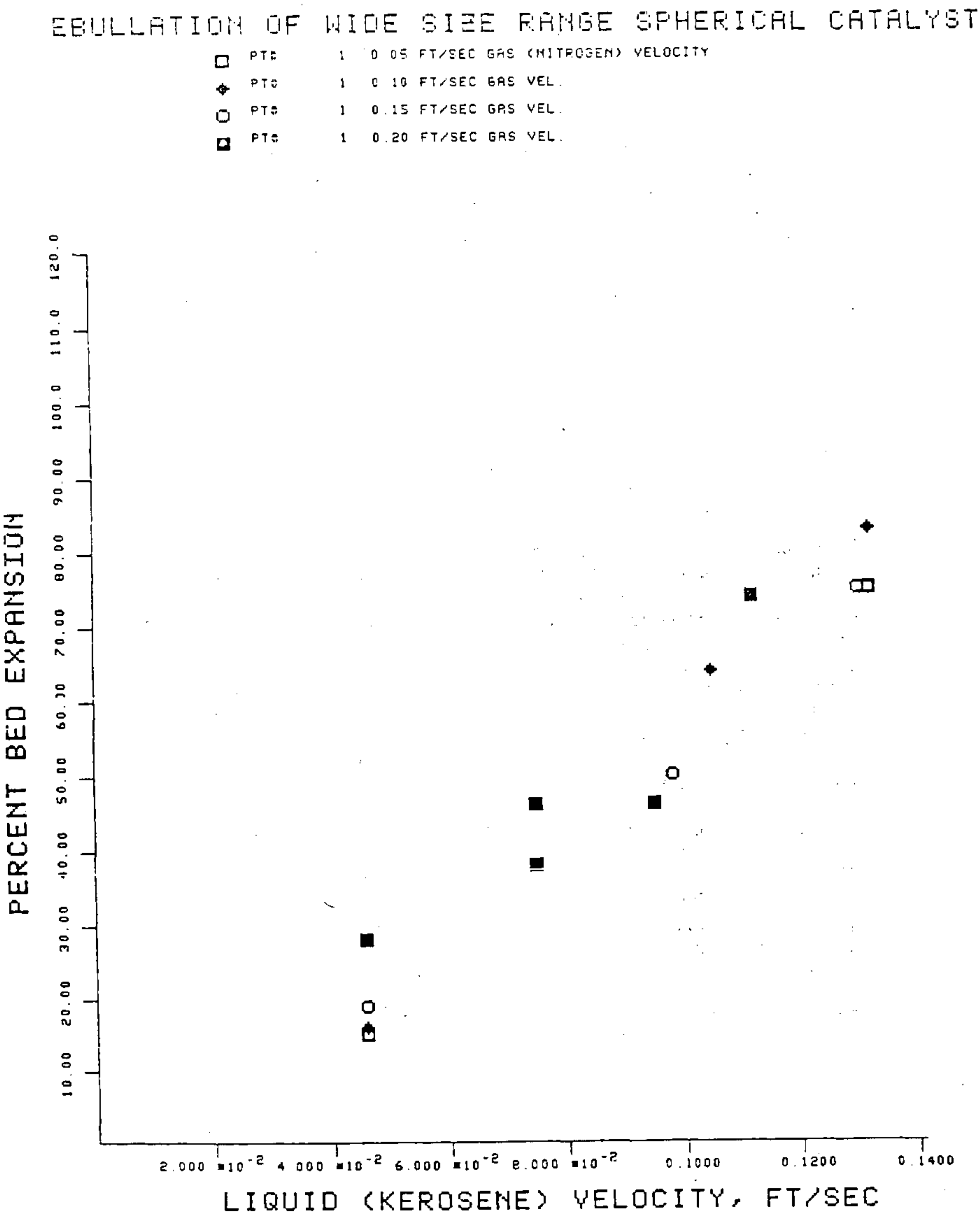
FIGURE 7


EBULLATION OF WIDE SIZE RANGE SPHERICAL CATALYST
PT# 1 0.05 FT/SEC GAS (NITROGEN) VELOCITY
PT# 1 0.10 FT/SEC GAS VEL.
PT# 1 0.15 FT/SEC GAS VEL.
PT# 1 0.20 FT/SEC GAS VEL.
PERCENT BED EXPANSION
120.0
110.0
100.0
90.00
80.00
70.00
60.00
50.00
40.00
30.00
20.00
10.00
2.000 *10^-2
4.000 *10^-2
6.000 *10^-2
8.000 *10^-2
0.1000
0.1200
0.1400
LIQUID (KEROSENE) VELOCITY, FT/SEC

5' - DIAMETER VESSEL PLENUM STUDIES

The flow pattern of the kerosene/nitrogen system was studied in a 5-foot-diameter vessel (Unit 216), similar to the reactor in the Pilot Plant. Spent catalyst from the Pilot Plant (HDS-1442A) was used in the fluidized-bed study. Flow-pattern studies were carried out for both the original plenum design and a modified design. During this study, the unit's back pressure was kept between 0 and 3 psig and the inlet temperature was kept between 75° and 85°F.

ORIGINAL PLENUM STUDY

Sight Glass Observations

Flow patterns within the 5' - diameter vessel were visually studied through eighteen sight glasses placed in the vessel wall. Below the distributor plate, the measurements taken were: wave frequency, wave amplitude, and the location of the gas-liquid interface measured as the distance from the top of the sight glass to the center line of the wave. Above the distributor plate, visual observations were made to study the flow on all sides of the vessel. The sight glasses at the 8-9 foot level and those above the recycle cup permitted comparative observations, in particular the bubble diameter and the amount of turbulence.

The first set of data were taken during a kerosene/nitrogen run. Beneath the distributor plate, a comparison was made of flow patterns observed from the sight glass above the inlet with those from opposite the inlet, at which points the greatest measurable differences were noticed. At all velocity conditions the liquid-level fluctuation opposite the inlet had approximately a 25% greater wave frequency, as shown in Figure 8, and the wave amplitude was greater by approximately 35-40% than that above the inlet. The average liquid level above the inlet was 1 - 1-1/2 inches lower than that opposite the inlet for all gas velocities, at the 0.025, 0.05 and 0.075 ft/sec liquid velocities. At 0.10 ft/sec and 0.125 ft/sec liquid velocities and gas velocities up to 0.10 ft/sec the difference in liquid levels was 1 - 1-3/4 inches. Liquid velocities of 0.10 ft/sec and 0.125 ft/sec, and gas velocities of 0.15 ft/sec and 0.20 ft/sec resulted in differences of 2 - 2-1/2 inches from one side of the vessel to the other. Obviously, the liquid level was tilted upwardly from above the inlet to opposite the inlet. This difference in liquid levels increased as the gas and liquid velocities increased, as shown in Figure 9.

