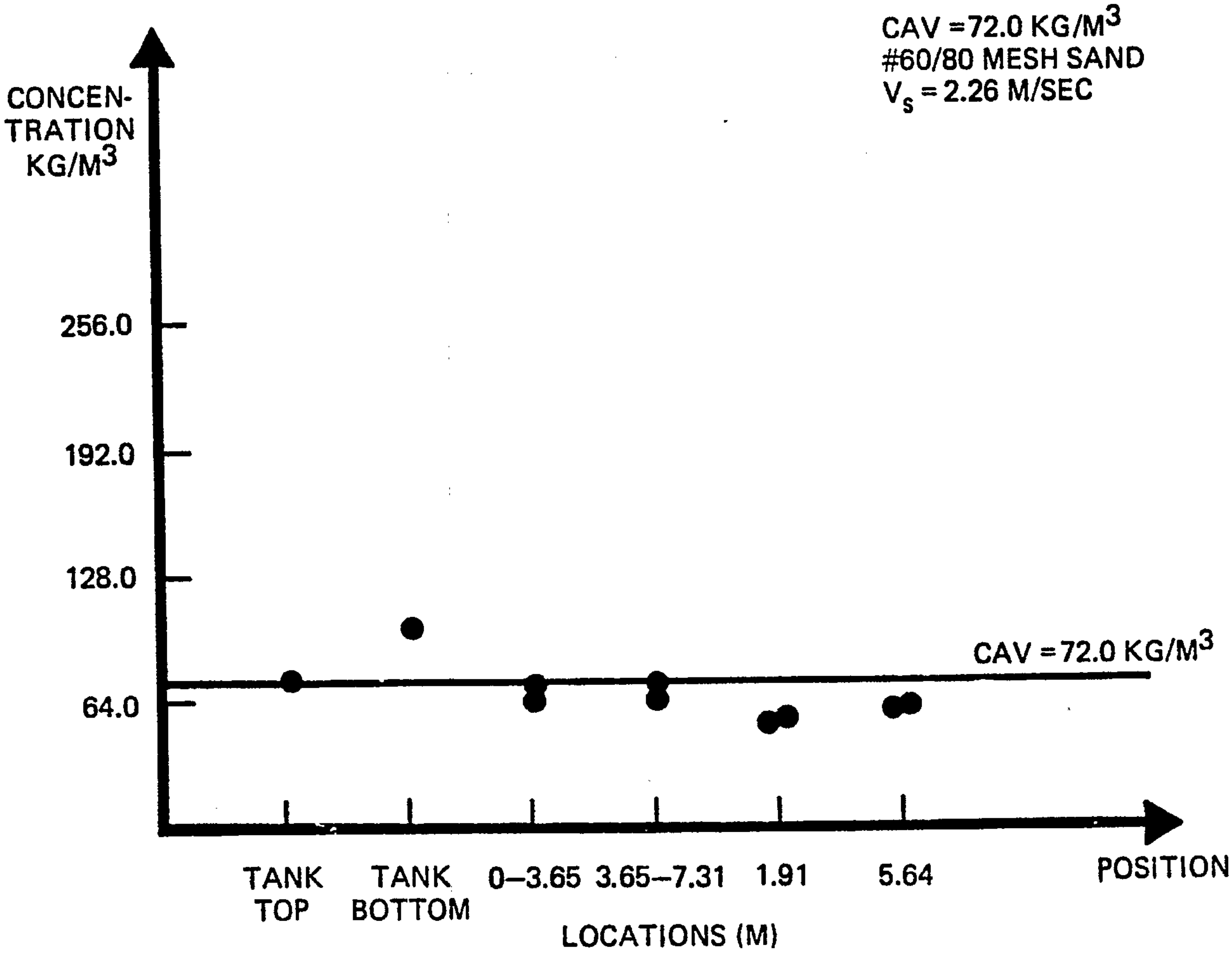


Figure 16
Slurry Concentration at Different Locations
(60/80-Mesh Sand; V_s = 2.26 m/sec)



relative error of measurement for these different-sized sands of 5 and 20%, respectively (140 and 60/80 mesh), the total relative error for the above-mentioned sample can be calculated as follows:

$$\frac{C_{measured}}{C_{actual}} = \frac{25(1.05) + 5(1.2)}{30}$$
$$= 1.075$$

Thus, the relative error from the true concentration in the tube is ±7.5%.

Figures 17 and 18 illustrate results from two additional tests with the 60/80-mesh sand, at the same concentration tested earlier, 5 lb/ft^3, at two different velocities. The only difference in these tests is that the sampling tube that was previously located at the side wall of the flow tube (at 6 and 18 ft) was now pushed to the middle of the flow tube to see if the accuracy of measurement could be improved. As the figures show, accuracy deteriorated tremendously. This could be because the slurry was avoiding the sampling tube, since the diameter of the sampling tube was large compared to the diameter of the plexiglas tube sampled. Rearrangement of the direction of flow because of the size of the sampling tube could have created an area of low concentration near the sampling port itself.

Based on these tests, the layout shown in Figure 8 was selected as the configuration for the 24 tubes to be sampled to measure slurry concentration.

Slurry Flow Rate Measurements

Performance of Task 1 also required a method to measure the slurry flow rate. Because of the proximity of the pipes to each other and the abrasiveness of the slurry, no direct means of measurement could be employed. However, an interesting indirect method that could be used to measure the flow rates was devised, using an optical electronic detection system (see Figure 19).

The piping setup assembled for the concentration tests was also used here. The optical electronic system was added at two points (Stations A and B), located a known distance apart along the tube. Each station consisted of a miniature incandescent light on one side, powered

Figure 17
Slurry Concentration at Different Locations
(60/80-Mesh Sand; V_s = 1.70 m/sec;
Sampling Tube in Center of Plexiglas Tube)

