LARGE-SCALE DISSOLVER COLD-FLOW MODELLING. SRC-I QUARTERLY TECHNICAL REPORT, OCTOBER-DECEMBER 1981. SUPPLEMENT

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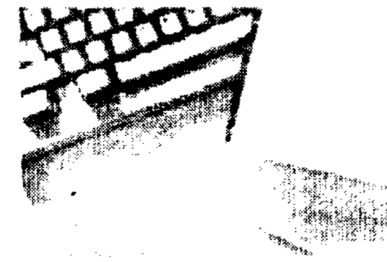

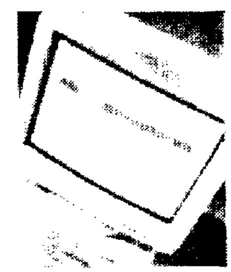

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Allentown, Pennsylvania

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Note

Articles in this Supplement were previously withheld from publication to avoid premature disclosure of inventions. The article on page 1 was included in the January-March 1982 Quarterly Technical Report (DOE/ER/03054-7). Pagination has not been altered, since this could invalidate any cross-references.

LARGE-SCALE DISSOLVER COLD-FLOW MODELLING

W. T. McDermott and D. H. Ying*

Researchers at Allentown Laboratories have been studying the hydrodynamic flow within a large-scale column, 6 ft in diameter by 25 ft tall with a 1-ft-diameter inlet, to simulate two-phase (air/water) and three-phase (air/water/solids) mass transfer and solids removal in the coal dissolver for the SRC-I Demonstration Plant. Evidence from experiments at the Wilsonville Pilot Plant and predictions developed from cold-flow-derived correlations suggest that hydrogen consumption will not be controlled by the rate of gas/liquid mass transfer.

During this reporting period, studies began on gas/liquid contact during batch and continuous liquid flow after all project-related equipment had been fabricated and installed. The piping layout design simulates the interstage from the first dissolver to the second dissolver in series. Transparent transfer lines were installed in order to observe the patterns of gas/slurry flow.

ICRC researchers also completed studies of the dissolver design without internal parts. Air was fed into the dissolver vessel through either a 3-in.- or a 12-in.-diameter inlet in order to measure gas holdup, rates of gas/liquid mass transfer, and gas bubble size. Researchers filmed batch and continuous flow in the contact vessel to accurately record the full range of flow conditions during each mode of operation. Results indicate that vessel performance depends upon the size of inlet diameter. For example, gas holdup for the vessel configuration without using internals was less when the large, 12-in-diameter inlet was used, than when the small, 3-in.-diameter inlet was used. Like the coefficient for gas holdup, the coefficient of gas/liquid mass transfer was less with the large inlet than with the small

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*International Coal Refining Co., ICRC.

inlet; however, this difference resulted from the decreased size of the inlet opening but disappeared at high gas velocities.

PROCEDURE FOR EXPERIMENTATION

Preparing Equipment

The previous report specifies equipment needs in detail according to a plan that covers the entire experiment (McDermott and Ying 1981). Researchers used this overall plan as a basis for completing modifications to the equipment and piping layout for the 6-ft-diameter cold-flow simulator.

The piping has been designed so that gas/liquid contact can be studied for both two-phase gas/liquid, and three-phase gas/liquid/solid, systems. Figure 1 shows a schematic of the system constructed for these studies. Air is distributed through a manifold to instrument, operating, and process lines. One of the air process lines simulating the gas feed into the dissolver joins with the water/sand slurry line in a 12-in.-diameter pipe before entering the column. The other air supply line goes directly into the column through a gas sparger system or into the rear of the drain assembly. Nitrogen is sparged through the system to desaturate the water of oxygen, so that the rate of oxygen uptake from air into water can be measured to calculate the rate of gas/liquid mass trnasfer.