The sight glasses at the distributor plate level and at the 8-9 foot level exhibited uneven distribution of turbulence and bubble size. The side opposite the inlet showed less turbulence and smaller bubbles than the other two locations, at all run conditions. The sight glasses above the recycle cup showed uniform motion and bubble diameter on all sides. Bubble diameter varied between 1/16-3/16 inch over the entire vessel. From all sight glasses it was obvious that the bubble diameter increased as the gas velocity increased.

The second study was a three-phase run with kerosene, nitrogen and catalyst. Below the distributor plate, the liquid-level fluctuation opposite the inlet had a greater wave frequency (Figure 10) and wave amplitude, by approximately 30% and 50-100%, respectively. The liquid level measurements were essentially the same as those in the kerosene and nitrogen run (Figure 11). The most significant observations were made from the sight glasses above the distributor plate. The sight glass above the inlet showed less expansion of the catalyst bed than those opposite the inlet. This occurred at liquid velocities up to 0.10 ft/sec and gas velocities up to 0.22 ft/sec. As gas or liquid velocities increased, so did the expansion within the bed but, always with less expansion above the inlet. At 0.125 ft/sec and greater, a few very large bubbles were seen throughout the bed: they were approximately 1/4-1/2 inch.

At the 8-9 foot level, the three-phase flow appeared to be the same as the flow in the kerosene/nitrogen run, with the side opposite the inlet less turbulent. Also, at this level the bubble size was very small compared with the bubbles at the same level without catalyst. At liquid velocities of 0.075 ft/sec, 0.10 ft/sec and 0.125 ft/sec and gas velocities of 0.15 ft/sec, 0.20 ft/sec and 0.22 ft/sec a few catalyst particles were observed at this level, with catalyst appearance always less in the side of the vessel opposite the inlet. Also, at this level, catalyst appeared at the 0.125 ft/sec liquid and 0.10 ft/sec, and the 0.05 ft/sec liquid and the 0.22 ft/sec gas velocity conditions. The turbulence above the recycle cup was again similar to the run without catalyst, the difference being again a smaller overall bubble diameter.

GAS HOLDUP MEASUREMENTS CONFIRMED FLOW MALDISTRIBUTION WITH ORIGINAL PLENUM DESIGN

Maldistribution was observed in both the two- and three-phase studies of the original plenum. Figure 12 shows the pressure drops measured directly above the inlet and directly across from the inlet for a liquid velocity of 0.10 ft/sec in the two-phase study. The pressure drops directly above the inlet were lower than those across from the inlet. At typical H-Coal Pilot Plant flow conditions, the difference in the pressure drops was approximately 14%. This pressure difference corresponds directly to the difference in the local volume fractions of liquid at these locations. Table 11 gives a comparison of the calculated gas holdup in the region above the inlet with the average gas holdup.* The gas holdups have been plotted in Figure 13 as functions of superficial gas velocities of 0.024 ft/sec and 0.10 ft/sec. The higher gas holdup measured directly above the inlet confirmed that a significantly larger portion of the incoming gas went through this region than through other regions.

* Arithmetic average of the values obtained for three locations spaced 120° apart.

The same variations in pressure drop were measured in the three-phase study, with the pressure drops directly above the inlet lower than the pressure drops across from the inlet (Figure 14). The local volume fractions of liquid and gas cannot be calculated without knowing the local volume fraction of catalyst. However, these differences do support the hypothesis that the original plenum design caused maldistribution in the vessel.

Maldistribution is a major operational concern at the Pilot Plant. Following a run which ended in December 1980, coke and loosely agglomerated catalyst were found in the inlet side of the reactor and the plenum. The coke layer in the reactor was about 18 inches thick and extended from several inches above the distributor grid to the 13' level. It was suspected that the coking was caused by reduced flow in the inlet side of the reactor which is related to flow maldistribution in the reactor plenum. A baffle system similar to that used for the second plenum study was recommended as a solution to the coking problem at the Pilot Plant.

TABLE 11

GAS HOLDUP
(ORIGINAL PLENUM DESIGN)

RUN NO.	GAS VELOCITY (FT/SEC)	LIQUID VELOCITY (FT/SEC)	VOLUMETRIC GAS HOLDUP DPR-5	VOLUMETRIC GAS HOLDUP AVERAGE*
2	0.025	0.044	0.103	0.060
3	0.024	0.072	0.103	0.095
4	0.024	0.101	0.128	0.129
5	0.025	0.141	0.167	0.172
6	0.025	0.187	0.192	0.229
9	0.050	0.044	0.077	0.086
10	0.049	0.075	0.128	0.116
11	0.050	0.104	0.180	0.164
12	0.050	0.139	0.205	0.216
13	0.050	0.195	0.256	0.233
15	0.075	0.042	0.103	0.086
16	0.075	0.071	0.128	0.121
17	0.075	0.099	0.218	0.176
19	0.075	0.142	0.269	0.237
20	0.076	0.191	0.282	0.254
23	0.100	0.048	0.186	0.120
24	0.100	0.072	0.205	0.157
25	0.100	0.096	0.276	0.213
26	0.100	0.142	0.314	0.265
27	0.100	0.185	0.359	0.295
29	0.123	0.042	0.115	0.097
30	0.123	0.071	0.180	0.149
32	0.123	0.094	0.212	0.179
34	0.123	0.135	0.256	0.234

* Average of DPR-5, DPR-6 and DPR-10 spaced 120° apart.