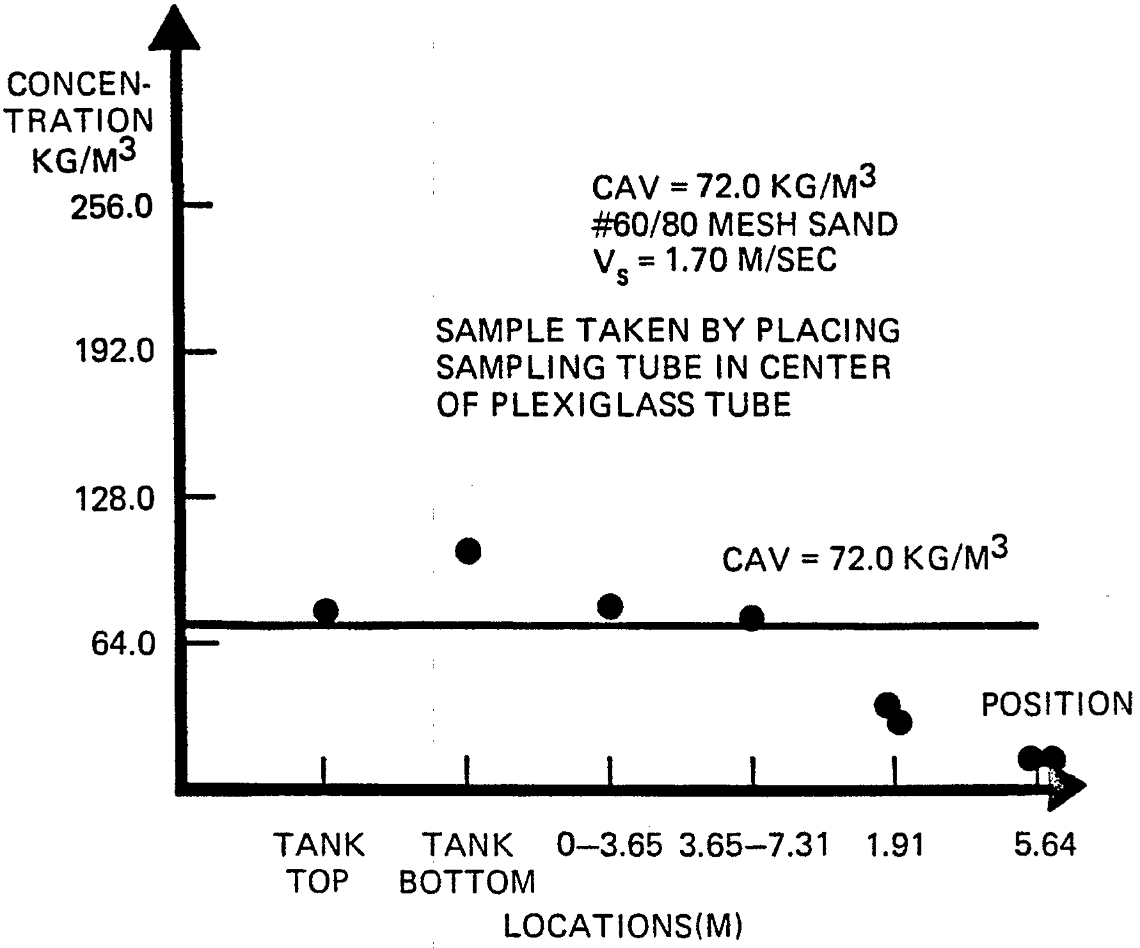


Figure 18
Slurry Concentration at Different Locations
(60/80-Mesh Sand; V_s = 2.28 m/sec;
Sampling Tube in Center of Plexiglas Tube)

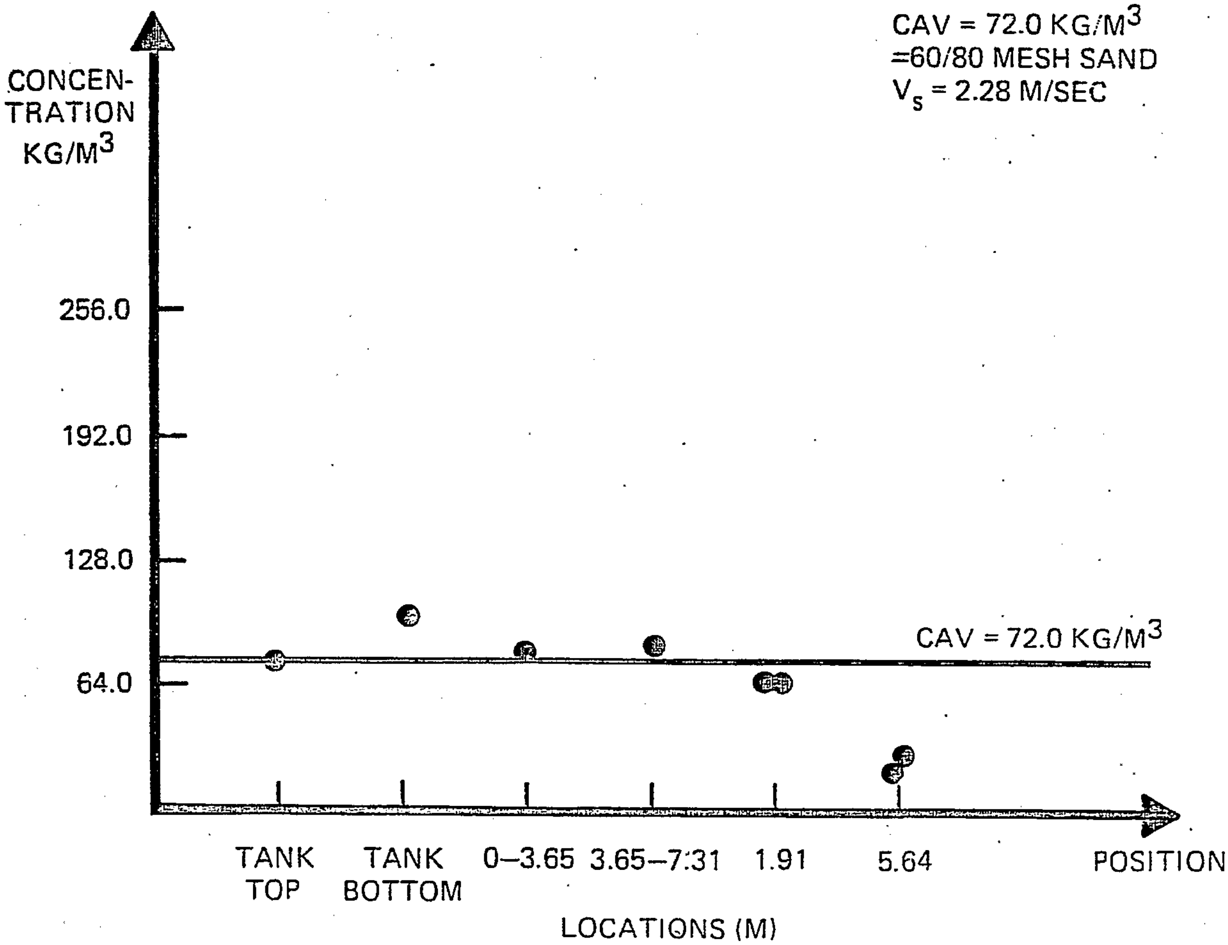
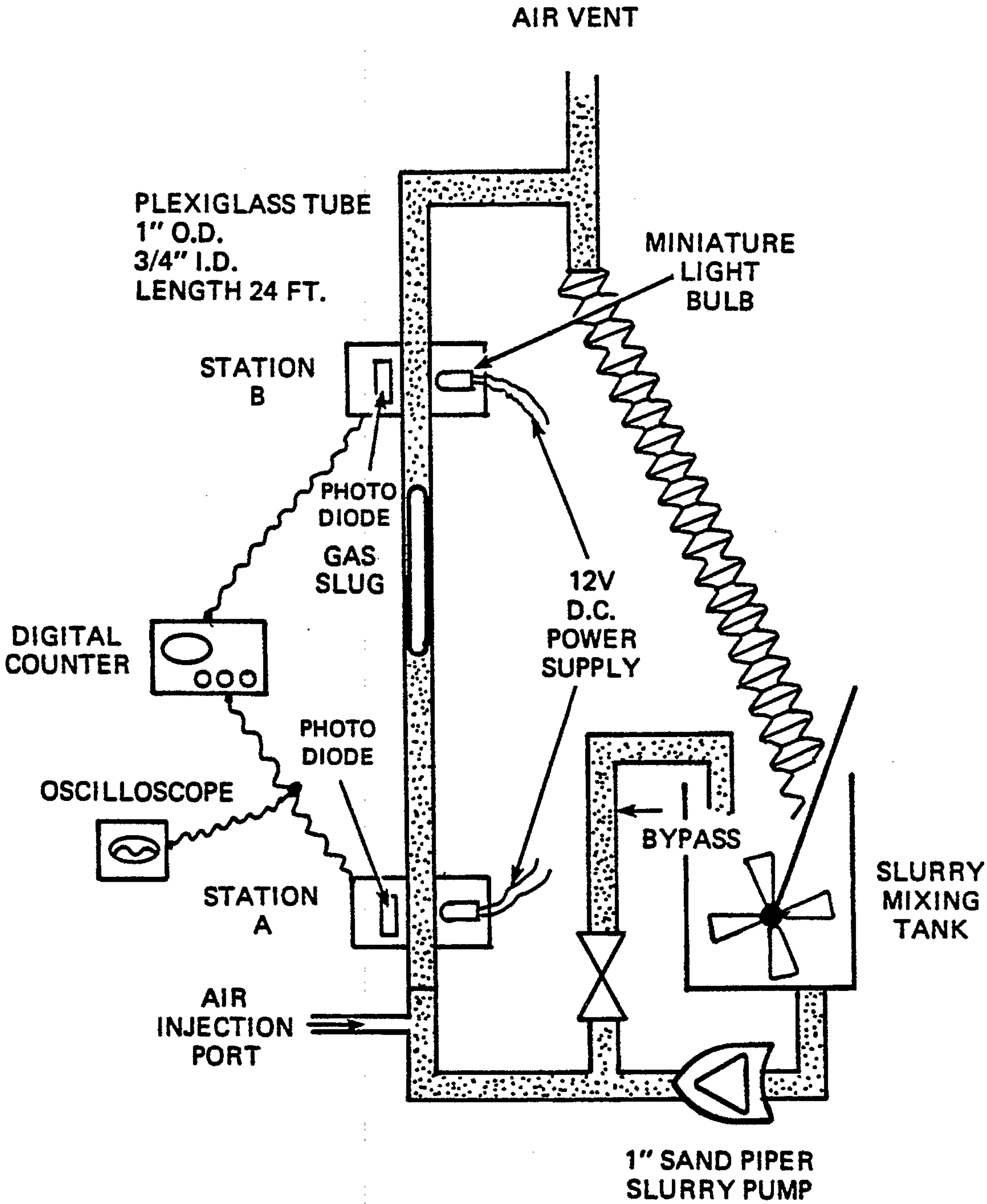


