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SYNTHOIL HYDRODYNAMICS

Combined Third and Fourth Quarterly Report
December 1, 1975 - May 31, 1976

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Introduction

During the first six months of this contract, we concentrated upon the overall behavior of two-phase upflow in a packed bed, under conditions corresponding to the Synthoil reactor: that is, we studied the effect of macroscopic operating variables -- such as gas and liquid flowrates and packing size -- upon macroscopic performance, as indicated by average liquid holdup fraction and the pressure-drop per unit length. Although the gas flux used in the cold model was much higher than in previous studies, the correlations resulting from the previous studies seem to apply to the cold model results to the extent allowed by experimental uncertainties. A new empirical correlation was developed and also some semi-theoretical predictions were made, based on highly simplified models of the phenomena. While the new empirical correlation worked better than previous ones, the semi-theoretical predictions generally showed only qualitative agreement with observations.

In the last six months of this contract, we have examined the behavior of the cold model at a smaller scale. Experiments performed on the cold model, using an electrical conductivity probe, revealed that the local conductivity varies substantially with radial position in the bed, suggesting that gas and liquid do not exist in the same proportions at all radial positions. Previous studies have observed a slow monotonic increase in liquid holdup near to the wall, but the conductivity studies on the cold model suggest nonmonotonic behavior.

Radial Distribution of Gas and Liquid

What causes non-uniform radial distribution of liquid? Is a non-uniform distribution more likely under high flowrates or low? In this report we postulate, on an intuitive basis, that the liquid distributes radially in a fashion which leads to the lowest pressure drop per unit length. In other words, the fluid selects the path of least resistance. Application of this postulate to determine the radial distribution in a packed bed is a complex mathematical problem. Before tackling this problem, we decided to test the postulate on a simpler case: suppose we have two-phase gas-liquid flow through a pipe (without packing) which splits into two identical tubes that run parallel for some length before joining together to form a single tube again. If there is no difference between the two tubes, will the gas and liquid split uniformly?

Appendix A presents details of calculations of pressure drop and holdup in this two-parallel-tube problem as a function of the assumed gas split. The liquid split was determined so that the pressure drop per unit length was identical for both tubes. This pressure matching is necessary, since the tubes are joined to a common manifold at either end. These calculations indicated that, for low liquid flowrates, a uniform split gives the lowest pressure drop; whereas for high liquid flows, nonuniform splits give lower pressure drop. In the latter case, the liquid holdup in the two tubes is substantially different at the split giving the lowest pressure drop. Hence by extrapolation high liquid flowrates may favor maldistribution of liquid in the Synthoil reactor. On the other hand, gas flowrate seemed to have little effect on the computed distribution. Since porosity affects the hydraulic radius of the bed, radial variations in bed porosity can be modeled by taking the two tubes to have different radii.

This leads to substantially nonuniform splits and different holdups in the two tubes, as might be expected. Of course, all the predictions of this model are highly speculative and should be further investigated.

Thermal Profiles Without Internal Heat Generation

James Bachelor's memo of March 17, 1976 caused us to shift emphasis from the Synthoil reactor, where the temperature is nearly constant, to the Synthoil preheater, where large changes in temperature occur. Bachelor requested an estimate of time which the liquid spends in the preheater at the reaction temperature. Besides the difference in temperature, the preheater used in runs FB-39 is of much larger diameter and contains much larger packing than the reactor. Consequently, the gas and liquid fluxes are much smaller than any used in the correlations developed from the cold model. Furthermore, the liquid viscosity increases by orders of magnitude at lower temperature. Hence either extrapolation of previous correlations of hold-up or more cold model experiments were needed. Effect of liquid viscosity is discussed below.

Independent of the liquid holdup estimate is the determination of the temperature profile within the preheater. This is needed to obtain the reactor volume which is at the reaction temperature. A few measurements of temperature along the length of the preheater were made in runs FB-39 and FB-53.

The number of temperature measurements made was insufficient to support direct correlation of the many variables involved. Therefore, two simplified models were considered in which the reactor wall is jacketed and heated to some temperature. The first model assumes the heat transfer is determined by the wall resistance and in the second model the transfer is

determined by the resistance in the packing. Assuming the flowrate per unit area is constant, the first model (see Appendix B for details and further discussion) suggests that the length of preheater needed to heat the inlet stream to a given temperature is proportional to the radius of the preheater, while the second suggests the length is proportional to the square of the radius. Unfortunately, the experimental data are not wholly consistent with either model. Apparently a more complex model is needed, but the amount of available data are insufficient to suggest what additional effects should be introduced.

Effect of Temperature on Holdup

Within the preheater, temperatures will vary between 100°C and 450°C . In this range the viscosity of the coal/oil slurry varies between 1 and 3200 cp and the gas density is halved. What effect will these property changes have upon the liquid holdup fraction? Appendix C describes how the effect was estimated using the extensive correlations available for vertical, two-phase flows in pipes. The results presented in the appendix show that the effect is not monotonic, partly because changes in gas density and liquid viscosity have opposing effects and partly because the oil viscosity does not depend monotonically with temperature. For the specific set of conditions studied the calculated holdup varied from 0.920 at 100°C to 0.592 at 235°C (where the viscosity has a sharp maximum) and back up to 0.833 at 450°C . Considering the enormity of the viscosity change, the computed holdup changes are unexpectedly small, but certainly not insignificant.

The cold model should be adapted to permit a study of the effect of liquid viscosity. Mixtures of glycerine and water could be used as the liquid to achieve higher viscosities.

Thermal Stability of the Reactor

Reactions occurring in the Synthoil reactor are exothermic. This heat must be removed, perhaps by an external cooling jacket, to keep the temperature within safe operations limits. If a jacket is employed, heat will be transferred radially from a hot center to a cooler wall. Of course, the rate of the reaction increases rapidly with increasing temperature. Unless the heat of reaction can be removed quickly by radial transport, local temperatures in the reactor will rise, increasing the rate of reaction and rate of heat generation, which in turn causes further increases in local temperature, etc. A thermal runaway may occur.

Appendix D is a critique of the analysis by George Quarderer of Dow Chemical Co. Taking reasonable values for activation energy, thermal conductivities and reaction rates, Quarderer estimates that the largest stable reactor has a diameter of six inches. Although we bring forth several minor criticisms of his analysis, we agree with his principal conclusion: there is ample reason for concern that a six-inch reactor may require some mode of heat removal in addition to an external jacket.

APPENDIX A

Radial Distribution of Gas and Liquid

In the past, when we spoke of liquid holdup we meant that fraction of the total reactor volume, external to the packing, which is occupied by liquid. We generally have implicitly considered that, if any subset of this interstitial volume had been examined, the same fraction would be found to contain liquid: i.e. the distribution of gas and liquid within the reactor is uniform. However, experiments show that this is not true.

Beiniesch and Kessler (AIChE J., 17:1160, 1971) report variations in local liquid holdup of from 0.15 to 0.50 with radial position. Similar observations have been made by others, including those by Dennis Smith at PERC. Hence, substantially nonuniform radial distribution of gas and liquid in cocurrent flow through packed beds is strongly indicated.

What are the implications of a nonuniform radial distribution in the Synthoil reactor? One important implication is the potential for developing local hot spots within the reactor. Since the reaction is exothermic and occurs only in the liquid, the rate of reaction and thus the rate of heat generation per unit of total volume will be higher in those locations where the liquid holdup is higher. This will increase the local temperature and further accelerate the reaction unless the heat generated can be readily adsorbed (the heat capacity per volume is also higher when holdup is higher), the temperature will rise, eventually destroying the catalyst locally and tending to plug the reactor.

What causes the distribution to be nonuniform? Local variation in bed porosity is one possible cause. Near the tube wall the porosity must approach unity as the packing can only touch the wall at discrete points,

leaving all the volume there void. Experiments nearly always find that the liquid holdup is also near unity near the wall. However, evidence for a link between local holdup and local porosity is only circumstantial.

Here we shall propose an alternate explanation: by unevenly distributing gas and liquid in the radial direction, the pressure drop is minimized. In other words, the flows through the bed seek that path (or distribution) of least resistance. To put this concept to use, again consider a bundle of parallel tubes as a model for a packed bed. Assuming that all tubes are joined to a common manifold at either end (see Figure A-1) the pressure drop across each tube must be equal. However, there are many combinations of gas and liquid flowrates through individual tubes which could lead to the same pressure drop. These different combinations of flowrates lead to different holdups. A mathematical model can be immediately formulated to predict the radial holdup distribution: the problem is to find the gas and liquid flowrates through each tube in the bundle, so that the pressure drop is minimized. Constraints include: pressure drop across each tube is the same, and the sum (over all tubes) of the gas (liquid) flowrate entering each tube must equal the flowrate of gas (liquid) entering the manifold.

If the number of tubes in the bundle is large, we may treat the distributions as if they were continuous in the radial direction. This should permit a solution using the calculus of variations.

Fig. A-1. A bundle of parallel tubes, manifolded at both ends, used as a model of a packed bed.

However, almost as much can be learned by studying the simpler problem where only two parallel tubes exist, which are joined to a manifold at either end. Consider the situation shown in Figure A-2. Gas and liquid flow into the bottom manifold at

fluxes of G_g and G_l ($\text{lb}_m/\text{ft}^2\text{-sec}$), respectively, before bifurcating

into the two tubes. If G_{gi} and

G_{li} are the fluxes in tube i and

A_i is the crosssection of that tube,

then:

Fig. A-2. Two parallel tubes, connecting two manifolds.

$$G_g A = G_{g1} A_1 + G_{g2} A_2 \quad (\text{A-1a})$$

$$G_l A = G_{l1} A_1 + G_{l2} A_2 \quad (\text{A-1b})$$

$$\Delta P(G_{g1}, G_{l1}, A_1) = \Delta P(G_{g2}, G_{l2}, A_2) \quad (\text{A-2})$$

Specifying the total flowrates, G_g and G_l , and the areas, A_1 and A_2 ($A = A_1 + A_2$), equations (1) and (2) provide three relations among four unknowns (G_{g1} , G_{g2} , G_{l1} , and G_{l2}). The extra degree of freedom can be used to minimize ΔP . An outline of the program to accomplish this is shown below:

- i) Guess G_{g1}/G_g .
- ii) Compute G_{g2}/G_g from equation (1a).
- iii) Guess G_{l1}/G_l .
- iv) Compute G_{l2}/G_l from equation (1b).
- v) Calculate the pressure drop across each tube, ΔP_1 and ΔP_2 , using the correlations for single tubes.
- vi) Check equation (2); if $\Delta P_1 \neq \Delta P_2$ go to (iii).
- vii) Repeat (i) through (vi) for various guesses of G_{g1}/G_g until ΔP is minimized.