FIGURE 8

LIQUID LEVEL WAVE FREQUENCY IN KEROSENE-NITROGEN SYSTEM

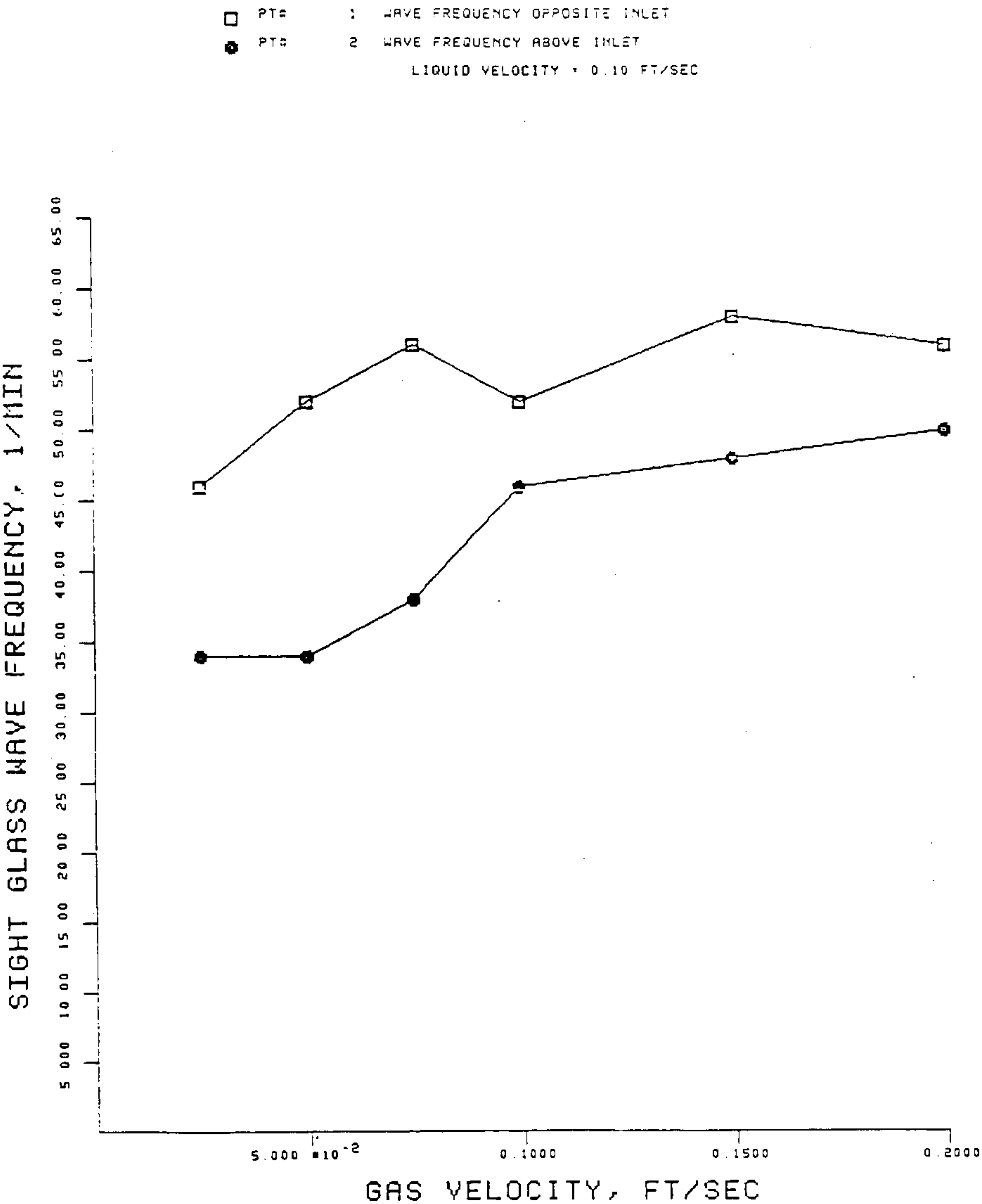


FIGURE 9

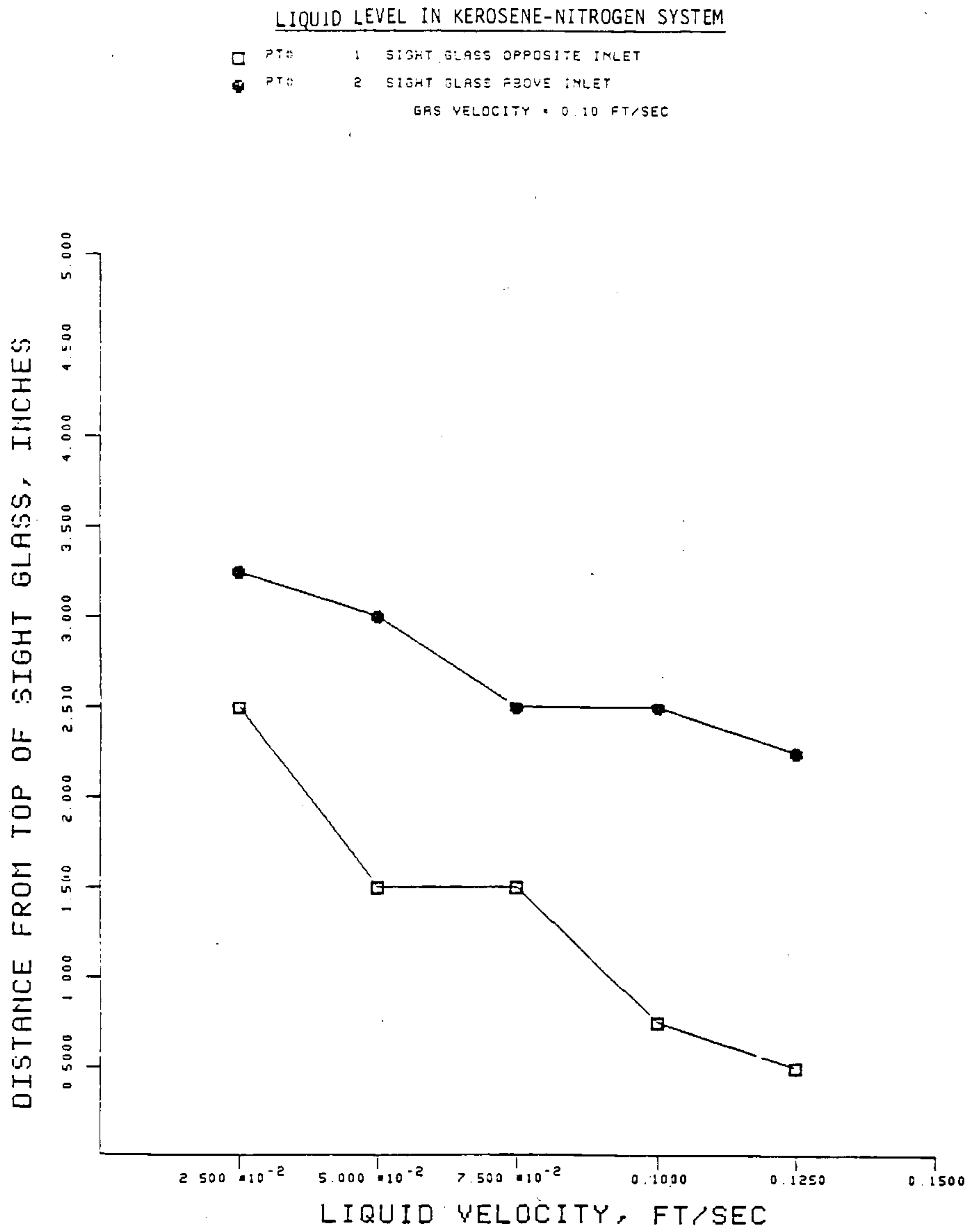

LIQUID LEVEL IN KEROSENE-NITROGEN SYSTEM
PT# 1 SIGHT GLASS OPPOSITE INLET
PT# 2 SIGHT GLASS ABOVE INLET
GAS VELOCITY = 0.10 FT/SEC
DISTANCE FROM TOP OF SIGHT GLASS, INCHES
5.000
4.500
4.000
3.500
3.000
2.500
2.000
1.500
1.000
0.5000
2.500 *10^-2
5.000 *10^-2
7.500 *10^-2
0.1000
0.1250
0.1500
LIQUID VELOCITY, FT/SEC