Figure 19
Diagram for Optical Electronics



by a 12-V d.c. battery, plus a light-sensitive photoelectric cell installed on the opposite side of the tube's cross-section. The photoelectric cell responded rapidly to light intensity changes and generated signals proportional to the changes in light that it received, when placed in the proper electronic circuit.

The system operated as follows: Through an air injection port, a known volume of air was introduced. Due to buoyancy effects and the upward flow of the slurry, this air slug would flow upward. Generally, the velocity of the air bubbles stabilizes after traveling a few pipe diameters up in the tube. This effect takes time because the forces of interfacial drag (i.e., between bubble and liquid) and the buoyancy force need time to balance out.

As the tip of the slug passed the first station (a), it interfered with the intensity of light received by the photodiode. This, in turn, triggered a voltage signal, which activated a digital counter that then started a clock. As the tip of the gas slug passed station B, it triggered the voltage signal from that photodiode, again stopping the clock.

Then, knowing the distance from Station A to B and the time it takes the slug to travel that distance, one can calculate the average slug velocity as follows:

$$V_{av} = L/t$$

where V_{av} = average slug velocity (ft/sec)

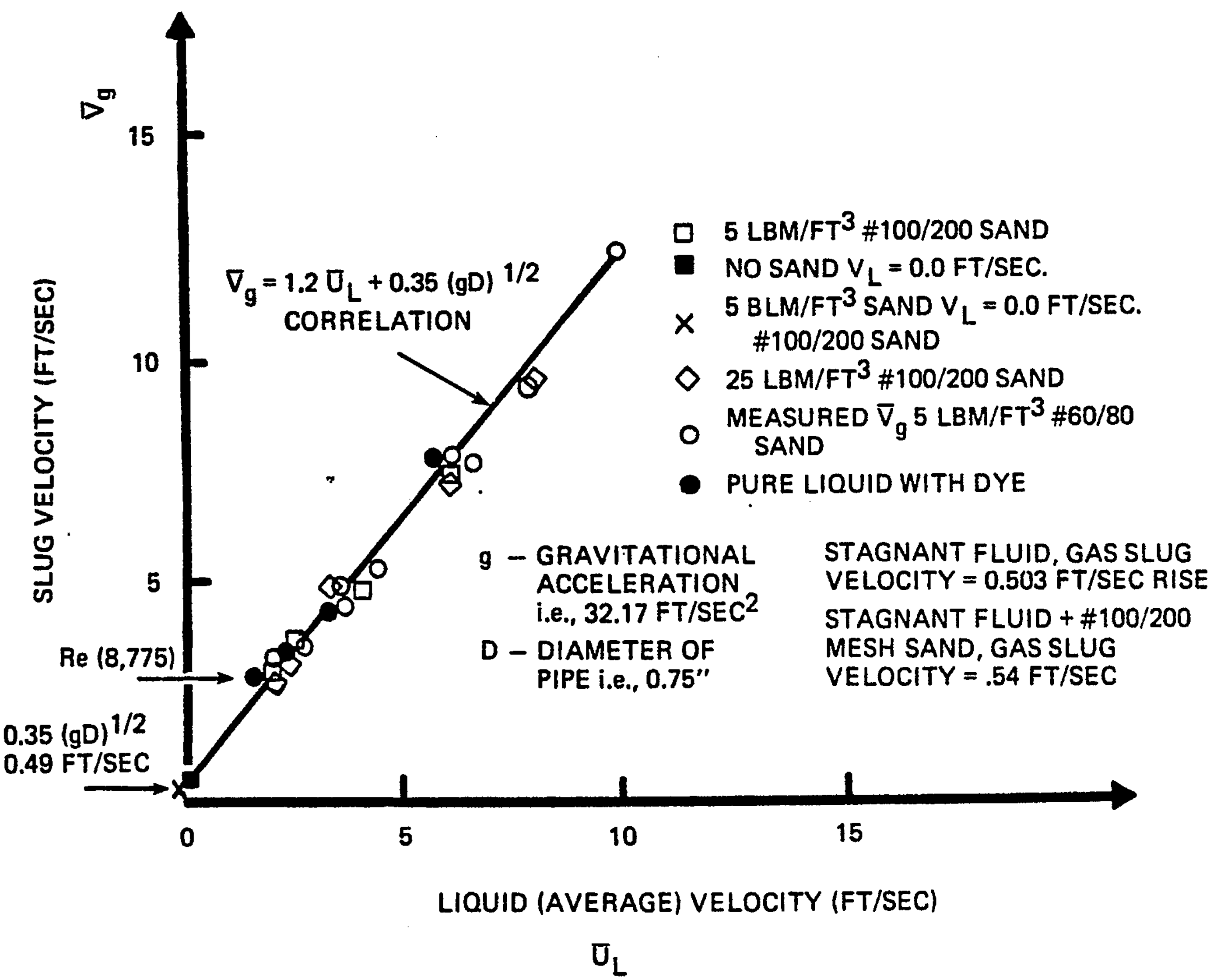
L = distance from A to B (ft)