The design of the 12-in.-diameter feed pipe simulates the interstage from the first dissolver to the second dissolver in series. Gas combines with the slurry in the elevated horizontal section of this feed pipe, which then bends 90 degrees, and drops to a lower level. Part of this vertical section is transparent so that researchers can study the downflow pattern. Then the gas/slurry flows through another horizontal section, before it enters the column at the bottom. Part of this second horizontal section is made of plexiglas so that researchers can observe the pattern of gas/slurry flow, before the slurry enters the column.

Gas/Liquid Contact

Experiments to determine the extent of gas/liquid contacting are being performed with and without any internals in the column, as described in the previous report (McDermott and Ying 1981). Each set of experiments using a particular design configuration is followed by a period for modifying equipment for use in the next experiment. To cover the full range of dissolver operating velocities, researchers are investigating superficial gas velocities of up to 0.5 ft/sec and liquid velocities of up to 0.05 ft/sec.

Measuring Gas Holdup

Researchers used liquid displacement to measure the difference in liquid levels with and without air bubbling through the liquid. The ratio of this difference in volume to the aerated liquid volume represents the overall gas hold-up. Liquid level measurements were made with a level-sighting tube external to the column and with a calibrated scale on the internal wall of the column. In an alternative measuring technique, the delta-h technique, the difference in height between the contents of the aerated vessel and the level of the unaerated liquid in the sighting tube is used to calculate the effective density of the vessel contents, which is then used to calculate the gas holdup.

Gas holdup was measured without using any internals and by using batch operation. The graphs of gas holdup, measured by the two techniques described above, agreed excellently (Figure 2). In addition, air was fed into the vessel either through the 12-in.- or through the 3-in.-diameter inlet pipe located at the rear of the drain assembly, as shown in Figure 1. The results presented in Figure 2 indicate that gas holdup is reduced when air is introduced through the 12-in. line. This difference is due to relatively large gas slugs that form when the gas is fed through the large, 12-in. inlet rather than the 3-in. The large slugs rise rapidly through the contact vessel and thereby reduce the average gas-phase residence time and holdup. To compare these results with others, Figure 2 also shows the line representing the correlation developed by Akita and Yoshida (1973).

As Figure 2 indicates, the prediction of Akita and Yoshida is higher than the the data for the experiment without internals. This difference is attributed to the clear liquid region generated by the combination of no internals and the dish-shaped bottom of the contact vessel. Figure 3 shows a schematic representation of this clear, unaerated region. Subsequent reports will investigate whether introducing internals in the vessel will successfully prevent the clear region from forming.

Measuring Gas/Liquid Mass Transfer Coefficient

Mass transfer was measured by using probes to detect dissolved oxygen to determine the rate of increase in oxygen saturation in the water during the sparging operation. Nitrogen was first sparged into the column to desaturate the water of oxygen. Simultaneous oxygen readings of three vertical column levels were used to determine the degree of uniformity of oxygen saturation throughout the column. The rate of increase in dissolved oxygen in the water was then used to calculate the mass-transfer coefficient.

Oxygen concentration in the liquid was measured as a function of time at three different levels: 7, 11, and 22 ft below the exit line for liquid (Figure 3). Figure 4 shows a typical plot of oxygen concentration versus time. Concentration profiles measured at the top (7 ft) and middle (11 ft) positions were identical; whereas, the concentration at the bottom (22 ft) lagged behind the others by approximately 10 sec.

Researchers observed that, in the design without vessel internals, the gas entered at the dish-shaped bottom of the contact vessel and rose conically, as shown in Figure 3, which thereby created the clear liquid region around the edge of the column bottom.

Approximately one contact vessel diameter was required for the gas to disperse uniformly across the vessel. Air introduced into the vessel produced, near the vessel bottoms, liquid with practically no bubbles present. The lowest probe, 22 ft below the exit line, was in the region of clear liquid. Therefore, when air flowed into the vessel of oxygen-desaturated water, the bottom probe reported a lag in time between the rates of oxygen uptake in the clear liquid region and in the aerated region monitored by the other two probes. Thus, water in the aerated region of the vessel became oxygenated faster than water in the virtually unaerated region.