To calculate the pressure-drop in each tube, we will use the Ros correlation for two-phase cocurrent flows in pipes. This correlation was discussed at some length in Appendix C of the Second Quarterly Report. However, there we only gave the results for Region II ("slug" or "froth" flow). Recall that, if the distribution is uniform, the entire range of flowrates used in Dennis Smith's cold model experiments fall in Region II. If we now consider nonuniform distributions, the range of flowrates in individual tubes in the bundle is greatly increased so that other flow regimes will also be encountered. Correlations of liquid holdup and pressure-drop are summarized below. Nomenclature is the same as the Second Quarterly Report.

Liquid Holdup:

$$\text{bubble flow (I):} \quad N_s = F_1 + F_2 N_{VL} + \left(F_3 - \frac{F_4}{N_D} \right) \left(\frac{N_{VG}}{1 + N_{VL}} \right)^2 \quad (\text{A-3a})$$

$$\text{slug flow (II):} \quad N_s = (1 + F_5) \frac{N_{VG}^{0.982} + 0.029 N_D + F_6}{(1 + F_7 N_{VL})^2} \quad (\text{A-3b})$$

$$\text{annular flow (III):} \quad N_s = 0 \quad (\text{A-3c})$$

The constants $F_1, F_2, \dots, F_7$ are functions of N_L (liquid viscosity number) only. For water under ambient conditions $N_L = 0.00224$, which corresponds to $F_1 = 1.25$, $F_2 = 0.24$, $F_3 = 0.83$, $F_4 = 16$, $F_5 = 0.22$, $F_6 = 0.72$, and $F_7 = 0.13$.

Pressure-Drop:

$$\left. \begin{array}{l} \text{bubble flow (I)} \\ \text{slug flow (II)} \end{array} \right\} \quad f_R = f_1 f_2 / f_3 \quad (\text{A-4a})$$

$$\text{annular flow (III)} \quad f_R = f'_1 \quad (\text{A-4b})$$

$$\text{where} \quad f_1 = \frac{16}{N_{ReL}} \quad \text{if} \quad N_{ReL} \equiv \frac{2\rho_L a_t V_{SL}}{\mu_L} < 700 \quad (\text{A-5})$$

$$f'_1 = \frac{0.0791}{N_{ReG}^{0.25}} \quad \text{if} \quad 2000 < N_{ReG} \equiv \frac{2\rho_G a_t V_{SG}}{\mu_G} < 10^5 \quad (\text{A-6})$$

$$f_2(\xi) = 1 - \frac{0.8\xi}{\xi + 2} \quad \text{if} \quad \xi \equiv f_1 \frac{V_{SG}}{V_{SL}} N_D^{2/3} \quad (\text{A-7})$$

$$\text{and} \quad f_3 = 1 \quad \text{if} \quad \mu_L / \rho_L < 50 \text{ centistokes} \quad (\text{A-8})$$

Flow Regimes:

$$\text{bubble flow (I):} \quad N_{VG} \leq 2 \quad (\text{A-9a})$$

$$\text{slug flow (II):} \quad 2 < N_{VG} < 75 + 84 N_{VL}^{0.75} \quad (\text{A-9b})$$

$$\text{annular flow (III):} \quad N_{VG} \geq 75 + 84 N_{VL}^{0.75} \quad (\text{A-9c})$$

It should be pointed out that the criteria for flow regime transitions listed above are not general. Some of the constants are specific for the air-water system when $N_D < 3$.

In the actual calculations, which are discussed at the end of this section, we have neglected variations in gas density along the tube lengths and have assumed the gas in both tubes have the same density (gas was always assumed to be at 15 psig, 68°F). Dividing equations (1) by the appropriate density, we obtain:

$$V_{SG}A = V_{SG1}A_1 + V_{SG2}A_2 \quad (A-10a)$$

$$V_{SL}A = V_{SL1}A_1 + V_{SL2}A_2 \quad (A-10b)$$

where the V's are superficial tube velocities. Program input includes V_{SG} , V_{SL} and the diameter of the tube whose area is A. Judging from our previous calculations on the cold-model data for 1/4-inch packing (see p 26 of Second Quarterly Report), we experienced:

$$5 \text{ ft/sec} \leq V_{SG} \leq 26 f_{ps}$$

$$0.02 \text{ ft/sec} \leq V_{SL} \leq 0.2 f_{ps}$$

This particular data corresponds to tube radii of 0.0387 in. The diameter of the manifold for a tube pair must be 0.109 in. (so that $A = A_1 + A_2$). For all the calculations below, this diameter was taken as 0.1 in.

Figures 3, 4, and 5 illustrate the effect of liquid flowrate, gas flowrate, and non-equal tube areas. Each figure gives the fraction of the liquid flow entering tube #1 (adjusted to make $\Delta P_1 = \Delta P_2$), the pressure drop along the tubes, and the liquid holdup fraction for each tube -- all as a function of the fraction of the gas flow assumed to enter tube #1.

The most noteworthy result is the pressure-drop curves in Figure 3. As one might expect, they are symmetric about the point where $V_{SG1} = 0.5 V_{SG}$ -- i.e. where the gas flow is uniformly distributed between two equal-area tubes. The interesting result is that, for $V_{SL} < 0.2 f_{ps}$, the uniform distribution leads to a minimum pressure drop, whereas for $V_{SL} > 0.2 f_{ps}$, non-uniform distributions lead to a lower pressure drop than that for a uniform one. If our hypothesis regarding minimum pressure drop is correct, Figure 3

Fig. A-3. Pressure-drop and liquid holdup in each of the two tubes in Fig. 2 for an assumed fractional split of gas entering the bottom manifold. V_{SL1}/V_{SL} is the corresponding fractional split of liquid entering the bottom manifold. Curves 1,2,...,5 correspond to $V_{SL} = 1, 0.5, 0.2, 0.1$, and 0.05 ft/sec, respectively. Other fixed parameters include $V_{SL} = 10$ ft/sec, $A_1/A = 0.5$ and the manifold diameter = 0.1 in

suggests, for example, that when $V_{SG} = 10 f_{ps}$ and $V_{SL} = 1 f_{ps}$, about 11% of the gas and about 84% of the liquid will flow through one tube, which has then a liquid holdup fraction of 0.593, while the balance of the flow will go through the second tube, which then has a liquid holdup of only 0.182. This nonuniform distribution between the two tubes would lead to a pressure-drop of 0.49 psi/ft compared to 0.90 psi/ft and a liquid holdup of 0.274 in both tubes if the distribution was uniform.

A few comments on general features of these figures may be helpful. Discontinuities in the curves correspond to points where a transition between flow regimes occurs. In all of these data, only slug-flow or bubble-flow were experienced. It should also be pointed out that some gas distributions are physically unachievable. For example, in Figure 3 curves 4 and 5 (low liquid flowrates) are not plotted for situations where nearly all of the gas goes through one tube. This is because the pressure drop in the tube where all the gas flows will be so large that, even if all the liquid is forced through the other tube, the pressure drop in the other tube cannot be made large enough to match. Of course, for larger liquid flowrates this problem does not occur, and the curves are plotted over the full range of distributions.

In Figure 4 the liquid flow is held at that value for which higher liquid flowrates lead to a local maximum in ΔP for uniform distribution and lower liquid flows lead to a local minimum. This series of runs was designed to see if changing the gas flowrate might favor uniform or non-uniform distributions such as the effect noted above for changes in liquid flowrate. However, gas flowrate does not seem to be as important to the distribution: all of the ΔP curves in Figure 4 display a local minimum

Fig. A-4. Pressure-drop and liquid holdup in each of the two tubes in Fig. 2 for an assumed fractional split of gas entering the bottom manifold. V_{SL1}/V_{SL} is the corresponding fractional split of liquid entering the bottom manifold. Curves 1,2,...,5 correspond to V_{SG} = 50, 20, 10, 5, and 2 ft/sec, respectively. Other fixed parameters include V_{SL} = 0.2 ft/sec, A_1/A = 0.5 and the manifold diameter = 0.1 inch.

when $V_{SG1} = 0.5 V_{SG}$. Apparently, the distribution is determined primarily the liquid flowrate.

Finally, Figure 5 illustrates what happens if the two parallel tubes have unequal crossections. Below we tabulate data taken from these curves at the point where ΔP is minimum:

A_1/A	V_{SG1}/V_{SG}	V_{SL1}/V_{SL}	$\min \Delta P$ (psi/ft)	H_{L1}	H_{L2}
0.5	0.500	0.500	0.307	0.229	0.229
0.4	0.300	0.492	0.307	0.285	0.199
0.3	0.21	0.41	0.310	0.33	0.19
0.2	0.15	0.29	0.318	0.36	0.19

The results for these two non-equal tubes should suggest what might be expected to result in a packed bed if radial variations in porosity occur.

A higher porosity at one radial position in the bed means that the hydraulic radius at that position is also higher. As might have been expected, unequal tube areas favor unequal gas and liquid distributions. Less expected is that the minimum ΔP and the average system holdup $((H_{L1} + H_{L2})/2)$ also increase if the tube areas are unequal.