t = time it takes slug to reach B from A (sec)

In actual operation, the air slug expanded slightly, due to a decrease in static pressure as the slug rose in the tube. However, this 1- to 2-in. expansion is small, considering the overall length from A to B is 6 ft.

The system was tested with several slurry concentrations and various particle sizes as well as a range of velocities to cover those expected during normal operation of the SRC-I plant. Figure 20 plots different slug velocities vs. corresponding liquid velocities, for 5 and 25 lb/ft^3 of minus 140-mesh sand and 5 lb/ft^3 of the 60/80-mesh sand.

Figure 20
Slug Velocity vs. Liquid Velocity



These results were correlated linearly with respect to the average slurry velocity and the inside diameter of the tube. A simplified linear relationship is seen to be arrived at, which compares well with that of Nicklin et al. (1962), as follows:

$$\bar{V}_g = 1.2\bar{U}_c + 0.35(gD)^{\frac{1}{2}}$$

where $\bar{V}_g$ = average gas slug velocity (ft/sec)

$\bar{U}_c$ = average slurry velocity (ft/sec)

g = gravitational acceleration (32.17 ft/sec^2)

D = diameter of tube (ft)

The constants 1.2 and 0.35 were fitted to the data to give a very good correlation. Note that most of these flow conditions would give a Reynolds number (Re) greater than 2,300, which is in the turbulent range. Thus, the index 1.2 is very closely related to the general ratio of the maximum to average velocity in a typical turbulent flow situation (i.e., 1.17).

As for the intercept on the gas slug velocity axis, i.e., $\bar{U}_c = 0.0$ (no slurry flow rate), it was confirmed that the rise in velocity for a slug is very close to $0.35(gD)^{\frac{1}{2}}$. This was also found to be the case when using a 5-lb/ft^3 concentration of minus 140-mesh sand in a stagnant slurry flow situation. Thus, this work established the fact that a feasible means of measuring the slurry flow velocity could be obtained by correlating it to the gas slug velocity.

Tests were also conducted in the downflow mode, but these were not as reliable as those for upflow, especially in the case of water. Hence, the idea of using this experimental setup to measure both upward and downward flow rates was abandoned. We suspect that the method could still work with more viscous fluids, which should not result in the severe breakdown of the gas slug that occurred with water.

If the optical electronics system could be used for the downflow case, the layout of the light source and photodiode arrangement with regard to the flow tube would be that drawn in Figure 21. Of course, another location would have to be considered with a similar optical electronics system to pick up the two signals required for the measurement.