Researchers calculated the coefficient for volumetric mass transfer, $k_L a$, by assuming that the liquid phase was completely backmixed.

In general, this assumption is valid for bubble columns that closely resemble backmixed reactors. Values of k_La were calculated by using the following equation:

$$k_La = \frac{1-\varepsilon_G}{\Delta t} \ln \frac{C^* - C_i}{C^* - C_f}$$

where k_La is the coefficient for volumetric liquid-phase mass transfer, (sec^{-1}); Δt is time in seconds; ε_G is gas holdup; C^* is completely saturated oxygen concentration; C_i is initial concentration of dissolved oxygen; and C_f is final concentration of dissolved oxygen.

Figure 5 shows coefficient values obtained with no vessel internals as well as the curve corresponding to the Akita and Yoshida correlation (1973). Akita and Yoshida indicate that the effect of diameter size in their correlation levels off after vessel diameter equals or exceeds 1.97 ft. Consequently, in Figure 5, ICRC researchers calculated the correlation curve for a vessel diameter of 1.97 ft, according to the suggestion of Akita and Yoshida.

For the entire range of gas velocities studied by feeding the gas through both the 3-in.- and 12-in.-diameter inlets, the values predicted by Akita and Yoshida were consistently higher than those measured by ICRC. At gas velocities up to 0.4 ft/sec, the mass transfer coefficient was higher when air was fed through the 3-in.-diameter inlet. This difference may be due to the relatively large gas slugs observed when the 12-in.-diameter inlet was used, because the large slugs reduce the interfacial surface area for gas/liquid transfer. However, at higher gas velocities, this difference diminished, possibly because turbulence is extremely high at high flow rates.

Subsequent reports will discuss volumetric mass-transfer coefficients measured with the various proposed internal configurations.

Data Recorded on Film

When no internals were used inside the vessel, with magnetic tape and audio-visual equipment, researchers filmed batch and continuous flow to record bubble size and dispersion as well as surface behavior. In the contact vessel, many finely dispersed bubbles, approximately 1/8 to 1/4 in. in diameter, remained throughout the full range of flow conditions investigated. Larger bubbles, which frequently had diameters close to that of the inlet, were superimposed on the smaller, dispersed bubbles. The large slugs were not generated when gas was introduced through the 3-in. inlet. In the 12-in.-diameter, vertical downflow section of the feed pipe, gas/liquid mixtures flowed annularly for each liquid and gas flow rate investigated. That is, liquid flowed down the walls of the pipe; gas flowed down the center of the pipe.

LITERATURE CITED

Akita, K., and F. Yoshida. 1973. Gas holdup and volumetric mass transfer coefficient in bubble columns. Ind. Eng. Chem. Process Des. Develop. 12(1):76.

McDermott, W. T. and D. H. Ying. 1981. Large-scale dissolver cold-flow modeling. Pages 21-29 in Draft SRC-I quarterly technical report, supplement, July-September 1981. International Coal Refining Co. Allentown, Pa.