FIGURE 10

LIQUID LEVEL WAVE FREQUENCY IN KEROSENE-NITROGEN-CATALYST SYSTEM

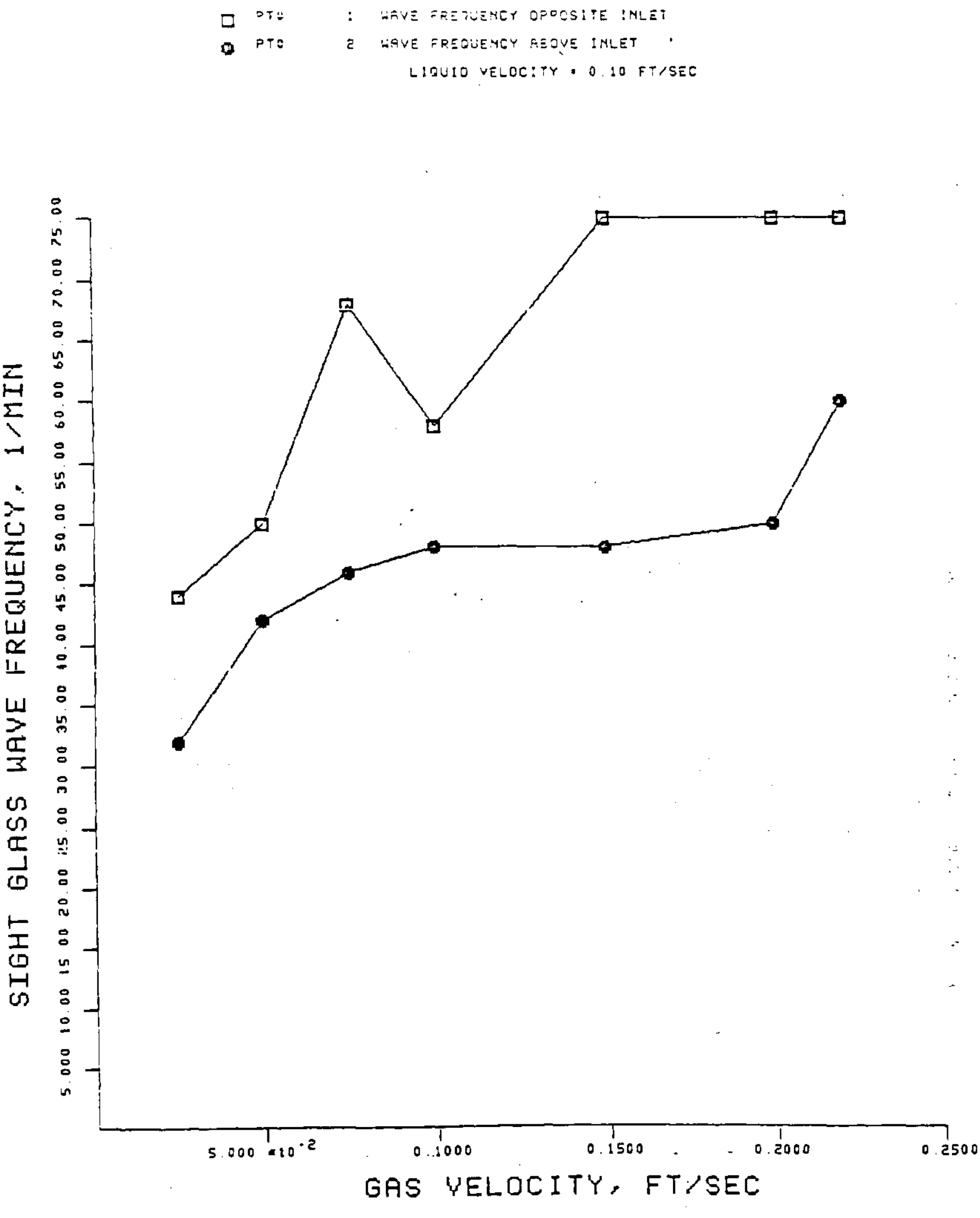


FIGURE 11

LIQUID LEVEL IN KEROSENE-NITROGEN-CATALYST SYSTEM

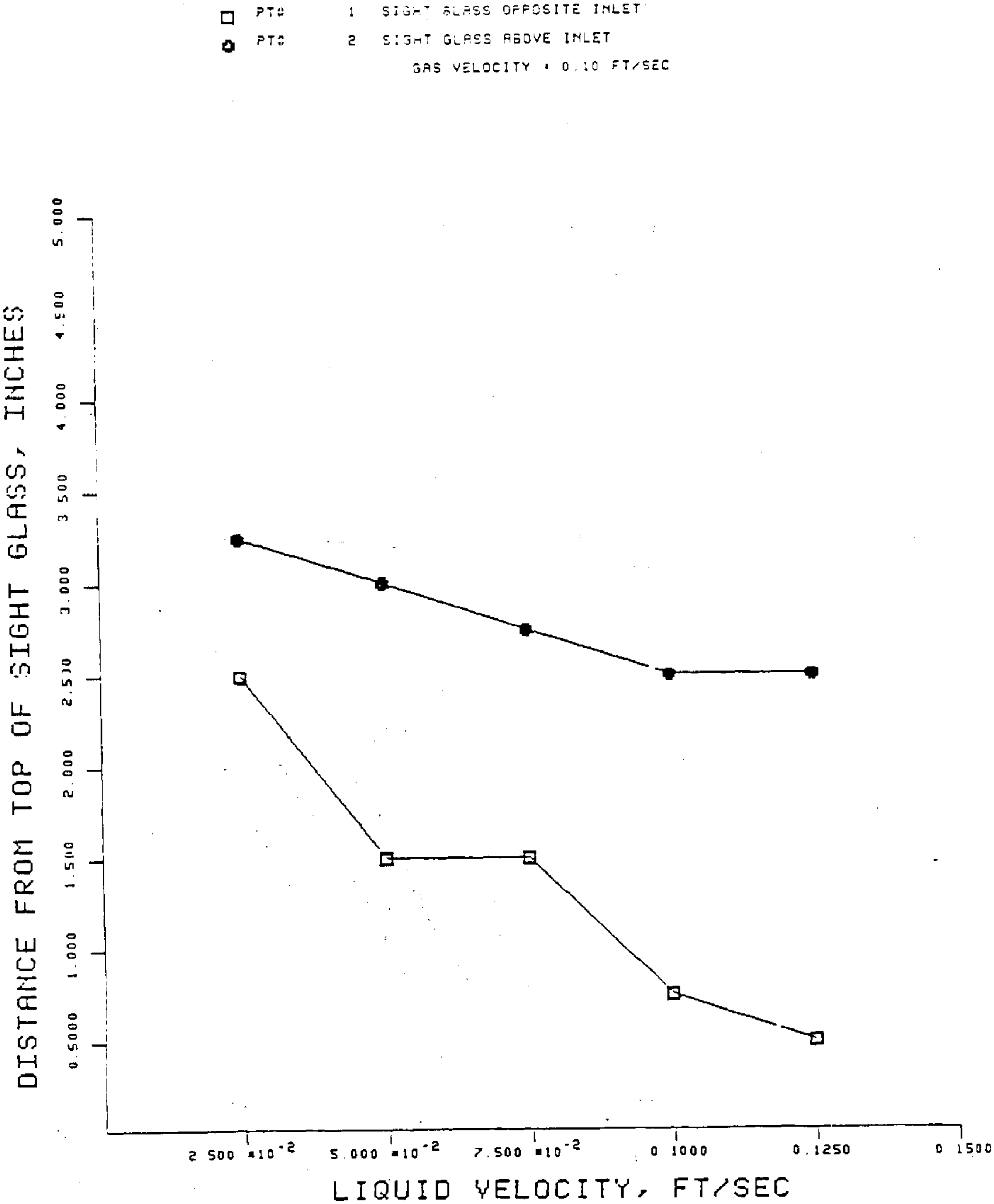


FIGURE 12