Figure 21
Method for Measuring Average Slurry Velocity in a Tube

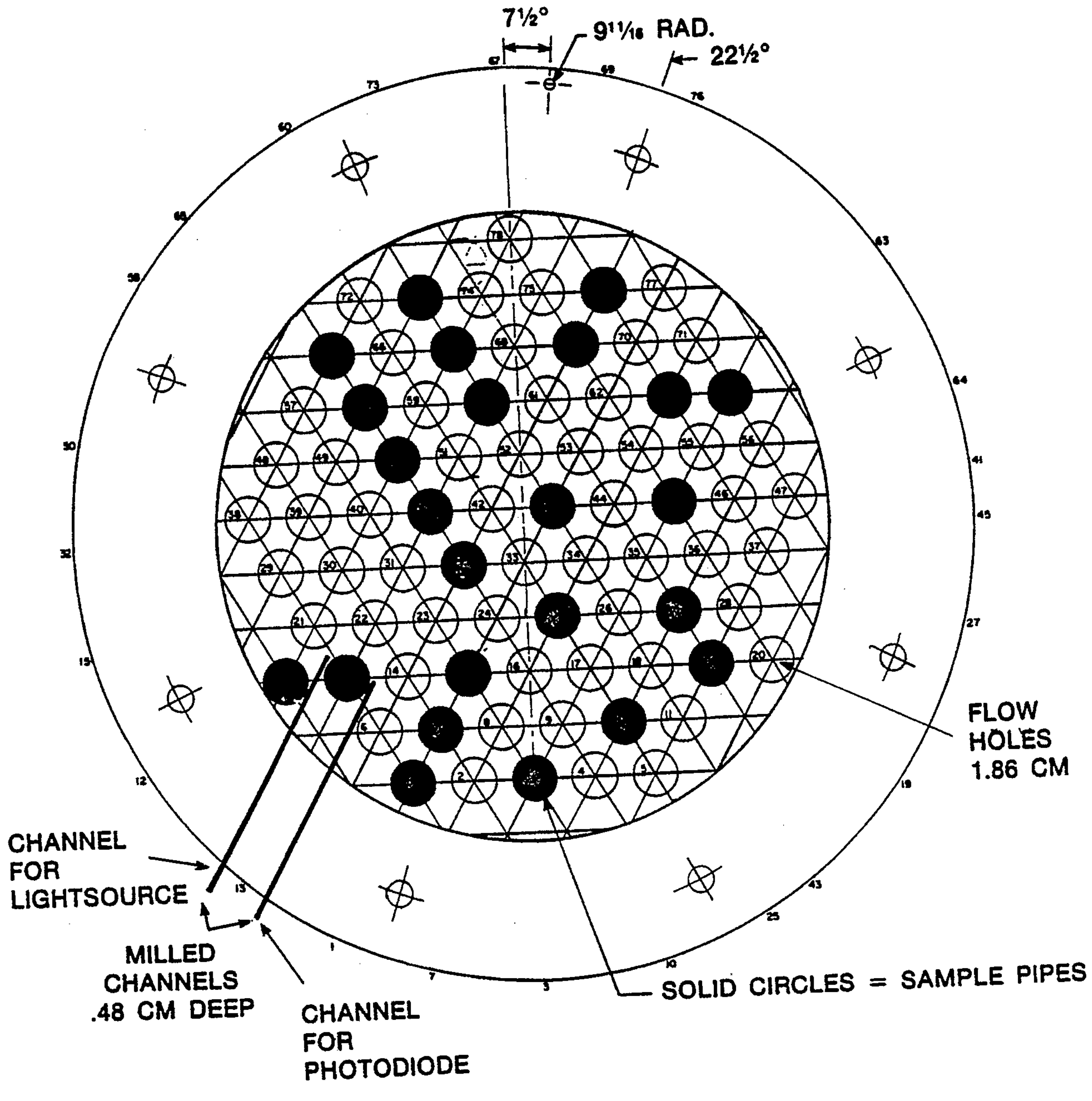


Figure 22 details the heat exchanger design, which includes the relative positions of the slurry concentration and flow rate sampling blocks.

To circumvent the problem of not being able to measure slug velocities in upflow and downflow modes, an alternate method was designed (see Figure 23). This setup consists of a mixing box arrangement, which would replace the reducer at the top or bottom, depending on the direction of flow. The advantage of this box is that it can still mix and manifold the flow coming from the 54 tubes that are not going to be sampled. At the same time, individual flow rates from the 24 sampling tubes can be measured, as well as their concentration. This configuration offers a compromise as far as the geometry of the heat exchanger is concerned, because it is felt that the existence of the reducers in general will help redistribute the slurry flow and concentrations in the heat exchanger. We expect that the relatively low pressure drop due to friction in the heat exchanger compared to its gravitational pressure drop will not significantly change the results.

Termination of the Task

Unfortunately, due to budget constraints, the funding for this project was terminated at the request of DOE and ICRC. At that time, the preliminary studies for Task 1 had resulted in a method that could be used to sample slurry concentration and flow rate. Construction of the experimental apparatus was 100% complete, and about 70% of the planned layout had been assembled.