Fig. A-5. Pressure-drop and liquid holdup in each of the two tubes in Fig. 2 for an assumed fractional split of gas entering the bottom manifold V_{SL1}/V_{SL} is the corresponding fractional split of liquid entering the bottom manifold. Curves 1,2,3,4 correspond to $A_1/A = 0.5, 0.4, 0.3,$ and 0.2 , respectively. Other fixed parameters include $V_{SL} = 0.2$ ft/sec, $V_{SG} = 10$ ft/sec and the manifold diameter = 0.1 inch

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100 COMMON/PDROPC/D,VSG,VSL,P,KTRAC,XA1,XG1,XL1,DP(2),REG(2),REL(2), -
200 -HL(2),NREGM(2),KTRB(2),VGNO(2),VLNO(2)
300 INTEGER ICHAR(7)/'S','G','A','B',' ',' ',' ',' ',' ',' ' -1//
400 INTEGER LETRS(7)/'I','B','S','R','P','X','U','/'
500 NAMEDLIST/CHANGE/D,VSG,VSL,P,XG1,XGIF,NXG,MODX,KTRAC,KPLOT,KPRNT,-
550 -XA1
600 C
700 C-----THIS MAINLINE USES ROS'S CORRELATION OF FRICTIONAL
800 C-----PRESSURE-DROP AND HOLDUP IN TWO-PHASE FLOW THROUGH VERTICAL
900 C-----TUBES. ASSUMING TWO VERTICAL TUBES ARE ARRANGED IN PARALLEL,
1000 C-----THIS PROGRAM DETERMINES THE PRESSURE-DROP AND HOLDUP DISTRIBUTION
1100 C-----FOR AN ASSUMED GAS FLOW DISTRIBUTION.
1200 C
1300 C-----THE FOLLOWING PARAMETERS CHARACTERIZE THE FLOW BEFORE SPLITTING:
1400 C D = TUBE DIAMETER (INCHES)
1500 C VSG = SUPERFICIAL GAS VELOCITY (FPS)
1600 C VSL = SUPERFICIAL LIQUID VELOCITY (FPS)
1700 C P = AVERAGE PRESSURE IN TUBES (PSIG)
1800 C XA1 = RATIO OF TUBE AREA NO. 1 TO TOTAL AREA
1900 C
2000 C-----TRIAL VALUES FOR GAS DISTRIBUTION (XG1 = FRACTION OF TOTAL GAS
2100 C-----FLOW WHICH ENTERS TUBE NO. 1) ARE DETERMINED FROM:
2200 C XG11 = INITIAL TRIAL GAS DISTRIBUTION
2300 C XG1F = FINAL TRIAL GAS DISTRIBUTION
2400 C NXG = NUMBER OF TRIALS
2500 C MODX = 1 - UNIFORMLY SPACED TRIALS
2600 C 2 - LOGARITHMICALLY SPACED TRIALS
2700 C KTRAC = 1 - TRACE LIQUID DISTRIBUTION ON PRINTER ONLY
2800 C -1 - TRACE LN PRINTER AND TERMINAL
2900 C KPRNT = PRINT FREQUENCY FOR TERMINAL (FILE4)
3000 C KPLOT = SET =1 TO WRITE RESULTS ON FILE2 FOR PLOTTING
3100 C
3200 D=.10
3300 VSG=10.
3400 VSL=1.
3500 XA1=.5
3600 P=15.
3700 XG11=.0
3800 XG1F=1.
3900 NXG=11
4000 MODX=1
4100 KTRAC=0
4200 KPRNT=1
4300 KPLOT=0
4400 10 WRITE(4,1)
4500 1 FORMAT(' ENTER &CHANGE')
4600 READ(5,CHANGE)
4700 IF(NXG.LE.0)STOP
4800 WRITE(1,2)D,VSG,VSL,P,XA1,(1,I=1,2)
4900 2 FORMAT(' INLET TUBE DIAMETER =',F7.4,' INCH ',
5000 -' TOTAL GAS SUPERF. VEL. =',F7.4,' FT/SEC TOTAL LIQUID',
5100 -' SUPERF. VEL. =',F7.4,' FT/SEC'/' AVERAGE PRESSURE =',F5.1,
5200 -' PSIG',5X,' AREA FRACTION FOR TUBE NO. 1 =',F6.3//

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902 908/04/76 18:45:24

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5300 - ' DISTRIBUTION FACTORS PRS DROP',2(2X,1H<,2J(1H-),'TUBE #',I1, -
5400 -19(1H-),1H>))' GAS IN #1 LIQ IN #1 (PSI/FT)', -
5500 -2(2X,'G-REYNUS L-REYNUS G-VEL-# L-VEL-# HOLDUP *')/)
5600 WRITE(4,5)
5700 3 FORMAT(5X,'XG1',7X,'XL1',7X,'DP',7X,'HL1',7X,'HL2',4X,'* *')
5800 XG1=XG1I
5900 DO 40 I=1,NXG
6000 IF(NXG.EQ.1)GO TO 20
6100 XG1=XG1I+(XG1F-XG1I)*(1-I.)/(NXG-1.)
6200 IF(MODX.EQ.2)XG1=XG1I*(XG1F/XG1I)**((1-I.)/(NXG-1.))
6300 20 CALL GETXL1(XG1,DPRS)
6400 DO 30 J=1,2
6500 K=NREGM(J)
6600 30 NREGM(J)=ICHR(K)
6700 ICK=ICHR(5)
6800 IF(ABS(DP(2)-DP(1))/DP(1)).GT.1.E-3)ICK=ICHR(6)
6900 KK=KTRB(1)
7000 IF(KK.GT.0)ICK=LETRS(KK)
7100 WRITE(1,4)XG1,XL1,DP(1),ICK,(REG(J),REL(J),VGNO(J),VLNO(J),HL(J), -
7200 -NREGM(J),J=1,2)
7300 4 FORMAT(1PZE11.3,E10.2,1A1,E9.2,E10.2,2E9.1,E10.2,1X,1A1, -
7400 -ZE10.2,2E9.1,E10.2,1X,1A1)
7500 IF(MOD(1,KPRNT).EQ.0)WRITE(4,5)XG1,XL1,DP(1),ICK,HL,NREGM
7600 5 FORMAT(1PSE10.2,1A1,E9.2,E10.2,1X,2A2)
7700 NCHAR=ICHR(5)
7800 IF(I.EQ.NXG)NCHAR=ICHR(7)
7900 IF(KPLOT.NE.0)WRITE(2,6)XG1,XL1,DP(1),HL,NCHAR
8000 6 FORMAT(1PSE11.3,A4)
8100 40 CONTINUE
8200 GO TO 10
8300 END

```

GETXL ,Jc/03/75 10:21:49

```
100      SUBROUTINE GETXL1(XG,DPRS)
200      EXTERNAL GETDP
300      COMMON/PDRDPC/D,VSG,VSL,P,KTRAC,XA1,XG1,XL1,DP(2),REG(2),REL(2), -
400      -HL(2),NREGM(2),KTRB(2),VGNO(2),VLNU(2)
500      C
600      C-----USING THE GAS DISTRIBUTION SPECIFIED AS XG1, DETERMINE THE LIQUID
700      C-----DISTRIBUTION XL1 SO THAT THE PRESSURE DROP IS THE SAME IN BOTH TUB
800      C
900      XG1=XG
1000     KTRL=-1
1100     CALL SEEKER(GETDP,0.,0.1,10,1.E-3,10,1.E-6,20,KTRL,ROOT)
1200     DPRS=DP(1)
1300     KTRB(1)=KTRL
1400     RETURN
1500     END
```

GETDP ,06/03/79 16:20:30

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100 SUBROUTINE GETDP(XL,ERR,*)
200 COMMON/PDRDPC/D,VSG,VSL,P,KTRAC,XA1,XG1,XL1,DP(2),REG(2),REL(2),
300 -HL(2),NREGM(2),KTRB(2),VGNO(2),VLNO(2)
400 C
500 C-----DETERMINE PRESSURE DROP IN EACH TUBE FOR A GIVEN GAS AND LIQUID
600 C-----DISTRIBUTION
700 C
800 DENSX=29.*(P+14.7)/(10.72*528.)
900 XL1=XL
1000 DO 60 I=1,2
1100 NREGM(I)=1
1200 VG=XG1*VSG
1300 VL=XL1*VSL
1400 D1=D*SQRT(XA1)/12.
1500 IF(1.EQ.1)GO TO 10
1600 VG=(1.-XG1)*VSG
1700 VL=(1.-XL1)*VSL
1800 D1=D*SQRT(1.-XA1)/12.
1900 10 REG(I)=VG*DENSX*D1/1.23E-5
2000 REL(I)=VL*62.4*D1/6.18E-4
2100 VLNO(I)=VL/0.537
2200 VGNO(I)=VG/0.537
2300 TDNO=D1*12./0.107
2400 C
2500 C-----CHECK FOR ONE-PHASE GAS FLOW (BSL EQN 6.2-15)
2600 C
2700 IF(VL.GT.0.)GO TO 20
2800 IF(REG(I).LE.0.)GO TO 15
2900 NREGM(I)=2
3000 F=0.0791/REG(I)**0.25
3100 DP(I)=(2.*DENSX*VG*VG*F/D1)/(32.2*144.)
3200 HL(I)=0.
3300 GO TO 60
3400 15 DP(I)=2.*15.*0.2*VG*6.18E-4/(D1*D1*32.2*144.)
3500 HL(I)=0.
3600 IF(VGNO(I).LE.75.)GO TO 25
3700 GO TO 60
3800 C
3900 C-----CHECK FOR ANNULAR FLOW
4000 C
4100 20 IF(VGNO(I).GE.75.+VLNO(I)**0.75)GO TO 50
4200 F1=15.7/REL(I)
4300 F2X=F1*TDNO**(2./3.)*VG/VL
4400 F2=1.-0.5*F2X/(F2X+2.)
4500 LP(I)=(2.*F1*F2*VL*VL*62.4*(1.+VG/VL)/D1)/(32.2*144.)
4600 C
4700 C-----CHECK FOR BUBBLE FLOW
4800 C
4900 25 IF(VGNO(I).LE.2.)GO TO 30
5000 FCP=0.029*TDNO+0.72
5100 SN=1.22*(VGNO(I)**0.982+FCP)/(1.+0.15*VLNO(I))**2
5200 GO TO 40
5300 30 FCP=0.83+16./TDNO

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5400      SN=1.25+J.24*VLND(I)+F3P*(VGND(I)/(1.+VLND(I)))*2
5500      IF(VL.GT.0.)NREGM(I)=4
5600      40 SLIP=SN*J.537
5700      HL(I)=((SLIP-VG-VL)+SQRT(((SLIP-VG-VL)**2+4.*SLIP*VL)))/(2.*SLIP)
5800      DP(I)=DP(I)+62.4*HL(I)/144.
5900      GO TO 60
6000      50 F=0.0791/REG(I)**0.25
6100      HL(I)=VL/(VL+VG)
6200      DP(I)=(2.*DENS*VG*VG*F/D1)/(32.2*144.)+62.4*HL(I)/144.
6300      NREGM(I)=3
6400      60 CONTINUE
6500      ERR=DP(1)-DP(2)
6600      0
6700      0----PRINT TRACE IF REQUESTED
6800      0
6900      IF(TRACE.NE.0)WRITE(1,1)XG1,XL1,ERR,(REG(J),REL(J),VGND(J),VLND(J)-
7000      -,HL(J),NREGM(J),J=1,2)
7100      1  FORMAT(1P2E11.3,E10.2,2(2E10.2,2E9.2,E10.2,12))
7200      IF(TRACE.LT.0)WRITE(4,2)XG1,XL1,ERR,HL,NREGM
7300      2  FORMAT(1P2E11.3,X,2I2)
7400      RETURN
7500      END
```