PRESSURE DROP IN KEROSENE-NITROGEN SYSTEM

□ PT# 153 INTERNAL DP (DPR5), PSI (Δp tap above inlet)

● PT# 155 INTERNAL DP (DPR10), PSI (Δp tap opposite inlet)

LIQUID VELOCITY = 0.10 FT/SEC

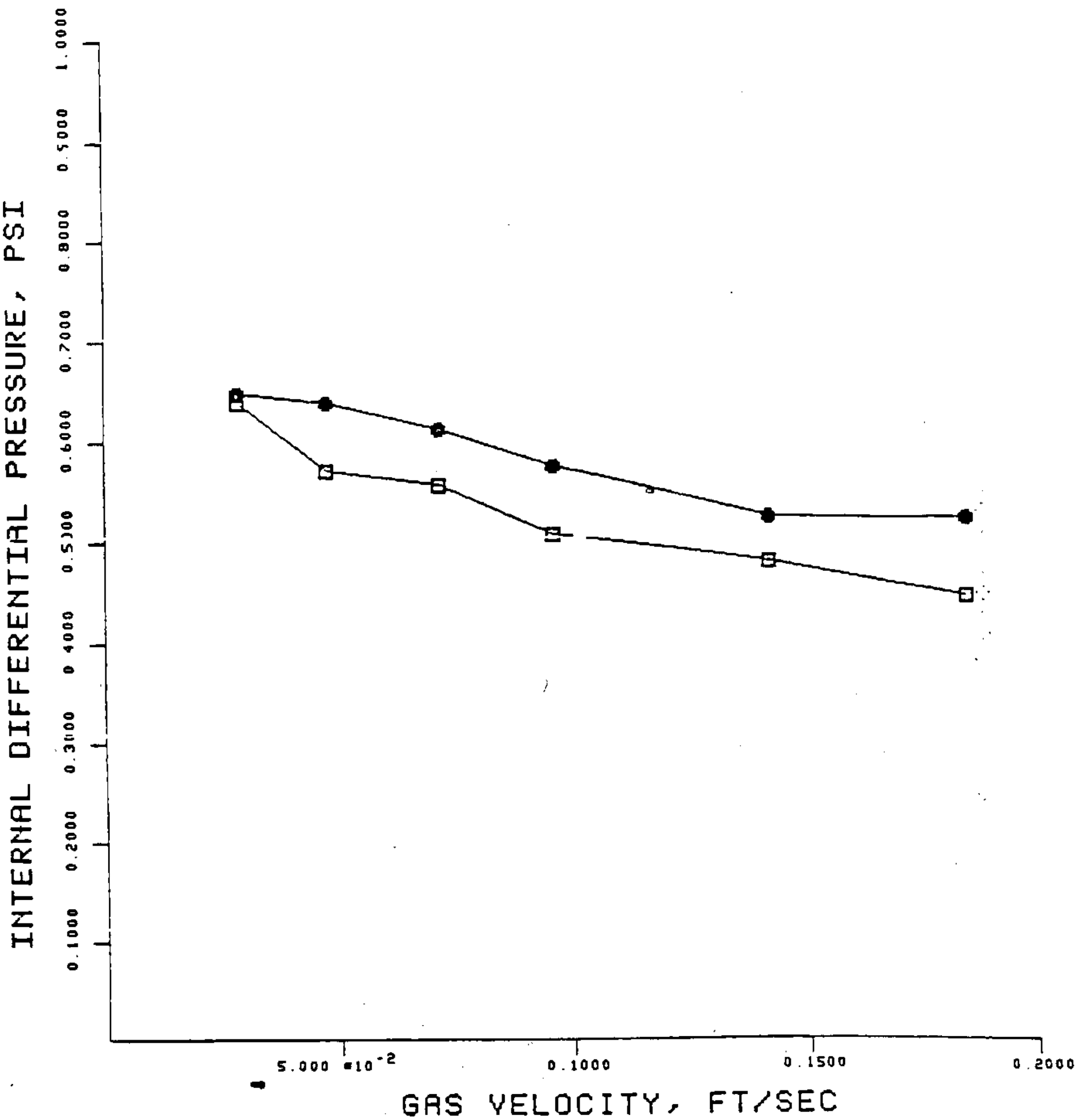
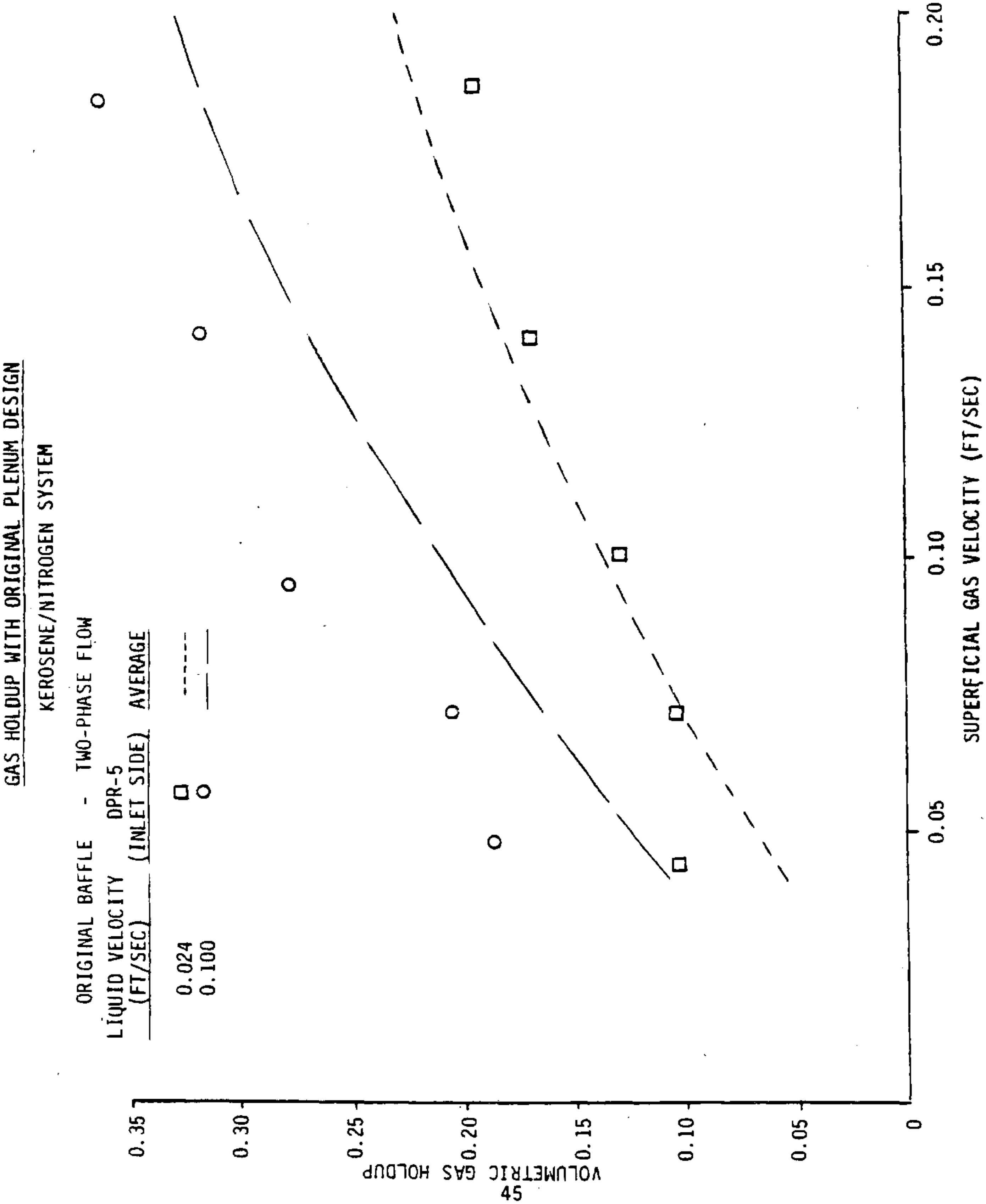