Figure 22
Schematic of the Heat Exchanger Showing Different Sampling Ports

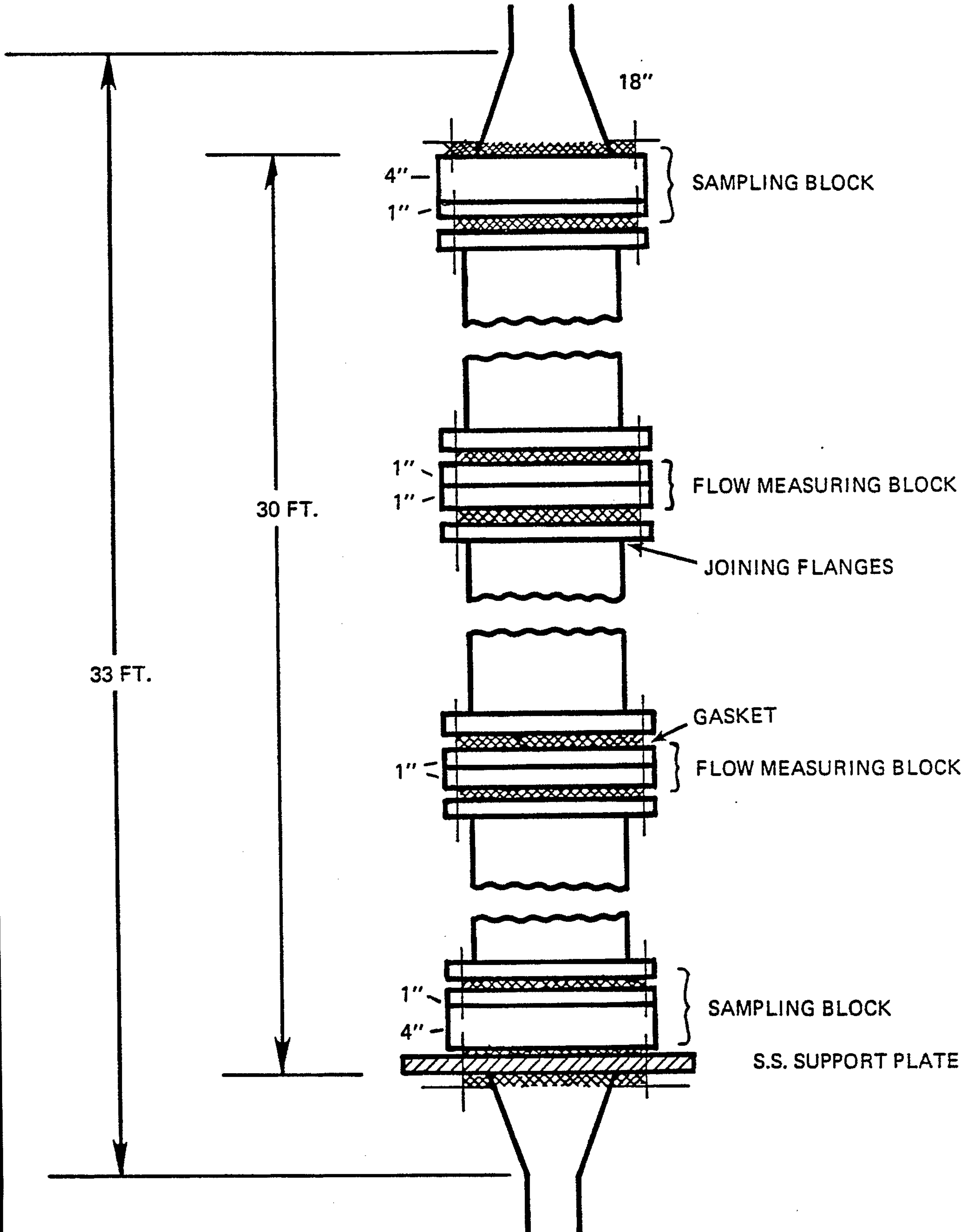
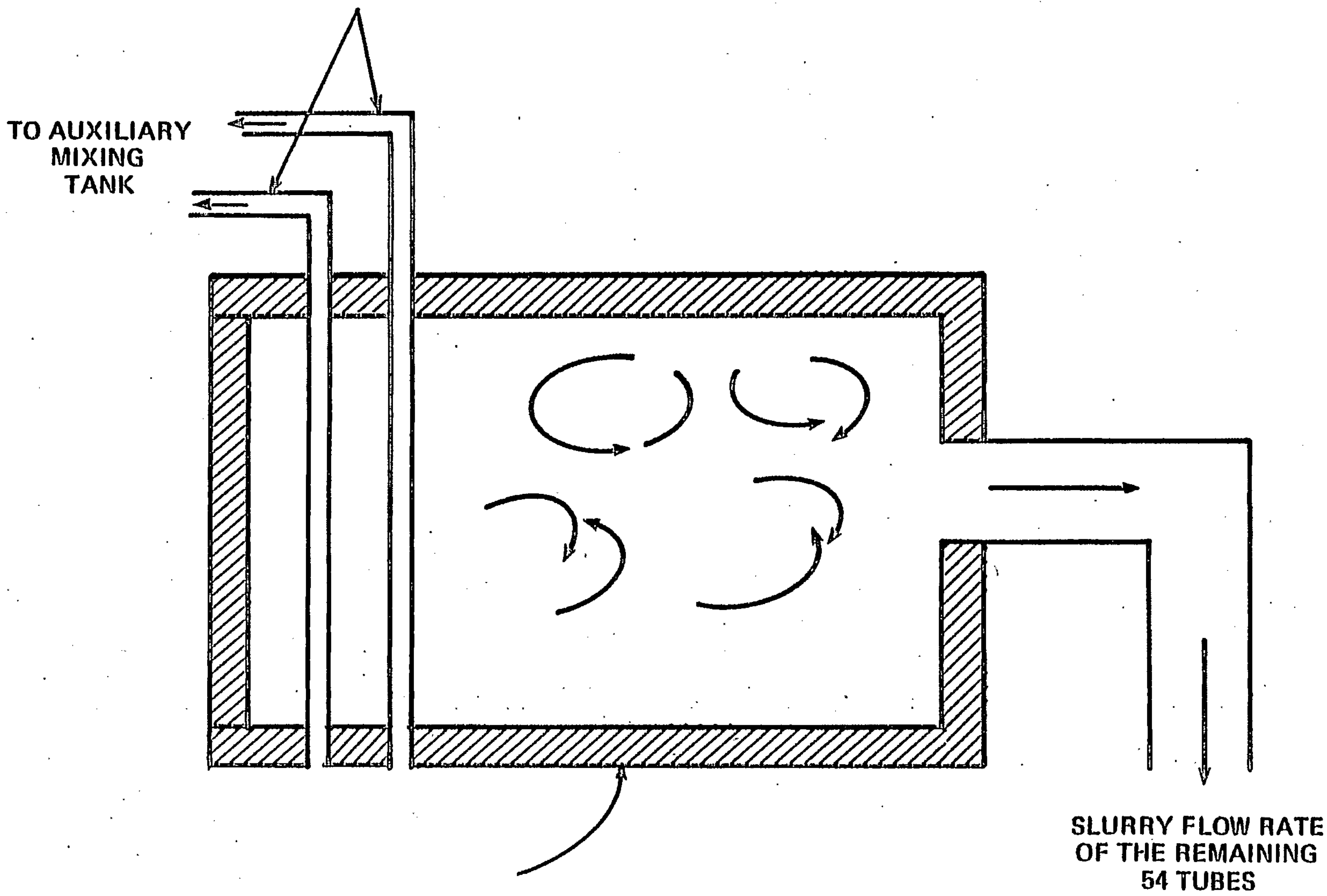


Figure 23
Mixing Box Diagram



TASK 2: TRANSFER-LINE STUDY

Transfer of the hydrogen gas/solvent/coal mixture from the SRC-I Demonstration Plant preheater to the dissolver is another example of a process step whose design depends on flow pattern. Transfer from the preheater to the dissolver will be accomplished through four 8-in. nominal diameter pipes, which will be combined in a manifold before entering the dissolver. Two design configurations are being considered. The manifold could be located at an elevated position, from which it would descend through a single 2-ft-diameter pipe entering into the dissolver. Or, the manifold could be placed at the same level as the inlet to the dissolver. Modeling three-phase flow behavior in this geometry will be extremely helpful in identifying the best design configuration.

Thus, the specific objective of this test was to study the hydrodynamics and flow patterns of gas/liquid and gas/slurry mixtures in the dissolver inlet manifold, the transport lines between the dissolver, and the line that leads from the second dissolver to the separation units downstream.

Experimental Plan

Pressure pulsation is a major concern in transfer lines such as the fired heater pipe. The alternate gas and slurry slugs flowing into the dissolver inlet manifold are a prime source of pressure pulsuation, which will affect flow both downstream and upstream of the dissolver. Several inlet manifold designs have been suggested by Catalytic, but none of these has addressed the problem of pressure pulsation. The experimental plan for part 1 of this task was to construct a Plexiglas model to simulate the demonstration plant design. The effect of pressure pulsation on transfer-line performance was to be investigated and liquid holdup in these systems was to be measured.

The plan specified use of air and methocel solution or a glycol/water mixture in this two-phase study. A minimum of two liquid viscosities varying between 1 and 10 cP was to be examined. The ranges of liquid and gas flow rates in the transfer lines were: liquid superficial velocity, 0.5-5 ft/sec; gas superficial velocity, 1-20 ft/sec.

Different combinations of gas and liquid flow rates were to be investigated to simulate the possibility of temporary failure of slurry pumps and gasifier. Included in the plan were a minimum of 18 runs for two liquid viscosities, at three different liquid flow rates, each at three different gas velocities.

The experimental plan for the two-phase study has the advantage of providing visual observation of the mixing of the two phases. These observations are important in understanding the inlet manifold and transfer-line performance.

The second part of this task was designed to determine the effect of solid particles on the transfer lines and dissolver inlet manifold performance observed in the gas/liquid two-phase system. Whether solids are settled or suspended will significantly affect the performance of these systems.

The plan specified use of sand and magnetite to simulate the solids leaving the fired heater and the solids in the dissolver effluents. Two particle sizes were to be examined: 150 and 60 mesh. The larger particles are used to simulate any oversized and insoluble solids that pass the screening in the coal preparation plant. The same ranges of flow conditions used in part 1 were to be tested, and solids concentrations were to vary between 10 and 15 wt % of the slurry feed.