INLET TUBE DIAMETER = 0.1000-INCH
AVERAGE PRESSURE = 10.0-PSIG

TOTAL GAS SUPERF. VEL. = 10.0000 FT/SEC
AREA FRACTION FOR TUBE NO. 1 = 0.500

TOTAL LIQUID SUPERF. VEL. = 1.0000 FT/SEC

DISTRIBUTION FACTORS		PRS DROP (PSI/FT)	< TUBE #1					> TUBE #2					
GAS IN #1	LQ IN #1		G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP
0.0000	0.048E-01	5.36E-01	0.00	4.97E-02	0.0	1.6E-00	1.00E-00	7.29E-02	9.85E-01	1.9E-01	3.1E-01	1.74E-01	S
5.632E-03	0.373E-01	5.29E-01	4.11E-00	4.98E-02	1.0E-01	1.6E-00	9.68E-01	7.25E-02	9.68E-01	1.9E-01	3.0E-01	1.74E-01	S
1.126E-02	0.392E-01	5.24E-01	8.21E-00	4.99E-02	2.1E-01	1.6E-00	9.39E-01	7.21E-02	9.57E-01	1.8E-01	3.0E-01	1.74E-01	S
1.659E-02	0.403E-01	5.20E-01	1.23E-01	5.00E-02	3.1E-01	1.6E-00	9.15E-01	7.17E-02	9.50E-01	1.8E-01	3.0E-01	1.75E-01	S
2.253E-02	0.403E-01	5.17E-01	1.64E-01	5.00E-02	4.2E-01	1.6E-00	8.96E-01	7.13E-02	9.50E-01	1.8E-01	3.0E-01	1.75E-01	S
2.816E-02	0.393E-01	5.19E-01	2.05E-01	4.99E-02	5.2E-01	1.6E-00	8.82E-01	7.03E-02	9.56E-01	1.8E-01	3.0E-01	1.75E-01	S
3.379E-02	0.375E-01	5.22E-01	2.46E-01	4.93E-02	6.3E-01	1.6E-00	8.72E-01	7.04E-02	9.67E-01	1.8E-01	3.0E-01	1.76E-01	S
3.942E-02	0.349E-01	5.26E-01	2.87E-01	4.97E-02	7.3E-01	1.6E-00	8.66E-01	7.00E-02	9.82E-01	1.8E-01	3.1E-01	1.77E-01	S
4.505E-02	0.317E-01	5.31E-01	3.28E-01	4.95E-02	8.4E-01	1.5E-00	8.63E-01	6.96E-02	1.00E-02	1.8E-01	3.1E-01	1.77E-01	S
5.068E-02	0.281E-01	5.37E-01	3.69E-01	4.93E-02	9.4E-01	1.5E-00	8.62E-01	6.92E-02	1.02E-02	1.8E-01	3.2E-01	1.78E-01	S
5.632E-02	0.240E-01	5.43E-01	4.11E-01	4.90E-02	1.0E-00	1.5E-00	8.63E-01	6.83E-02	1.05E-02	1.8E-01	3.3E-01	1.79E-01	S
6.195E-02	0.198E-01	5.51E-01	4.52E-01	4.83E-02	1.2E-00	1.5E-00	8.66E-01	6.84E-02	1.07E-02	1.7E-01	3.4E-01	1.79E-01	S
6.758E-02	0.155E-01	5.55E-01	4.93E-01	4.85E-02	1.3E-00	1.5E-00	8.69E-01	6.80E-02	1.10E-02	1.7E-01	3.4E-01	1.80E-01	S
7.321E-02	0.107E-01	5.66E-01	5.34E-01	4.82E-02	1.4E-00	1.5E-00	8.73E-01	6.78E-02	1.13E-02	1.7E-01	3.5E-01	1.81E-01	S
7.884E-02	0.060E-01	5.74E-01	5.75E-01	4.80E-02	1.5E-00	1.5E-00	8.77E-01	6.72E-02	1.15E-02	1.7E-01	3.5E-01	1.81E-01	S
8.447E-02	0.012E-01	5.82E-01	6.16E-01	4.77E-02	1.6E-00	1.5E-00	8.82E-01	6.67E-02	1.18E-02	1.7E-01	3.7E-01	1.82E-01	S
9.011E-02	7.96E-01	5.90E-01	6.57E-01	4.74E-02	1.7E-00	1.5E-00	8.86E-01	6.60E-02	1.21E-02	1.7E-01	3.8E-01	1.83E-01	S
9.574E-02	7.91E-01	5.98E-01	6.98E-01	4.71E-02	1.8E-00	1.5E-00	8.90E-01	6.59E-02	1.24E-02	1.7E-01	3.9E-01	1.84E-01	S
1.014E-01	7.66E-01	6.06E-01	7.39E-01	4.68E-02	1.9E-00	1.5E-00	8.95E-01	6.55E-02	1.27E-02	1.7E-01	4.0E-01	1.84E-01	S
1.070E-01	7.517E-01	6.14E-01	7.80E-01	4.65E-02	2.0E-00	1.5E-00	8.99E-01	6.51E-02	1.30E-02	1.7E-01	4.1E-01	1.85E-01	S

INLET TUBE DIAMETER = 0.1000-INCH
AVERAGE PRESSURE = 15.0-PSIG

TOTAL GAS SUPERF. VEL. = 10.0000 FT/SEC
AREA FRACTION FOR TUBE NO. 1 = 0.500

TOTAL LIQUID SUPERF. VEL. = 1.0000 FT/SEC

DISTRIBUTION FACTORS		PRS DROP	TUBE #1						TUBE #2					
GAS IN #1	LIQ IN #1	(PSI/FT)	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*
8.930E-01	2.183E-01	6.14E-01	6.51E 02	1.30E 02	1.7E 01	4.1E-01	1.85E-01	S	7.80E 01	4.65E 02	2.0E 00	1.5E 00	8.99E-01	B
8.986E-01	2.134E-01	6.06E-01	6.55E 02	1.27E 02	1.7E 01	4.0E-01	1.84E-01	S	7.39E 01	4.68E 02	1.9E 00	1.5E 00	8.95E-01	B
9.043E-01	2.085E-01	5.98E-01	6.59E 02	1.24E 02	1.7E 01	3.9E-01	1.84E-01	S	6.98E 01	4.71E 02	1.8E 00	1.5E 00	8.90E-01	B
9.099E-01	2.036E-01	5.90E-01	6.63E 02	1.21E 02	1.7E 01	3.8E-01	1.83E-01	S	6.57E 01	4.74E 02	1.7E 00	1.5E 00	8.86E-01	B
9.155E-01	1.988E-01	5.82E-01	6.67E 02	1.18E 02	1.7E 01	3.7E-01	1.82E-01	S	6.16E 01	4.77E 02	1.6E 00	1.5E 00	8.82E-01	B
9.212E-01	1.940E-01	5.74E-01	6.72E 02	1.15E 02	1.7E 01	3.6E-01	1.81E-01	S	5.75E 01	4.80E 02	1.5E 00	1.5E 00	8.77E-01	B
9.268E-01	1.893E-01	5.66E-01	6.76E 02	1.13E 02	1.7E 01	3.5E-01	1.81E-01	S	5.34E 01	4.82E 02	1.4E 00	1.5E 00	8.73E-01	B
9.324E-01	1.847E-01	5.58E-01	6.80E 02	1.10E 02	1.7E 01	3.4E-01	1.80E-01	S	4.93E 01	4.85E 02	1.3E 00	1.5E 00	8.69E-01	B
9.381E-01	1.802E-01	5.51E-01	6.84E 02	1.07E 02	1.7E 01	3.4E-01	1.79E-01	S	4.52E 01	4.88E 02	1.2E 00	1.5E 00	8.66E-01	B
9.437E-01	1.760E-01	5.43E-01	6.88E 02	1.05E 02	1.8E 01	3.3E-01	1.79E-01	S	4.11E 01	4.90E 02	1.0E 00	1.5E 00	8.63E-01	B
9.493E-01	1.719E-01	5.37E-01	6.92E 02	1.02E 02	1.8E 01	3.2E-01	1.78E-01	S	3.69E 01	4.93E 02	9.4E-01	1.5E 00	8.62E-01	B
9.549E-01	1.683E-01	5.31E-01	6.96E 02	1.00E 02	1.8E 01	3.1E-01	1.77E-01	S	3.28E 01	4.95E 02	8.4E-01	1.5E 00	8.60E-01	B
9.606E-01	1.651E-01	5.26E-01	7.00E 02	9.82E 01	1.8E 01	3.1E-01	1.77E-01	S	2.87E 01	4.97E 02	7.3E-01	1.6E 00	8.58E-01	B
9.662E-01	1.625E-01	5.22E-01	7.04E 02	9.67E 01	1.8E 01	3.0E-01	1.76E-01	S	2.46E 01	4.98E 02	6.3E-01	1.6E 00	8.72E-01	B
9.718E-01	1.607E-01	5.19E-01	7.08E 02	9.56E 01	1.8E 01	3.0E-01	1.76E-01	S	2.05E 01	4.99E 02	5.2E-01	1.6E 00	8.82E-01	B
9.775E-01	1.597E-01	5.19E-01	7.13E 02	9.50E 01	1.8E 01	3.0E-01	1.75E-01	S	1.64E 01	5.00E 02	4.2E-01	1.6E 00	8.96E-01	B
9.831E-01	1.597E-01	5.20E-01	7.17E 02	9.50E 01	1.8E 01	3.0E-01	1.75E-01	S	1.23E 01	5.00E 02	3.1E-01	1.6E 00	9.15E-01	B
9.887E-01	1.608E-01	5.24E-01	7.21E 02	9.57E 01	1.8E 01	3.0E-01	1.74E-01	S	8.21E 00	4.99E 02	2.1E-01	1.6E 00	9.39E-01	B
9.944E-01	1.627E-01	5.29E-01	7.25E 02	9.68E 01	1.9E 01	3.0E-01	1.74E-01	S	4.11E 00	4.93E 02	1.0E-01	1.6E 00	9.66E-01	B
1.000E 00	1.652E-01	5.36E-01	7.29E 02	9.83E 01	1.9E 01	3.1E-01	1.74E-01	S	4.35E-00	4.97E 02	1.1E-00	1.6E 00	1.00E 00	B

INLET 1 DIAMETER = 0.1000-INCH
 AVERAGE PRESSURE = 15.0-PSIG
 AREA FRACTION FOR TUBE NO. 1 = 0.500
 TOTAL GAS SUPERF. VEL. = 10.0000 FT/SEC
 TOTAL LIQUID SUPERF. VEL. = 3.5000 FT/SEC