Figure 1
Flow Diagram of the Gas/Liquid Contact System

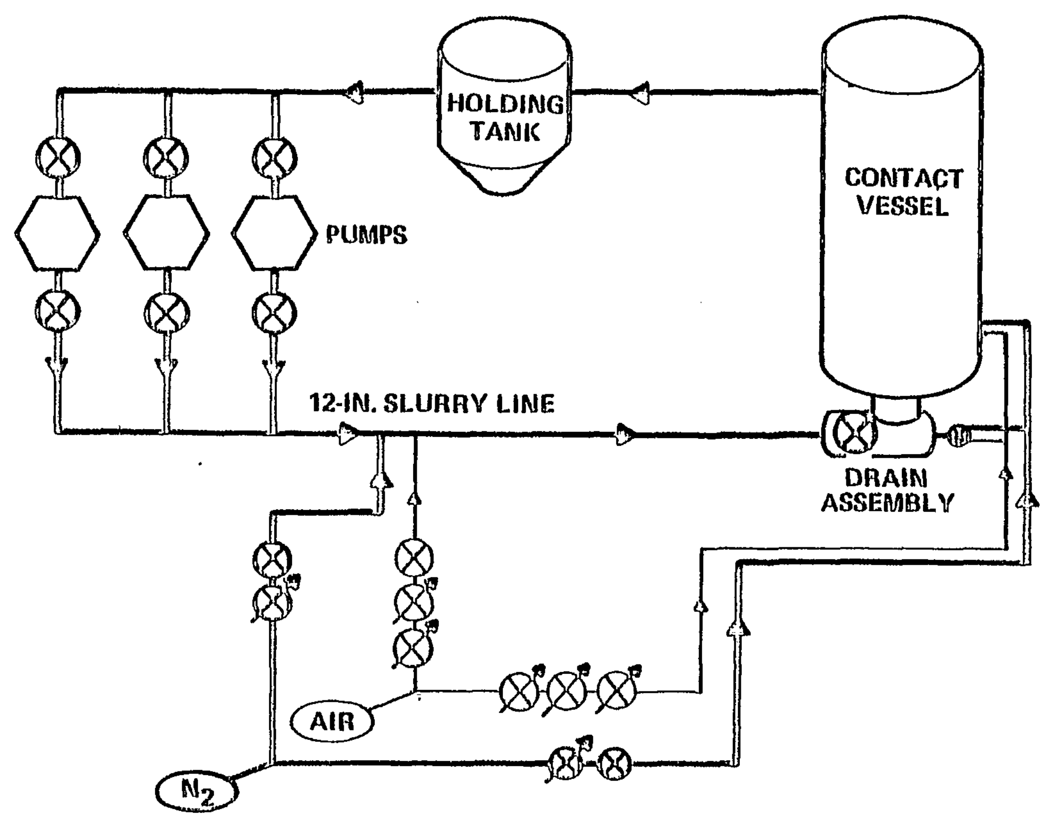


Figure 2
Measured Gas-Phase Holdup:
Batch Operation, No Internals

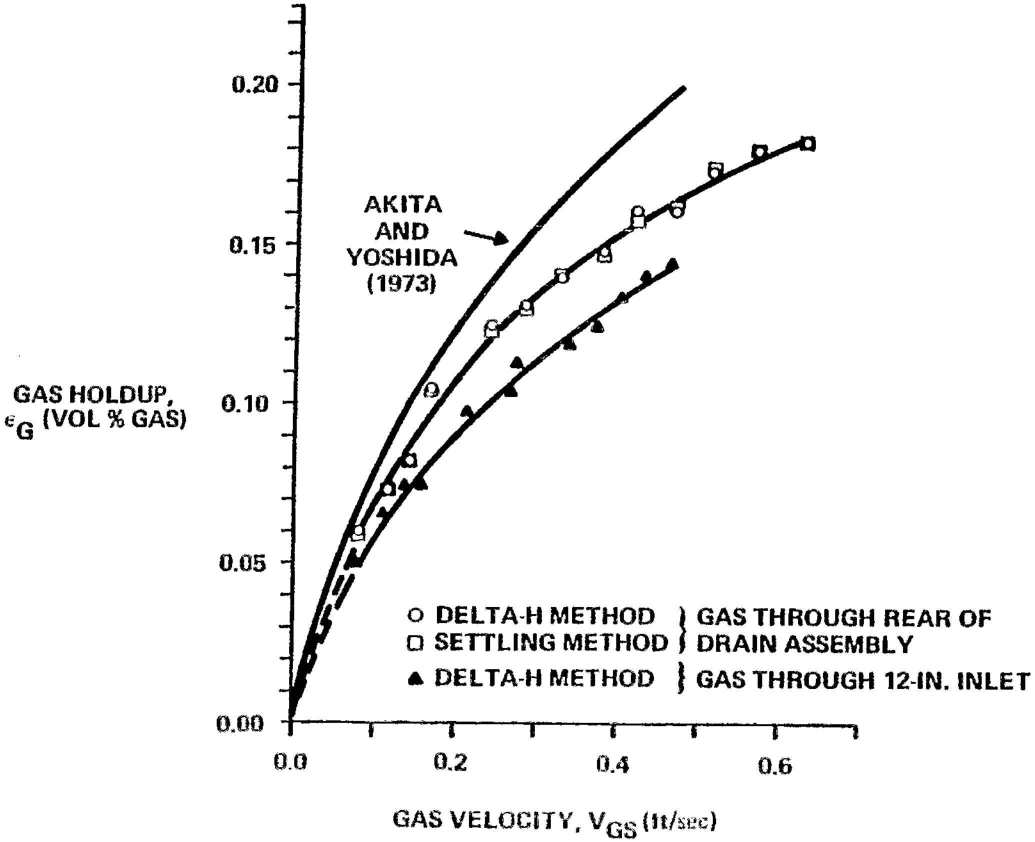
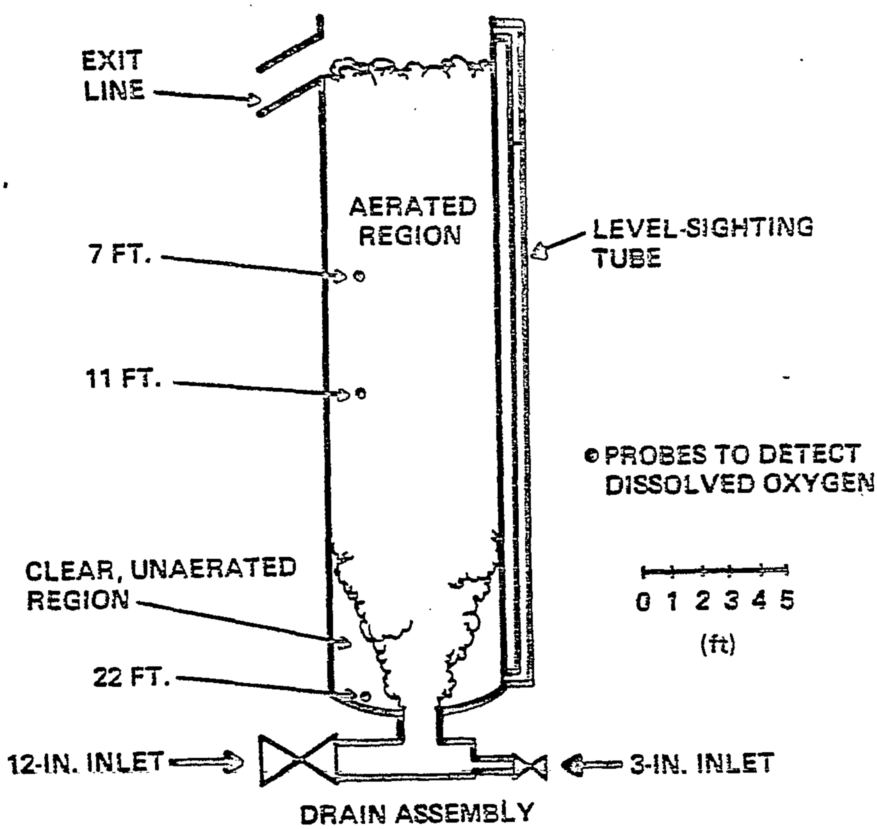