In addition, experiments were to be performed to simulate start-up of the transfer lines after temporary shutdown. Particulate attention was to focus on solids suspension, providing a data base for plant operation.

<u>Accomplishments</u>

Figure 24 diagrams the overall layout of the equipment that was designed to be used in common for both this task and the preceding task. Both tasks were to share the common 1,000-gallon feed tank, in which all the slurry is mixed and then fed into respective systems.

Figure 25 details the specific layout for this task. At first, we thought that the centrifugal pump used for Task 1 could be used to circulate the slurry in these experiments. However, further thought indicated that this would not be a good way to achieve the same slurry flow rate in the four legs of the manifold connected to the dissolver.

Figure 24
Layout of Tasks 1 and 2

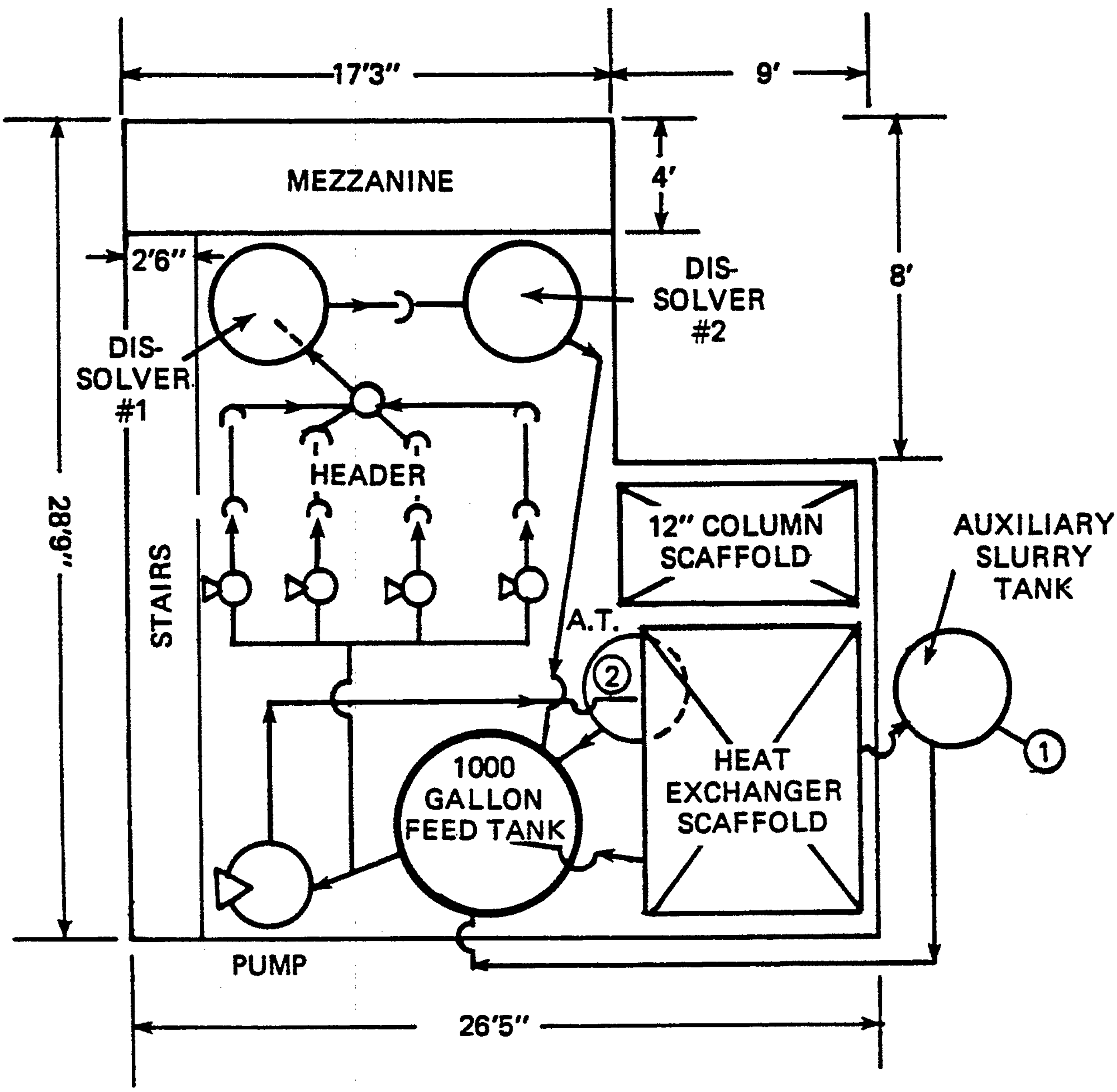
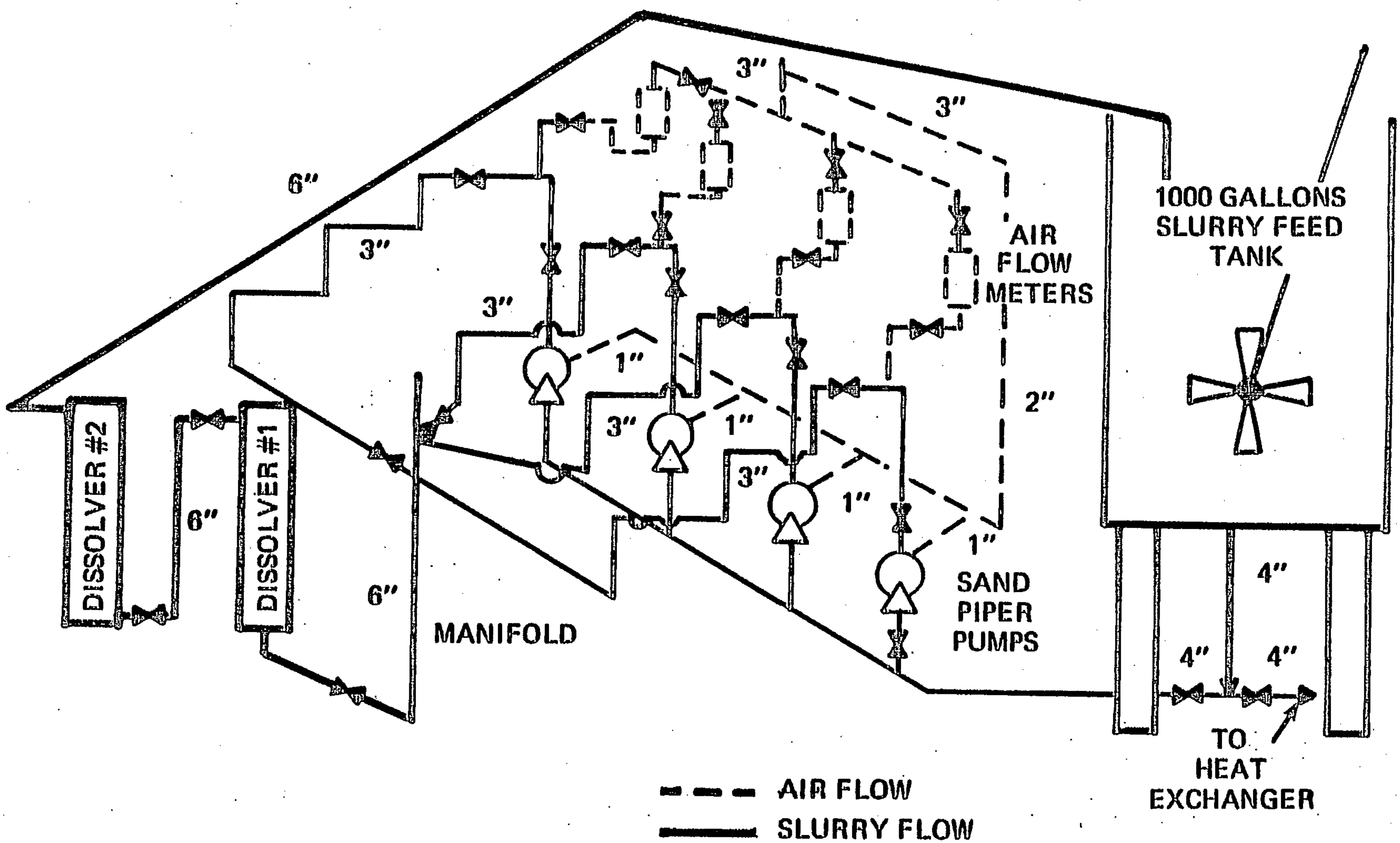