DISTRIBUTION FACTORS		PRS DRDP		<-----TUBE #1----->		<-----TUBE #2----->		* HOLDUP		G-REYNOS		L-REYNOS		G-VEL-#		L-VEL-#		HOLDUP		* HOLDUP									
GAS IN #1	LIT IN #1	(PSI/FT)	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	* HOLDUP	G-REYNOS	L-REYNOS	G-VEL-#	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	* HOLDUP	GAS IN #1	LIT IN #1	(PSI/FT)	G-REYNOS	L-REYNOS	G-VEL-#	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	* HOLDUP		
0.000	7.268E-01	4.78E-01	0.00	2.16E	02	0.0	6.8E-01	1.00E	00	8	7.29E	02	8.13E	01	1.9E	01	2.5E	01	1.73E	01	1.73E	01	1.73E	01	2.5E	01	1.73E	01	
5.532E-03	7.387E-01	4.65E-01	4.11E	00	2.20E	02	1.0E-01	6.9E-01	9.53E	01	8	7.25E	02	7.71E	01	1.9E	01	2.4E	01	1.73E	01	1.73E	01	1.73E	01	2.4E	01	1.73E	01
1.140E-02	7.458E-01	4.57E-01	8.21E	00	2.22E	02	2.1E-01	6.9E-01	9.18E	01	8	7.21E	02	7.56E	01	1.8E	01	2.4E	01	1.73E	01	1.73E	01	1.73E	01	2.4E	01	1.73E	01
1.689E-02	7.459E-01	4.55E-01	1.23E	01	2.22E	02	3.1E-01	7.3E-01	8.97E	01	8	7.17E	02	7.53E	01	1.8E	01	2.4E	01	1.74E	01	1.74E	01	1.74E	01	2.4E	01	1.74E	01
2.253E-02	7.431E-01	4.57E-01	1.64E	01	2.21E	02	4.2E-01	6.9E-01	8.88E	01	8	7.13E	02	7.64E	01	1.8E	01	2.4E	01	1.74E	01	1.74E	01	1.74E	01	2.4E	01	1.74E	01
2.816E-02	7.359E-01	4.63E-01	2.05E	01	2.19E	02	5.2E-01	6.9E-01	8.86E	01	8	7.08E	02	7.86E	01	1.8E	01	2.5E	01	1.75E	01	1.75E	01	1.75E	01	2.5E	01	1.75E	01
3.579E-02	7.269E-01	4.70E-01	2.46E	01	2.16E	02	6.3E-01	6.8E-01	8.90E	01	8	7.04E	02	8.12E	01	1.8E	01	2.5E	01	1.75E	01	1.75E	01	1.75E	01	2.5E	01	1.75E	01
3.944E-02	7.171E-01	4.78E-01	2.87E	01	2.13E	02	7.3E-01	6.7E-01	8.95E	01	8	7.00E	02	8.42E	01	1.8E	01	2.6E	01	1.76E	01	1.76E	01	1.76E	01	2.6E	01	1.76E	01
4.505E-02	7.070E-01	4.87E-01	3.28E	01	2.10E	02	8.4E-01	6.6E-01	9.02E	01	8	6.96E	02	8.72E	01	1.8E	01	2.7E	01	1.77E	01	1.77E	01	1.77E	01	2.7E	01	1.77E	01
5.068E-02	6.970E-01	4.95E-01	3.69E	01	2.07E	02	9.4E-01	6.5E-01	9.09E	01	8	6.92E	02	9.01E	01	1.8E	01	2.8E	01	1.77E	01	1.77E	01	1.77E	01	2.8E	01	1.77E	01
5.552E-02	6.872E-01	5.04E-01	4.11E	01	2.04E	02	1.0E	00	9.15E	01	8	6.88E	02	9.31E	01	1.8E	01	2.9E	01	1.78E	01	1.78E	01	1.78E	01	2.9E	01	1.78E	01
6.195E-02	6.777E-01	5.12E-01	4.52E	01	2.02E	02	1.2E	00	9.21E	01	8	6.84E	02	9.59E	01	1.7E	01	3.0E	01	1.79E	01	1.79E	01	1.79E	01	3.0E	01	1.79E	01
6.758E-02	6.685E-01	5.20E-01	4.93E	01	1.99E	02	1.3E	00	9.27E	01	8	6.80E	02	9.86E	01	1.7E	01	3.1E	01	1.79E	01	1.79E	01	1.79E	01	3.1E	01	1.79E	01
7.321E-02	6.595E-01	5.27E-01	5.34E	01	1.96E	02	1.4E	00	9.32E	01	8	6.76E	02	1.01E	02	1.7E	01	3.2E	01	1.80E	01	1.80E	01	1.80E	01	3.2E	01	1.80E	01
7.894E-02	6.508E-01	5.35E-01	5.75E	01	1.94E	02	1.5E	00	9.36E	01	8	6.72E	02	1.04E	02	1.7E	01	3.3E	01	1.81E	01	1.81E	01	1.81E	01	3.3E	01	1.81E	01
8.447E-02	6.424E-01	5.42E-01	6.16E	01	1.91E	02	1.6E	00	9.40E	01	8	6.67E	02	1.06E	02	1.7E	01	3.4E	01	1.81E	01	1.81E	01	1.81E	01	3.4E	01	1.81E	01
9.011E-02	6.341E-01	5.49E-01	6.57E	01	1.89E	02	1.7E	00	9.44E	01	8	6.63E	02	1.09E	02	1.7E	01	3.4E	01	1.82E	01	1.82E	01	1.82E	01	3.4E	01	1.82E	01
9.574E-02	6.260E-01	5.55E-01	6.98E	01	1.86E	02	1.8E	00	9.47E	01	8	6.59E	02	1.11E	02	1.7E	01	3.5E	01	1.83E	01	1.83E	01	1.83E	01	3.5E	01	1.83E	01
1.070E-01	6.181E-01	5.62E-01	7.39E	01	1.84E	02	1.9E	00	9.50E	01	8	6.55E	02	1.14E	02	1.7E	01	3.6E	01	1.84E	01	1.84E	01	1.84E	01	3.6E	01	1.84E	01

INLET DIAMETER = 0.1000-INCH
AVERAGE PRESSURE = 15.0-PSIG

TOTAL GAS SUPERF. VEL. = 10.0000 FT/SEC
AREA FRACTION FOR TUBE NO. 1 = 0.500

TOTAL LIQUID SUPERF. VEL. = 0.5000 FT/SEC

DISTRIBUTION FACTORS		PKS DROP (PSI/FT)	TUBE #1					*	TUBE #2					*
GAS IN #1	LIQ IN #1		G-REYNDS	L-REYNDS	G-VEL-#	L-VEL-#	HOLDUP		G-REYNDS	L-REYNDS	G-VEL-#	L-VEL-#	HOLDUP	
8.930E-01	3.897E-01	5.58E-01	6.51E 02	1.16E 02	1.7E 01	3.6E-01	1.84E-01 S		7.80E 01	1.32E 02	2.0E 00	5.7E-01	9.53E-01 B	
8.986E-01	3.819E-01	5.62E-01	6.55E 02	1.14E 02	1.7E 01	3.6E-01	1.83E-01 S		7.39E 01	1.84E 02	1.9E 00	5.8E-01	9.50E-01 B	
9.043E-01	3.740E-01	5.55E-01	6.59E 02	1.11E 02	1.7E 01	3.5E-01	1.83E-01 S		6.98E 01	1.86E 02	1.8E 00	5.3E-01	9.47E-01 B	
9.099E-01	3.659E-01	5.49E-01	6.63E 02	1.09E 02	1.7E 01	3.4E-01	1.82E-01 S		6.57E 01	1.89E 02	1.7E 00	5.9E-01	9.44E-01 B	
9.155E-01	3.576E-01	5.42E-01	6.67E 02	1.06E 02	1.7E 01	3.3E-01	1.81E-01 S		6.16E 01	1.91E 02	1.6E 00	6.0E-01	9.40E-01 B	
9.212E-01	3.492E-01	5.35E-01	6.72E 02	1.04E 02	1.7E 01	3.3E-01	1.81E-01 S		5.75E 01	1.94E 02	1.5E 00	6.1E-01	9.36E-01 B	
9.268E-01	3.405E-01	5.27E-01	6.76E 02	1.01E 02	1.7E 01	3.2E-01	1.80E-01 S		5.34E 01	1.96E 02	1.4E 00	6.1E-01	9.32E-01 B	
9.324E-01	3.315E-01	5.20E-01	6.80E 02	9.86E 01	1.7E 01	3.1E-01	1.79E-01 S		4.93E 01	1.99E 02	1.3E 00	6.2E-01	9.27E-01 B	
9.381E-01	3.223E-01	5.12E-01	6.84E 02	9.59E 01	1.7E 01	3.0E-01	1.79E-01 S		4.52E 01	2.02E 02	1.2E 00	6.3E-01	9.21E-01 B	
9.437E-01	3.123E-01	5.04E-01	6.88E 02	9.31E 01	1.8E 01	2.9E-01	1.78E-01 S		4.11E 01	2.04E 02	1.0E 00	6.4E-01	9.15E-01 B	
9.493E-01	3.030E-01	4.95E-01	6.92E 02	9.01E 01	1.8E 01	2.8E-01	1.77E-01 S		3.69E 01	2.07E 02	9.4E-01	6.5E-01	9.09E-01 B	
9.549E-01	2.930E-01	4.87E-01	6.96E 02	8.72E 01	1.8E 01	2.7E-01	1.77E-01 S		3.28E 01	2.10E 02	8.4E-01	6.6E-01	9.02E-01 B	
9.605E-01	2.829E-01	4.78E-01P	7.00E 02	8.42E 01	1.8E 01	2.6E-01	1.76E-01 S		2.87E 01	2.13E 02	7.3E-01	6.7E-01	8.95E-01 B	
9.662E-01	2.731E-01	4.70E-01	7.04E 02	8.12E 01	1.8E 01	2.5E-01	1.75E-01 S		2.46E 01	2.16E 02	6.3E-01	6.8E-01	8.90E-01 B	
9.718E-01	2.641E-01	4.63E-01	7.08E 02	7.86E 01	1.8E 01	2.5E-01	1.75E-01 S		2.05E 01	2.19E 02	5.2E-01	6.9E-01	8.83E-01 B	
9.775E-01	2.569E-01	4.57E-01	7.13E 02	7.64E 01	1.8E 01	2.4E-01	1.74E-01 S		1.64E 01	2.21E 02	4.2E-01	6.9E-01	8.83E-01 B	
9.831E-01	2.531E-01	4.55E-01	7.17E 02	7.53E 01	1.8E 01	2.4E-01	1.74E-01 S		1.23E 01	2.22E 02	3.1E-01	7.0E-01	8.97E-01 B	
9.887E-01	2.542E-01	4.57E-01	7.21E 02	7.56E 01	1.8E 01	2.4E-01	1.73E-01 S		8.21E 00	2.22E 02	2.1E-01	6.9E-01	9.18E-01 B	
9.944E-01	2.613E-01	4.65E-01	7.25E 02	7.77E 01	1.9E 01	2.4E-01	1.73E-01 S		4.11E 00	2.20E 02	1.0E-01	6.9E-01	9.53E-01 B	
1.000E 00	2.732E-01	4.78E-01	7.29E 02	8.13E 01	1.9E 01	2.5E-01	1.73E-01 S		4.55E-05	2.18E 02	1.1E-01	6.5E-01	1.00E 00 B	