Figure 3
Flow Patterns in the Gas/Liquid Contact Vessel, No Internals



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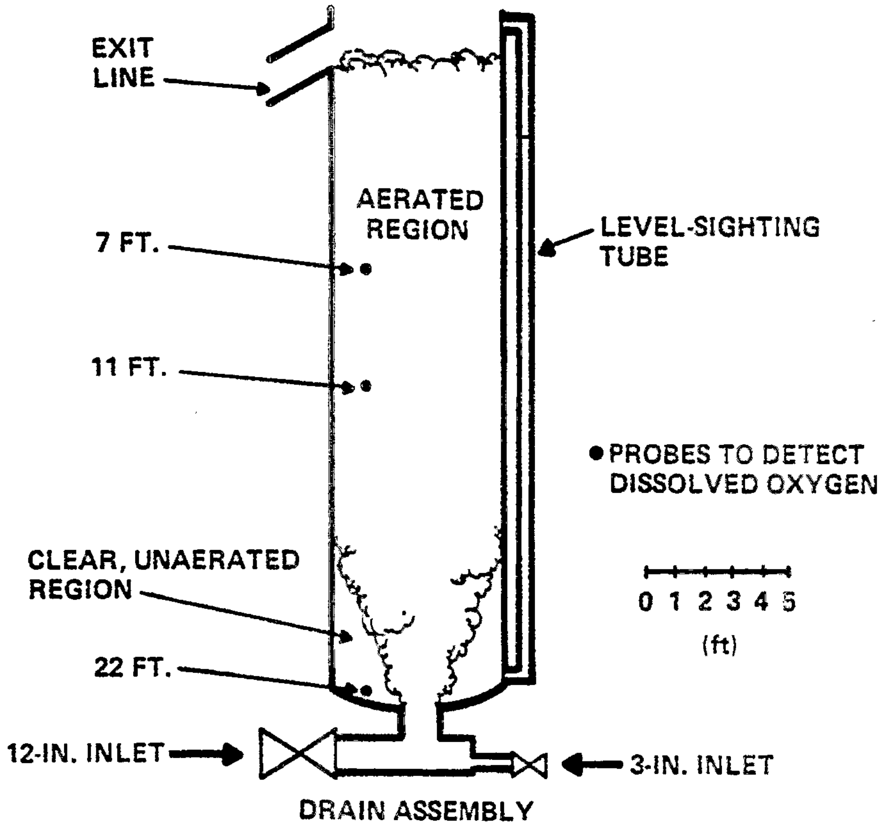