Figure 25
Detailed Piping and Valving Layout of Task 2



In the actual design, four individual pumping systems will pump slurry into the manifold. Thus, the task of redistributing the flow after it enters a manifold tends to be difficult, because each adjustment to an individual leg is bound to change the setting on all others.

Consequently, the need to use a battery of four available 3-in. Sandpiper pumps was evident. These pumps will have their suction side connected to a common manifold, but will discharge separately into different legs of the manifold connected to the dissolver. The flow can then be controlled by means of the air pressure used to activate the pumps, coupled with the use of dual receiver/amplifier Dynasonics flowmeters, which were calibrated as discussed previously.

Figure 25 also shows how the geometry of the different legs was duplicated from the actual design, which requires in some cases more than one bend and elbow. The flow was discharged into the manifold from each of the legs, perpendicular to the manifold.

Many on-off valves were installed in this system in order to measure the gas holdup in some of the legs that transport the three phases. From such measurements along with the average superficial gas and liquid velocities, a correlation can be derived to construct some type of a flow regime map for this particular correlation.

Pressure transducers were to be used for dynamic pressure and pressure-drop measurements across the different legs and transport legs to relate that to the flow maps. For task 2, all equipment and materials plus the four air flowmeters were purchased. Also, the air compressor that was supposed to provide the power for the slurry pumps and the required air flow in the experimental apparatus was brought on site. Hence, the construction for that task was about 50% complete.

Unfortunately, termination of funding prevented any further construction or completion of the experimental plan.

TASK 3: EFFECT OF PRESSURE ON TWO- AND THREE-PHASE FLOW IN TRANSFER LINES

Another area of the design requiring modeling of three-phase flow behavior in a vertical downflow system is the interstage transfer line between the two dissolvers in series. The pressure drop will depend on both the geometry of the transfer line and the flow conditions. A bottom return bend immediately upstream of the second dissolver will cause appreciable pressure surge, which could conceivably affect the flow pattern both upstream and downstream. The degree of mixing, as well as the mass transfer in the transfer line, will depend on whether coring-bubble flow or froth flow occurs in this line. Likewise, other three-phase transport lines, such as the line from the second dissolver to the dissolver effluent separator, behave in a similar manner. All such transfer-line designs will benefit from a general understanding of the hydrodynamics in these systems.

The objective of this task was to determine the effect of pressure on the results derived from the cold-flow simulation study, in order to justify its application to the coal liquefaction process. The most obvious change was the increase of gas density with increasing system pressure. In this task, the configuration of one of the transfer lines from the preheater to the dissolver under Task 2 was to be constructed with material suitable for moderate-pressure operation (90 psig) to simulate the gas density in the actual process. Plans called for design and construction of a closed system employing Freon/water components, and examination of two gas densities ranging from 1 to 3 lb/ft^3. The ranges of gas and liquid flow rates were identical to those chosen for Task 2. A total of 36 runs were to be performed.

Unfortunately, because of the funding cutbacks, work on this task was never started.

TASK 4: EFFECTIVENESS OF DESIGN OF SLURRY FEED PUMP MANIFOLD

Parts of the coal liquefaction plant will involve two-phase solid/liquid slurry, without gas flow. Typical examples are in the coal/solvent preparation area and in the deashing unit lines. After coal and process solvent are thoroughly mixed in large holding vessels, the coal/solvent mixture will feed into a manifold, which in turn will feed into four slurry pumps. One of these four slurry pumps is a spare unit. The slurry flow distribution depends on the manifold design. Furthermore, without the turbulence induced by the gas phase, the likelihood of solid sedimentation will undoubtedly increase in upflow, and even more so in horizontal flow. The situation will become more serious at turndown velocities. The major concern is solid sedimentation on the suction side of the pump and its effect on pump operation. In addition to pipeline orientation, particle size, fluid viscosity, and flow conditions will influence solid sedimentation. The homogeneity of the mixture and distribution among the feed pipes are also important to plant operation. An understanding of solid/liquid flow behavior in this geometry is required to confirm the slurry feed manifold and piping design.

The experimental plan for this task was to design and construct a scale-down model of the demonstration plant slurry-feed pump manifold. Plexiglas pipes were to be used to provide visibility of solids behavior. Methocel solution was to be used to simulate the process solvent, at viscosities varying from 1 to 40 cP. Coal, sand, and magnetite were to be used at 40 wt % in this study, the sand and magnetite simulating the insoluble particles present in the coal. Two particle sizes (150 and 60 mesh) were to be examined. In addition to visual observation of solids behavior in the manifold, critical slurry velocity, which is defined as the minimum velocity to maintain particles in suspension, was to be determined for all three solids at the two particle sizes described above. Finally, start-up simulation was also to be performed.

Unfortunately, because of the funding cutbacks, work on this task was never started.

LITERATURE CITED

Nicklin, D. J., J. O. Wilkes, and J. F. Davidson. 1962. Two-phase flow in vertical tubes. Trans. Instr. Chem. Eng., Vol. 40.