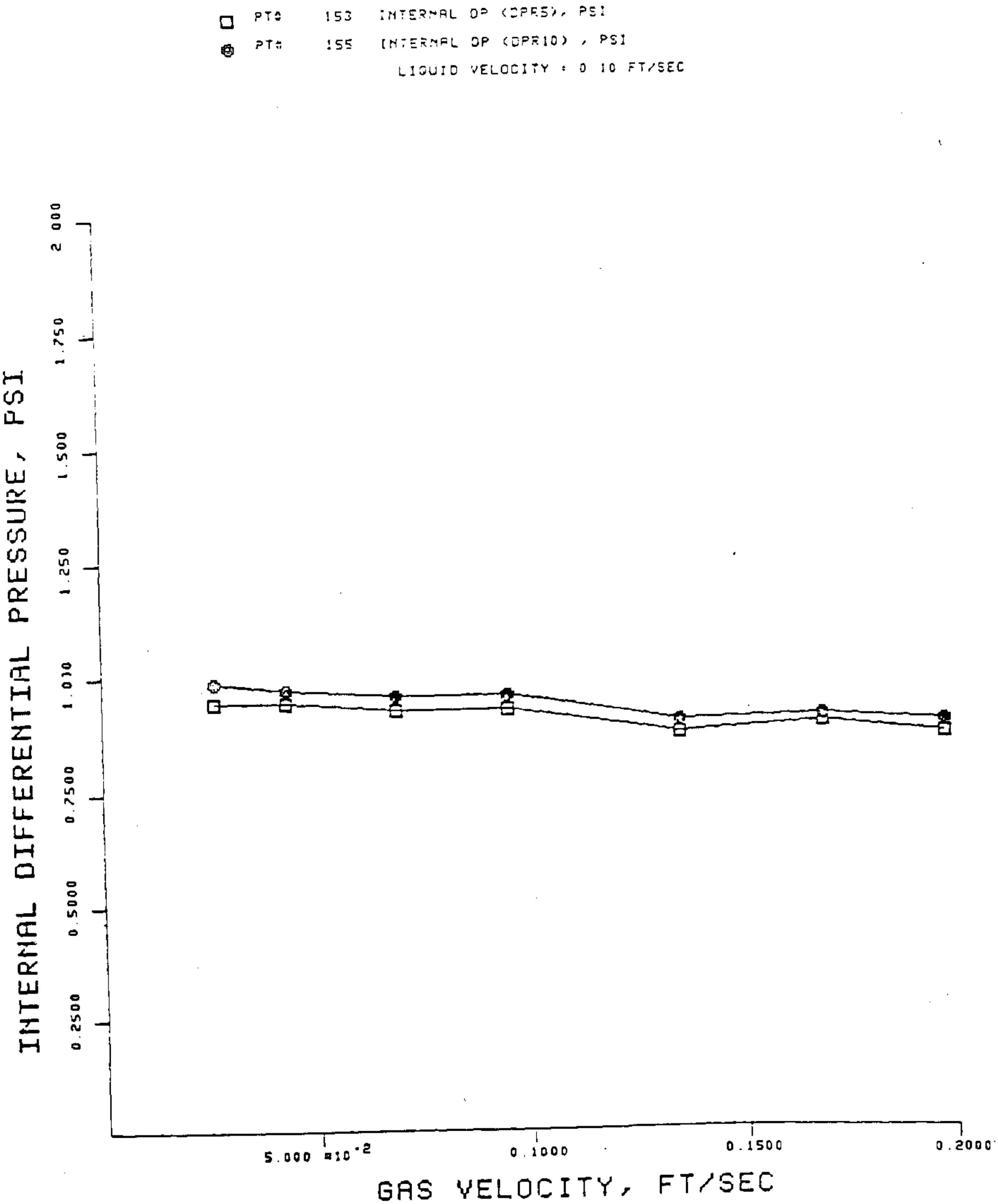
FIGURE 13



GAS HOLDUP WITH ORIGINAL PLENUM DESIGN
KEROSENE/NITROGEN SYSTEM
ORIGINAL BAFFLE - TWO-PHASE FLOW
LIQUID VELOCITY (FT/SEC)
DPR-5 (INLET SIDE)
AVERAGE
0.024
0.100
0.35
0.30
0.25
0.20
0.15
0.10
0.05
0
0.05
0.10
0.15
0.20
VOLUMETRIC GAS HOLDUP
SUPERFICIAL GAS VELOCITY (FT/SEC)

FIGURE 14

PRESSURE DROPS IN KEROSENE-NITROGEN CATALYST SYSTEM



SECOND PLENUM STUDY

SIGHT GLASS OBSERVATIONS

After the installation of a newly-designed, disk-doughnut type baffle over the inlet, a kerosene-nitrogen run was made. Similar measurements were taken in this run as in the previous run.