INLET TUBE DIAMETER = 0.1000-INCH
 AVERAGE PRESSURE = 15.0-PSIG

TOTAL GAS SUPERF. VEL. = 10.0000 FT/SEC
 AREA FRACTION FOR TUBE NO. 1 = 0.500

TOTAL LIQUID SUPERF. VEL. = 0.6000 FT/SEC

DISTRIBUTION FACTORS		PRS DROP		G-REYNOLDS		L-REYNOLDS		TUBE #1		HOLDUP		G-REYNOLDS		L-REYNOLDS		TUBE #2		HOLDUP	
GAS IN #1	LIQ IN #1	IPSI/FT						G-VEL-#	L-VEL-#		*	G-VEL-#	L-VEL-#			G-VEL-#	L-VEL-#		*
6.000	4.081E-01	4.43E-01	0.00	4.86E-01	0.0	1.5E-01	1.00E-00	7.29E-02	7.04E-01	1.9E-01	2.2E-01	1.74E-01	1.74E-01	1.9E-01	2.2E-01	1.74E-01	1.74E-01	1.74E-01	1.74E-01
5.63E-03	4.602E-01	4.23E-01	8.21E-00	5.48E-01	1.0E-01	1.7E-01	9.37E-01	7.25E-02	6.42E-01	1.9E-01	2.0E-01	1.73E-01	1.73E-01	1.8E-01	2.0E-01	1.73E-01	1.73E-01	1.73E-01	1.73E-01
1.12E-02	4.701E-01	4.17E-01	1.23E-01	5.52E-01	3.1E-01	1.7E-01	9.08E-01	7.21E-02	6.38E-01	1.8E-01	2.0E-01	1.73E-01	1.73E-01	1.8E-01	2.0E-01	1.73E-01	1.73E-01	1.73E-01	1.73E-01
1.68E-02	4.639E-01	4.20E-01	1.64E-01	5.25E-01	4.2E-01	1.6E-01	9.11E-01	7.13E-02	6.05E-01	1.8E-01	2.1E-01	1.74E-01	1.74E-01	1.8E-01	2.1E-01	1.74E-01	1.74E-01	1.74E-01	1.74E-01
2.25E-02	4.413E-01	4.26E-01	2.03E-01	4.97E-01	5.2E-01	1.5E-01	9.30E-01	7.04E-02	7.17E-01	1.8E-01	2.2E-01	1.75E-01	1.75E-01	1.8E-01	2.2E-01	1.75E-01	1.75E-01	1.75E-01	1.75E-01
2.01E-02	4.180E-01	4.33E-01	2.46E-01	4.73E-01	6.3E-01	1.5E-01	9.38E-01	7.00E-02	7.39E-01	1.8E-01	2.3E-01	1.76E-01	1.76E-01	1.8E-01	2.3E-01	1.76E-01	1.76E-01	1.76E-01	1.76E-01
3.94E-02	3.791E-01	4.40E-01	2.87E-01	4.51E-01	7.3E-01	1.4E-01	9.44E-01	6.96E-02	7.57E-01	1.8E-01	2.4E-01	1.77E-01	1.77E-01	1.8E-01	2.4E-01	1.77E-01	1.77E-01	1.77E-01	1.77E-01
4.20E-02	3.635E-01	4.50E-01	3.69E-01	4.33E-01	8.4E-01	1.4E-01	9.50E-01	6.92E-02	7.73E-01	1.8E-01	2.4E-01	1.77E-01	1.77E-01	1.8E-01	2.4E-01	1.77E-01	1.77E-01	1.77E-01	1.77E-01
5.00E-02	3.503E-01	4.53E-01	4.11E-01	4.10E-01	1.0E-01	1.3E-01	9.55E-01	6.88E-02	7.88E-01	1.8E-01	2.5E-01	1.78E-01	1.78E-01	1.8E-01	2.5E-01	1.78E-01	1.78E-01	1.78E-01	1.78E-01
5.03E-02	3.473E-01	4.59E-01	4.52E-01	3.90E-01	1.3E-01	1.2E-01	9.59E-01	6.84E-02	8.00E-01	1.7E-01	2.5E-01	1.78E-01	1.78E-01	1.8E-01	2.5E-01	1.78E-01	1.78E-01	1.78E-01	1.78E-01
6.15E-02	3.170E-01	4.02E-01	4.93E-01	3.78E-01	1.4E-01	1.1E-01	9.62E-01	6.80E-02	8.12E-01	1.7E-01	2.6E-01	1.79E-01	1.79E-01	1.8E-01	2.6E-01	1.79E-01	1.79E-01	1.79E-01	1.79E-01
7.32E-02	3.087E-01	4.04E-01	5.75E-01	3.57E-01	1.5E-01	1.1E-01	9.67E-01	6.72E-02	8.42E-01	1.7E-01	2.6E-01	1.79E-01	1.79E-01	1.8E-01	2.6E-01	1.79E-01	1.79E-01	1.79E-01	1.79E-01
7.68E-02	3.001E-01	4.06E-01	6.57E-01	3.48E-01	1.6E-01	1.1E-01	9.72E-01	6.67E-02	8.54E-01	1.7E-01	2.6E-01	1.80E-01	1.80E-01	1.8E-01	2.6E-01	1.80E-01	1.80E-01	1.80E-01	1.80E-01
8.44E-02	2.925E-01	4.08E-01	7.39E-01	3.39E-01	1.7E-01	1.1E-01	9.75E-01	6.63E-02	8.59E-01	1.7E-01	2.6E-01	1.81E-01	1.81E-01	1.8E-01	2.6E-01	1.81E-01	1.81E-01	1.81E-01	1.81E-01
9.01E-02	2.850E-01	4.09E-01	8.58E-01	3.31E-01	1.8E-01	1.0E-01	9.75E-01	6.59E-02	8.67E-01	1.7E-01	2.7E-01	1.81E-01	1.81E-01	1.8E-01	2.7E-01	1.81E-01	1.81E-01	1.81E-01	1.81E-01
9.57E-02	2.779E-01	4.71E-01	9.39E-01	3.22E-01	1.9E-01	1.0E-01	9.75E-01	6.55E-02	8.75E-01	1.7E-01	2.7E-01	1.82E-01	1.82E-01	1.8E-01	2.7E-01	1.82E-01	1.82E-01	1.82E-01	1.82E-01
1.01E-01	2.710E-01	4.72E-01	1.00E-01	3.15E-01	2.0E-01	9.8E-02	9.76E-01	6.51E-02	8.75E-01	1.7E-01	2.7E-01	1.82E-01	1.82E-01	1.8E-01	2.7E-01	1.82E-01	1.82E-01	1.82E-01	1.82E-01
1.070E-01	2.644E-01	4.73E-01	1.00E-01	3.15E-01	2.0E-01	9.8E-02	9.76E-01	6.51E-02	8.75E-01	1.7E-01	2.7E-01	1.82E-01	1.82E-01	1.8E-01	2.7E-01	1.82E-01	1.82E-01	1.82E-01	1.82E-01