Figure 4
Typical Curves of Dissolved-Oxygen Concentration

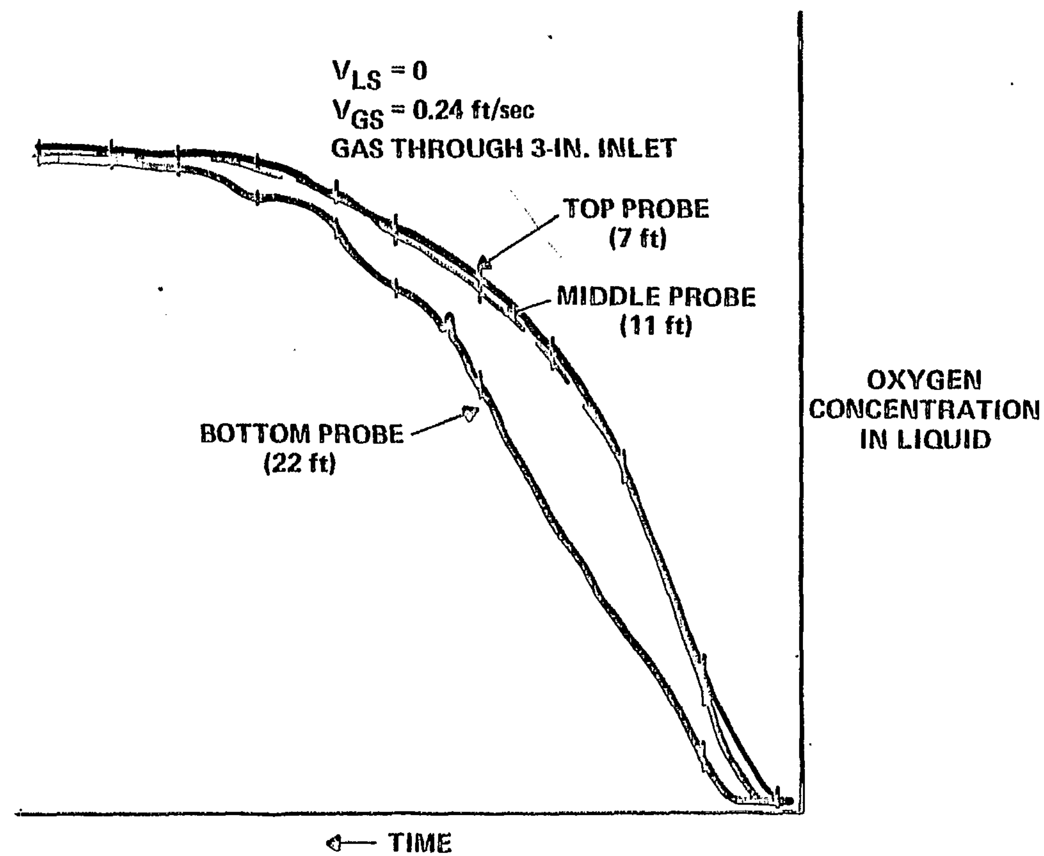
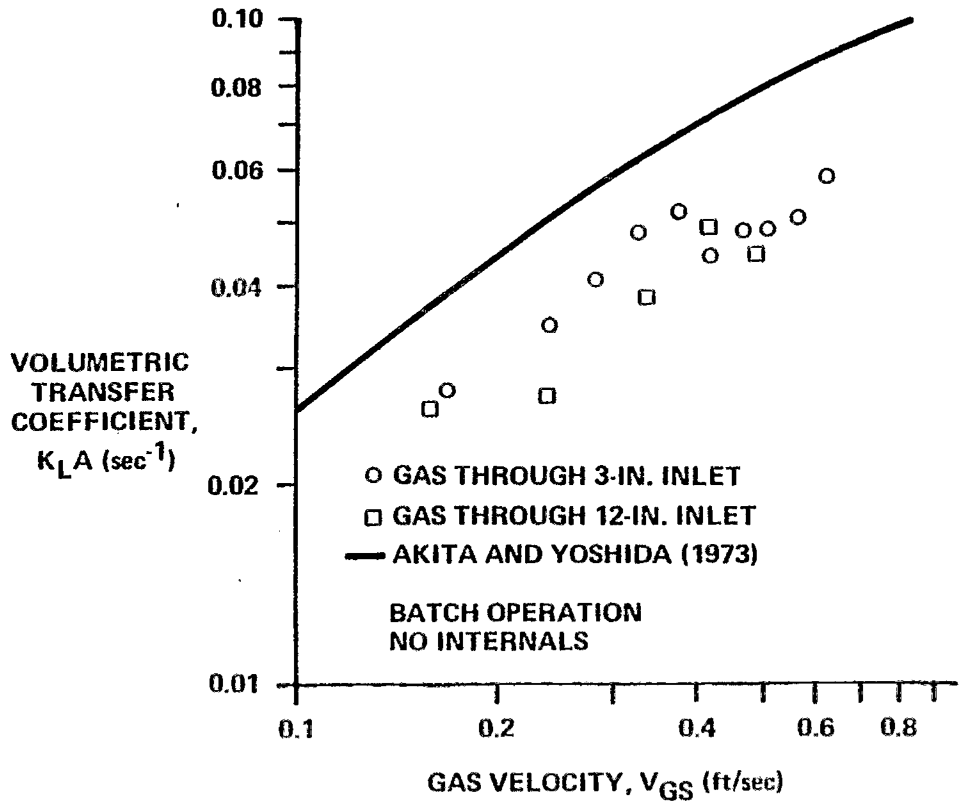


Figure 5
Measured Volumetric Mass-Transfer Coefficients, Averaged $K_L A$ Values



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