The sight glass directly above the inlet exhibited extreme turbulence at all liquid and gas velocities with 0.025 ft/sec to 0.05 ft/sec gas velocities. Comparing this sight glass to the one opposite the inlet at those conditions, the one above the inlet exhibited approximately a 30-40% greater wave amplitude, and a 30% greater wave frequency. Owing to the turbulent conditions above the inlet at all other liquid-gas velocities, the sight glass 45° to the left of the inlet was used for measurements during this run.

The frequency of waves in the 45° sight glass, compared to that opposite the inlet, showed differences from 4% to 20%. The wave amplitude in the 45° sight glass was anywhere from 16% to 30% greater than the sight glass opposite the inlet. The liquid level in the sight glass opposite the inlet was 3/4 inch lower than in the sight glass 45° from the inlet, for all conditions. The average difference in the liquid level was 1/4 inch. These findings indicate that the liquid beneath the distributor was tilted only slightly downward from 45° left of the inlet to opposite the inlet.

Above the distributor plate the turbulence and bubbles were minute to small, and appeared to be uniform on all sides of the vessel. The sight glasses at the 8-9 foot level again exhibited uniform motion and bubble size on all sides of the vessel. Above the recycle cup, the sight glasses also appeared consistent all around, with bubble size varying from small to 1/8 inch.

The liquid level differences were quite a bit less than those in the previous runs, 3/4-inch compared to 2-1/2 inches. The slight tilt in liquid level was opposite that of the previous runs. With this new baffle, the liquid level tilted down from the inlet side to opposite the inlet. This slightly uneven liquid level appears to have been caused by the asymmetrical flow and high kinetic energy of the inlet streams. Although flow uniformity above the distributor plate appeared to have been improved, further modifications of the plenum design may be required to eliminate the uneven flow.

NEW PLENUM DESIGN IMPROVED FLOW DISTRIBUTION

The pressure drops measured above the inlet were compared to those opposite the inlet at various combinations of gas and liquid velocities. Figure 15 presents the comparison for a liquid velocity of 0.125 ft/sec. The pressure drop above the inlet was approximately 3% less than the pressure drop opposite the inlet at typical Pilot Plant reactor flow conditions. Although a difference in pressure drop remained between the inlet side and the opposite side, it was substantially less than in the original plenum study (Figure 12). A comparison of the gas holdup in the region directly above the inlet with the average gas holdup is presented in Table 12 and Figure 16. The local variations of the gas holdup were rather small and cannot be discerned from experimental error. The improvement in inlet design appears to have corrected the flow maldistribution caused with the original design.

A new inlet nozzle similar to this design was installed in the plenum of the Pilot Plant reactor. After a 131-day run on Illinois coal (Run No. 8), the reactor was virtually coke free and the plenum was clean. The results from this extended Pilot Plant run confirmed that the new plenum design has improved flow distribution.

GAS HOLDUP FOR MODIFIED PLENUM DESIGN

RUN NO.	GAS VELOCITY (FT/SEC)	LIQUID VELOCITY (FT/SEC)	VOLUMETRIC GAS HOLDUP DPR-5	VOLUMETRIC GAS HOLDUP AVERAGE*
10	0.023	0.047	0.056	0.048
11	0.023	0.095	0.133	0.112
12	0.023	0.198	0.224	0.205
14	0.050	0.074	0.140	0.119
15	0.050	0.139	0.231	0.205
16	0.075	0.046	0.105	0.107
17	0.075	0.097	0.175	0.162
18	0.076	0.184	0.266	0.242
19	0.101	0.070	0.161	0.138
1	0.100	0.136	0.238	0.214
2	0.100	0.189	0.175	0.267
4	0.123	0.048	0.140	0.128
5	0.123	0.068	0.140	0.146
6	0.123	0.095	0.210	0.214
7	0.123	0.134	0.278	0.237
8	0.123	0.181	0.266	0.269

* Average of DPR-5, DPR-6 and DPR-10 spaced 120° apart.

FIGURE 15

PRESSURE DROPS IN KEROSENE-NITROGEN SYSTEM

(NEW BAFFLE)