INLET TUBE DIAMETER = 0.1000-INCH
AVERAGE PRESSURE = 15.0-PSIG

TOTAL GAS SUPERF. VEL. = 10.0000 FT/SEC
AREA FRACTION FOR TUBE NO. 1 = 0.500

TOTAL LIQUID SUPERF. VEL. = 0.2000 FT/SEC

DISTRIBUTION FACTORS			PRS DROP		<-----TUBE #1----->										<-----TUBE #2----->									
GAS IN #1	LIQ IN #1	(PSI/FT)	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*		G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*		G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*		
8.930E-01	7.356E-01	4.73E-01	6.51E 02	8.75E 01	1.7E 01	2.7E-01	1.82E-01	S		7.80E 01	3.15E 01	2.0E 00	9.6E-02	9.76E-01	8									
8.980E-01	7.290E-01	4.72E-01	6.55E 02	8.67E 01	1.7E 01	2.7E-01	1.81E-01	S		7.39E 01	3.22E 01	1.9E 00	1.0E-01	9.75E-01	8									
9.043E-01	7.221E-01	4.71E-01	6.59E 02	8.59E 01	1.7E 01	2.7E-01	1.81E-01	S		6.98E 01	3.31E 01	1.8E 00	1.0E-01	9.73E-01	8									
9.099E-01	7.150E-01	4.69E-01	6.63E 02	8.51E 01	1.7E 01	2.7E-01	1.80E-01	S		6.57E 01	3.39E 01	1.7E 00	1.1E-01	9.72E-01	8									
9.155E-01	7.075E-01	4.68E-01	6.67E 02	8.42E 01	1.7E 01	2.6E-01	1.80E-01	S		6.16E 01	3.48E 01	1.6E 00	1.1E-01	9.70E-01	8									
9.212E-01	6.997E-01	4.66E-01	6.72E 02	8.35E 01	1.7E 01	2.6E-01	1.79E-01	S		5.75E 01	3.57E 01	1.5E 00	1.1E-01	9.67E-01	8									
9.268E-01	6.913E-01	4.64E-01	6.76E 02	8.23E 01	1.7E 01	2.6E-01	1.79E-01	S		5.34E 01	3.67E 01	1.4E 00	1.1E-01	9.65E-01	8									
9.324E-01	6.824E-01	4.62E-01	6.80E 02	8.12E 01	1.7E 01	2.5E-01	1.78E-01	S		4.93E 01	3.78E 01	1.3E 00	1.2E-01	9.62E-01	8									
9.381E-01	6.727E-01	4.59E-01	6.84E 02	8.00E 01	1.7E 01	2.5E-01	1.78E-01	S		4.52E 01	3.90E 01	1.2E 00	1.2E-01	9.59E-01	8									
9.437E-01	6.619E-01	4.57E-01	6.88E 02	7.88E 01	1.8E 01	2.5E-01	1.77E-01	S		4.11E 01	4.02E 01	1.0E 00	1.3E-01	9.55E-01	8									
9.493E-01	6.500E-01	4.55E-01	6.92E 02	7.73E 01	1.8E 01	2.4E-01	1.77E-01	S		3.69E 01	4.16E 01	9.4E-01	1.3E-01	9.50E-01	8									
9.549E-01	6.365E-01	4.50E-01	6.96E 02	7.57E 01	1.8E 01	2.4E-01	1.76E-01	S		3.28E 01	4.33E 01	8.4E-01	1.4E-01	9.44E-01	8									
9.606E-01	6.209E-01	4.45E-01	7.00E 02	7.39E 01	1.8E 01	2.3E-01	1.75E-01	S		2.87E 01	4.51E 01	7.3E-01	1.4E-01	9.38E-01	8									
9.662E-01	6.029E-01	4.40E-01	7.04E 02	7.17E 01	1.8E 01	2.2E-01	1.75E-01	S		2.46E 01	4.73E 01	6.3E-01	1.5E-01	9.30E-01	8									
9.718E-01	5.820E-01	4.33E-01	7.08E 02	6.93E 01	1.8E 01	2.2E-01	1.74E-01	S		2.05E 01	4.97E 01	5.2E-01	1.6E-01	9.21E-01	8									
9.775E-01	5.587E-01	4.26E-01	7.13E 02	6.65E 01	1.8E 01	2.1E-01	1.74E-01	S		1.64E 01	5.25E 01	4.2E-01	1.6E-01	9.11E-01	8									
9.831E-01	5.361E-01	4.20E-01	7.17E 02	6.33E 01	1.8E 01	2.0E-01	1.73E-01	S		1.23E 01	5.52E 01	3.1E-01	1.7E-01	9.04E-01	8									
9.887E-01	5.239E-01	4.17E-01	7.21E 02	6.23E 01	1.8E 01	2.0E-01	1.73E-01	S		8.21E 00	5.67E 01	2.1E-01	1.8E-01	9.03E-01	8									
9.944E-01	5.090E-01	4.23E-01	7.25E 02	6.42E 01	1.9E 01	2.0E-01	1.73E-01	S		4.11E 00	5.48E 01	1.0E-01	1.7E-01	9.37E-01	8									
1.000E 00	5.919E-01	4.43E-01	7.29E 02	7.04E 01	1.9E 01	2.2E-01	1.72E-01	S		4.35E-05	4.86E 01	1.1E-00	1.9E-01	1.00E 00	8									

LET TUBE DIAMETER = 0.1000-INCH TOTAL GAS SUPERF. VEL. = 10.0000 FT/SEC TOTAL LIQUID SUPERF. VEL. = 0.0000 FT/SEC
 VERAGE PRESSURE = 15.0-PSIG AREA FRACTION FOR TUBE NO. 1 = 0.500

DISTRIBUTION FACTORS		PRS DROP	TUBE #1						TUBE #2					
GAS IN #1	LIQ IN #1	(PSI/FT)	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*
4.100E-01	1.000E 00	2.29E-01*	2.99E 02	2.97E 01	7.6E 00	9.3E-02	2.39E-01 S		4.30E 02	7.09E-06	1.1E 01	2.2E-08	2.00E-01 S	
4.146E-01	1.000E 00	2.29E-01*	3.02E 02	2.97E 01	7.7E 00	9.3E-02	2.38E-01 S		4.27E 02	1.77E-05	1.1E 01	5.5E-08	2.01E-01 S	
4.192E-01	9.993E-01	2.30E-01	3.06E 02	2.97E 01	7.8E 00	9.3E-02	2.37E-01 S		4.23E 02	2.16E-02	1.1E 01	6.8E-05	2.01E-01 S	
4.238E-01	9.999E-01	2.29E-01	3.09E 02	2.89E 01	7.9E 00	9.0E-02	2.36E-01 S		4.20E 02	8.95E-01	1.1E 01	2.8E-03	2.02E-01 S	
4.285E-01	9.407E-01	2.29E-01P	3.12E 02	2.80E 01	8.0E 00	8.8E-02	2.34E-01 S		4.17E 02	1.76E 00	1.1E 01	5.5E-03	2.03E-01 S	
4.331E-01	9.117E-01	2.28E-01	3.16E 02	2.71E 01	8.1E 00	8.5E-02	2.33E-01 S		4.13E 02	2.63E 00	1.1E 01	8.2E-03	2.04E-01 S	
4.377E-01	8.828E-01	2.27E-01	3.19E 02	2.63E 01	8.2E 00	8.2E-02	2.32E-01 S		4.10E 02	3.49E 00	1.0E 01	1.1E-02	2.05E-01 S	
4.423E-01	8.540E-01	2.27E-01	3.22E 02	2.54E 01	8.2E 00	8.0E-02	2.31E-01 S		4.07E 02	4.34E 00	1.0E 01	1.4E-02	2.05E-01 S	
4.469E-01	8.253E-01	2.26E-01P	3.26E 02	2.46E 01	8.3E 00	7.7E-02	2.29E-01 S		4.03E 02	5.20E 00	1.0E 01	1.6E-02	2.06E-01 S	
4.515E-01	7.967E-01	2.26E-01	3.29E 02	2.37E 01	8.4E 00	7.4E-02	2.28E-01 S		4.00E 02	6.05E 00	1.0E 01	1.9E-02	2.07E-01 S	
4.562E-01	7.682E-01	2.25E-01	3.33E 02	2.29E 01	8.5E 00	7.2E-02	2.27E-01 S		3.96E 02	6.89E 00	1.0E 01	2.2E-02	2.08E-01 S	
4.608E-01	7.398E-01	2.25E-01	3.36E 02	2.20E 01	8.6E 00	6.9E-02	2.26E-01 S		3.93E 02	7.74E 00	1.0E 01	2.4E-02	2.09E-01 S	
4.654E-01	7.114E-01	2.25E-01	3.39E 02	2.12E 01	8.7E 00	6.6E-02	2.25E-01 S		3.90E 02	8.58E 00	1.0E 01	2.7E-02	2.10E-01 S	
4.700E-01	6.831E-01	2.24E-01	3.43E 02	2.03E 01	8.8E 00	6.4E-02	2.24E-01 S		3.86E 02	9.43E 00	9.9E 00	3.0E-02	2.11E-01 S	
4.746E-01	6.549E-01	2.24E-01	3.46E 02	1.95E 01	8.8E 00	6.1E-02	2.22E-01 S		3.83E 02	1.03E 01	9.8E 00	3.2E-02	2.11E-01 S	
4.792E-01	6.267E-01	2.24E-01	3.49E 02	1.86E 01	8.9E 00	5.8E-02	2.21E-01 S		3.80E 02	1.11E 01	9.7E 00	3.5E-02	2.12E-01 S	
4.838E-01	5.985E-01	2.24E-01P	3.53E 02	1.78E 01	9.0E 00	5.6E-02	2.20E-01 S		3.76E 02	1.19E 01	9.6E 00	3.7E-02	2.13E-01 S	
4.885E-01	5.703E-01	2.23E-01	3.56E 02	1.70E 01	9.1E 00	5.3E-02	2.19E-01 S		3.73E 02	1.28E 01	9.5E 00	4.0E-02	2.14E-01 S	
4.931E-01	5.422E-01	2.23E-01	3.59E 02	1.61E 01	9.2E 00	5.0E-02	2.18E-01 S		3.70E 02	1.36E 01	9.4E 00	4.3E-02	2.15E-01 S	
4.977E-01	5.141E-01	2.23E-01	3.63E 02	1.53E 01	9.3E 00	4.8E-02	2.17E-01 S		3.66E 02	1.45E 01	9.4E 00	4.5E-02	2.16E-01 S	
5.023E-01	4.859E-01	2.23E-01P	3.66E 02	1.45E 01	9.4E 00	4.5E-02	2.16E-01 S		3.63E 02	1.53E 01	9.3E 00	4.8E-02	2.17E-01 S	
5.069E-01	4.578E-01	2.23E-01	3.70E 02	1.36E 01	9.4E 00	4.3E-02	2.15E-01 S		3.59E 02	1.61E 01	9.2E 00	5.0E-02	2.18E-01 S	
5.115E-01	4.297E-01	2.23E-01	3.73E 02	1.28E 01	9.5E 00	4.0E-02	2.14E-01 S		3.56E 02	1.70E 01	9.1E 00	5.3E-02	2.19E-01 S	
5.162E-01	4.015E-01	2.24E-01P	3.76E 02	1.19E 01	9.6E 00	3.7E-02	2.13E-01 S		3.53E 02	1.78E 01	9.0E 00	5.6E-02	2.20E-01 S	
5.208E-01	3.733E-01	2.24E-01	3.80E 02	1.11E 01	9.7E 00	3.5E-02	2.12E-01 S		3.49E 02	1.86E 01	8.9E 00	5.9E-02	2.21E-01 S	
5.254E-01	3.451E-01	2.24E-01P	3.83E 02	1.03E 01	9.8E 00	3.2E-02	2.11E-01 S		3.46E 02	1.95E 01	8.8E 00	6.1E-02	2.22E-01 S	
5.300E-01	3.169E-01	2.24E-01	3.86E 02	9.43E 00	9.9E 00	3.0E-02	2.11E-01 S		3.43E 02	2.03E 01	8.8E 00	6.4E-02	2.24E-01 S	
5.346E-01	2.888E-01	2.25E-01	3.90E 02	8.53E 00	1.0E 01	2.7E-02	2.10E-01 S		3.39E 02	2.12E 01	8.7E 00	6.7E-02	2.25E-01 S	
5.392E-01	2.602E-01	2.25E-01	3.93E 02	7.74E 00	1.0E 01	2.4E-02	2.09E-01 S		3.36E 02	2.20E 01	8.6E 00	6.9E-02	2.26E-01 S	
5.438E-01	2.318E-01	2.25E-01P	3.96E 02	6.89E 00	1.0E 01	2.2E-02	2.08E-01 S		3.33E 02	2.29E 01	8.5E 00	7.2E-02	2.27E-01 S	
5.485E-01	2.033E-01	2.26E-01P	4.00E 02	6.05E 00	1.0E 01	1.9E-02	2.07E-01 S		3.29E 02	2.37E 01	8.4E 00	7.4E-02	2.28E-01 S	
5.531E-01	1.747E-01	2.26E-01	4.03E 02	5.20E 00	1.0E 01	1.6E-02	2.06E-01 S		3.26E 02	2.46E 01	8.3E 00	7.7E-02	2.29E-01 S	
5.577E-01	1.460E-01	2.27E-01	4.07E 02	4.34E 00	1.0E 01	1.4E-02	2.05E-01 S		3.22E 02	2.54E 01	8.2E 00	8.0E-02	2.31E-01 S	
5.623E-01	1.172E-01	2.27E-01	4.10E 02	3.49E 00	1.0E 01	1.1E-02	2.05E-01 S		3.19E 02	2.63E 01	8.2E 00	8.2E-02	2.32E-01 S	
5.669E-01	8.829E-02	2.26E-01	4.13E 02	2.63E 00	1.1E 01	8.2E-03	2.04E-01 S		3.16E 02	2.71E 01	8.1E 00	8.5E-02	2.33E-01 S	
5.715E-01	5.926E-02	2.29E-01	4.17E 02	1.76E 00	1.1E 01	5.5E-03	2.03E-01 S		3.12E 02	2.80E 01	8.0E 00	8.8E-02	2.34E-01 S	
5.762E-01	3.003E-02	2.29E-01B	4.20E 02	8.93E-01	1.1E 01	2.8E-03	2.02E-01 S		3.09E 02	2.89E 01	7.9E 00	9.0E-02	2.36E-01 S	
5.808E-01	7.324E-04	2.30E-01B	4.23E 02	2.13E-02	1.1E 01	6.8E-05	2.01E-01 S		3.06E 02	2.97E 01	7.8E 00	9.3E-02	2.37E-01 S	
5.854E-01	4.883E-05	2.31E-01B	4.27E 02	1.45E-03	1.1E 01	4.5E-06	2.01E-01 S		3.02E 02	2.97E 01	7.7E 00	9.3E-02	2.38E-01 S	
5.900E-01	4.883E-05	2.32E-01B	4.30E 02	1.45E-03	1.1E 01	4.5E-06	2.00E-01 S		2.99E 02	2.97E 01	7.6E 00	9.3E-02	2.39E-01 S	