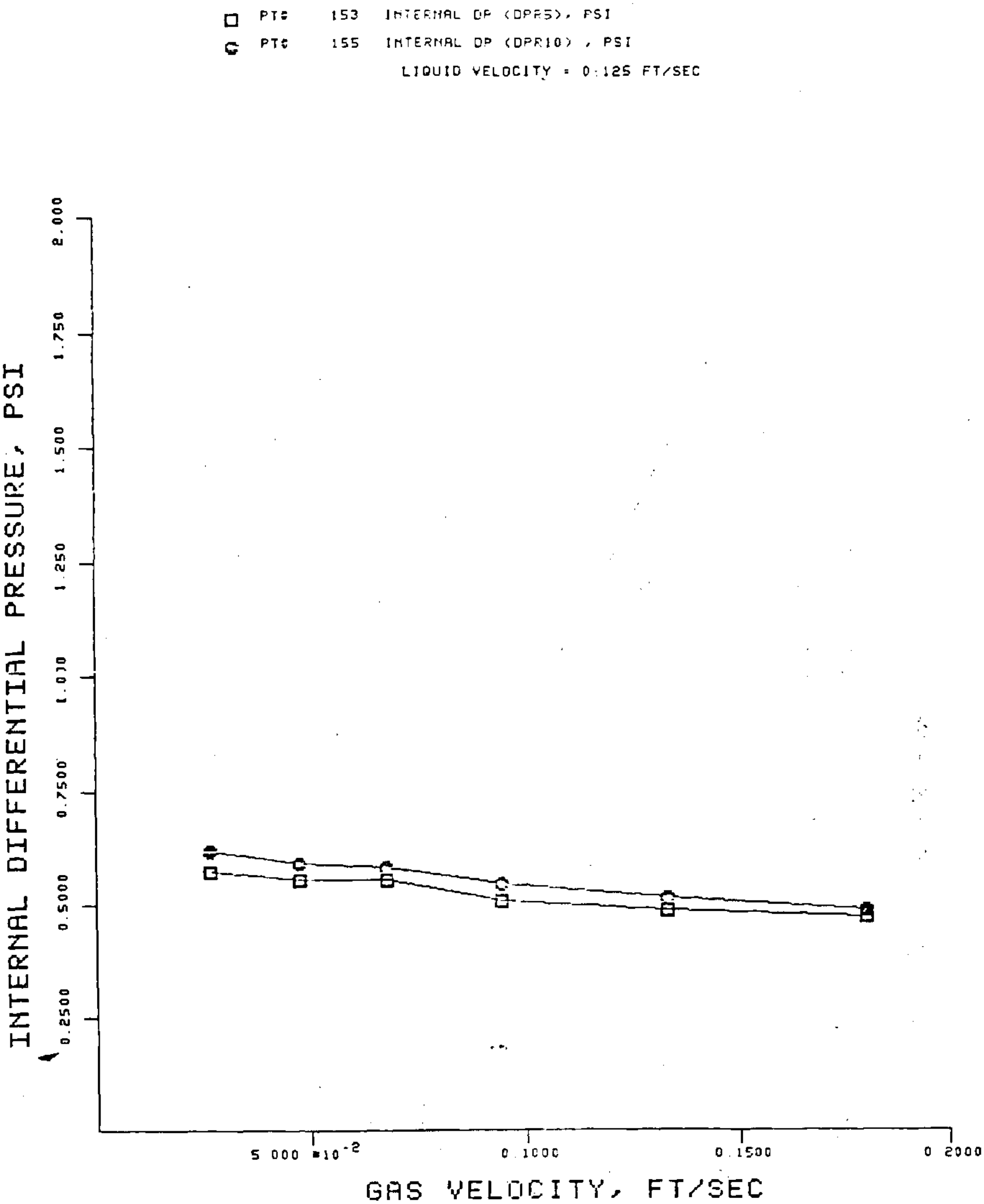
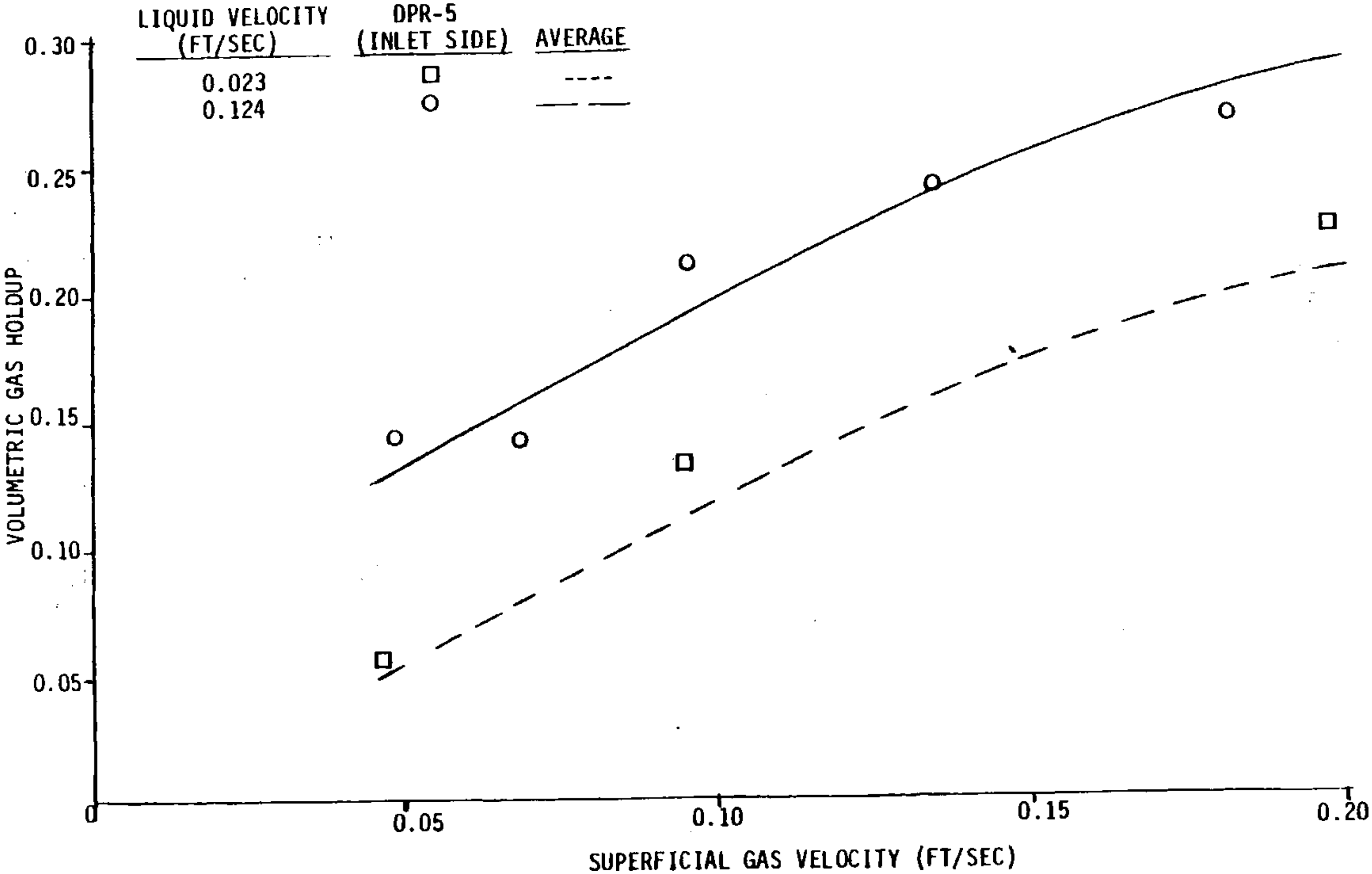


FIGURE 16

GAS HOLDUP WITH MODIFIED PLENUM DESIGN

KEROSENE/NITROGEN SYSTEM

NEW DIFFUSER - TWO-PHASE FLOW



SECTION IV

MODEL DEVELOPMENT

MODEL DEVELOPMENT

DEVELOPMENT OF BED-EXPANSION CORRELATION

The bed fluidization study employing the 6-inch-diameter glass column confirmed the strong effect of liquid (or slurry) velocity on bed expansion and the relatively weaker effect of gas velocity. It also showed that bed expansion varies with the fluidizing media. Obviously, any fluid property that affects the fluid flow pattern would have an influence on bed expansion. Viscosity of the liquid, for instance, determines the drag on the catalyst particles and can, therefore, be expected to be an important parameter. The apparent effect of liquid viscosity on bed expansion can be seen in Figure 3 where the data for the kerosene/N_2, Circosol 304/N_2, and Circosol 306/N_2 systems have been plotted. The bed expansion is seen to increase as the viscosity of the liquid increases. The data for the high-pressure and high-temperature PDU-10 run (presented in Table 13) have also been included for comparison. It is noticed that, despite the lower viscosity of the PDU-10 fluids, higher bed expansions were observed.

The effect of fines concentration on bed expansion has also been shown in Figure 2 where the data for kerosene/N_2 system with various concentrations of fines were plotted. The bed expansion increases as the fines concentration increases. One major cause for the higher bed expansions is the increase in the apparent viscosity of the slurry with increasing fines concentrations. However, the PDU-10 data for which the weight percentage of solids varies between 17% and 23% show higher bed expansions despite the lower apparent viscosities.