INLET TUBE DIAMETER = 0.1000-INCH
 AREA FRACTION FOR TUBE NO. 1 = 0.500
 TOTAL GAS SUPERF. VEL. = 50.000 FT/SEC
 TOTAL LIQUID SUPERF. VEL. = 20.000 FT/SEC

LISTRI BUTTIC FACTORS PRS DROP < TUBE #1 > < TUBE #2 >
 GAS IN #1 LIQ IN #1 (PSI/FT) G-REYNOS L-REYNOS G-VEL-# L-VEL-# HOLDUP *
 G-REYNOS L-REYNOS G-VEL-# L-VEL-# HOLDUP *

3.000E-01	1.000E 00	8.24E-01*	1.31E 03	1.19E 02	3.4E 01	3.7E-01	1.31E-01	2.33E 03	4.96E-05	6.0E 01	1.6E-07	1.29E-01	S
3.07E-01	1.001E 00	8.25E-01*	1.32E 03	1.19E 02	3.4E 01	3.7E-01	1.30E-01	2.31E 03	4.98E-05	5.9E 01	1.6E-07	1.28E-01	S
3.14E-01	1.002E 00	8.26E-01	1.33E 03	1.15E 02	3.5E 01	3.6E-01	1.29E-01	2.28E 03	3.96E 00	5.8E 01	1.2E-02	1.27E-01	S
3.21E-01	1.003E 00	8.27E-01	1.34E 03	1.15E 02	3.5E 01	3.6E-01	1.28E-01	2.25E 03	3.96E 00	5.8E 01	1.2E-02	1.26E-01	S
3.28E-01	1.004E 00	8.28E-01	1.35E 03	1.08E 02	3.6E 01	3.4E-01	1.27E-01	2.23E 03	3.96E 00	5.8E 01	1.2E-02	1.25E-01	S
3.35E-01	1.005E 00	8.29E-01	1.36E 03	1.08E 02	3.6E 01	3.4E-01	1.26E-01	2.20E 03	3.96E 00	5.8E 01	1.2E-02	1.24E-01	S
3.42E-01	1.006E 00	8.30E-01	1.37E 03	1.04E 02	3.7E 01	3.3E-01	1.25E-01	2.18E 03	3.96E 00	5.8E 01	1.2E-02	1.23E-01	S
3.49E-01	1.007E 00	8.31E-01	1.38E 03	1.04E 02	3.7E 01	3.3E-01	1.24E-01	2.16E 03	3.96E 00	5.8E 01	1.2E-02	1.22E-01	S
3.56E-01	1.008E 00	8.32E-01	1.39E 03	1.01E 02	3.8E 01	3.1E-01	1.23E-01	2.14E 03	3.96E 00	5.8E 01	1.2E-02	1.21E-01	S
3.63E-01	1.009E 00	8.33E-01	1.40E 03	1.01E 02	3.8E 01	3.1E-01	1.22E-01	2.12E 03	3.96E 00	5.8E 01	1.2E-02	1.20E-01	S
3.70E-01	1.010E 00	8.34E-01	1.41E 03	1.01E 02	3.8E 01	3.1E-01	1.21E-01	2.10E 03	3.96E 00	5.8E 01	1.2E-02	1.19E-01	S
3.77E-01	1.011E 00	8.35E-01	1.42E 03	1.01E 02	3.8E 01	3.1E-01	1.20E-01	2.08E 03	3.96E 00	5.8E 01	1.2E-02	1.18E-01	S
3.84E-01	1.012E 00	8.36E-01	1.43E 03	1.01E 02	3.8E 01	3.1E-01	1.19E-01	2.06E 03	3.96E 00	5.8E 01	1.2E-02	1.17E-01	S
3.91E-01	1.013E 00	8.37E-01	1.44E 03	1.01E 02	3.8E 01	3.1E-01	1.18E-01	2.04E 03	3.96E 00	5.8E 01	1.2E-02	1.16E-01	S
3.98E-01	1.014E 00	8.38E-01	1.45E 03	1.01E 02	3.8E 01	3.1E-01	1.17E-01	2.02E 03	3.96E 00	5.8E 01	1.2E-02	1.15E-01	S
4.05E-01	1.015E 00	8.39E-01	1.46E 03	1.01E 02	3.8E 01	3.1E-01	1.16E-01	2.00E 03	3.96E 00	5.8E 01	1.2E-02	1.14E-01	S
4.12E-01	1.016E 00	8.40E-01	1.47E 03	1.01E 02	3.8E 01	3.1E-01	1.15E-01	1.98E 03	3.96E 00	5.8E 01	1.2E-02	1.13E-01	S
4.19E-01	1.017E 00	8.41E-01	1.48E 03	1.01E 02	3.8E 01	3.1E-01	1.14E-01	1.96E 03	3.96E 00	5.8E 01	1.2E-02	1.12E-01	S
4.26E-01	1.018E 00	8.42E-01	1.49E 03	1.01E 02	3.8E 01	3.1E-01	1.13E-01	1.94E 03	3.96E 00	5.8E 01	1.2E-02	1.11E-01	S
4.33E-01	1.019E 00	8.43E-01	1.50E 03	1.01E 02	3.8E 01	3.1E-01	1.12E-01	1.92E 03	3.96E 00	5.8E 01	1.2E-02	1.10E-01	S
4.40E-01	1.020E 00	8.44E-01	1.51E 03	1.01E 02	3.8E 01	3.1E-01	1.11E-01	1.90E 03	3.96E 00	5.8E 01	1.2E-02	1.09E-01	S
4.47E-01	1.021E 00	8.45E-01	1.52E 03	1.01E 02	3.8E 01	3.1E-01	1.10E-01	1.88E 03	3.96E 00	5.8E 01	1.2E-02	1.08E-01	S
4.54E-01	1.022E 00	8.46E-01	1.53E 03	1.01E 02	3.8E 01	3.1E-01	1.09E-01	1.86E 03	3.96E 00	5.8E 01	1.2E-02	1.07E-01	S
4.61E-01	1.023E 00	8.47E-01	1.54E 03	1.01E 02	3.8E 01	3.1E-01	1.08E-01	1.84E 03	3.96E 00	5.8E 01	1.2E-02	1.06E-01	S
4.68E-01	1.024E 00	8.48E-01	1.55E 03	1.01E 02	3.8E 01	3.1E-01	1.07E-01	1.82E 03	3.96E 00	5.8E 01	1.2E-02	1.05E-01	S
4.75E-01	1.025E 00	8.49E-01	1.56E 03	1.01E 02	3.8E 01	3.1E-01	1.06E-01	1.80E 03	3.96E 00	5.8E 01	1.2E-02	1.04E-01	S
4.82E-01	1.026E 00	8.50E-01	1.57E 03	1.01E 02	3.8E 01	3.1E-01	1.05E-01	1.78E 03	3.96E 00	5.8E 01	1.2E-02	1.03E-01	S
4.89E-01	1.027E 00	8.51E-01	1.58E 03	1.01E 02	3.8E 01	3.1E-01	1.04E-01	1.76E 03	3.96E 00	5.8E 01	1.2E-02	1.02E-01	S
4.96E-01	1.028E 00	8.52E-01	1.59E 03	1.01E 02	3.8E 01	3.1E-01	1.03E-01	1.74E 03	3.96E 00	5.8E 01	1.2E-02	1.01E-01	S
5.03E-01	1.029E 00	8.53E-01	1.60E 03	1.01E 02	3.8E 01	3.1E-01	1.02E-01	1.72E 03	3.96E 00	5.8E 01	1.2E-02	1.00E-01	S
5.10E-01	1.030E 00	8.54E-01	1.61E 03	1.01E 02	3.8E 01	3.1E-01	1.01E-01	1.70E 03	3.96E 00	5.8E 01	1.2E-02	9.99E-02	S
5.17E-01	1.031E 00	8.55E-01	1.62E 03	1.01E 02	3.8E 01	3.1E-01	1.00E-01	1.68E 03	3.96E 00	5.8E 01	1.2E-02	9.98E-02	S
5.24E-01	1.032E 00	8.56E-01	1.63E 03	1.01E 02	3.8E 01	3.1E-01	9.99E-02	1.66E 03	3.96E 00	5.8E 01	1.2E-02	9.97E-02	S
5.31E-01	1.033E 00	8.57E-01	1.64E 03	1.01E 02	3.8E 01	3.1E-01	9.98E-02	1.64E 03	3.96E 00	5.8E 01	1.2E-02	9.96E-02	S
5.38E-01	1.034E 00	8.58E-01	1.65E 03	1.01E 02	3.8E 01	3.1E-01	9.97E-02	1.62E 03	3.96E 00	5.8E 01	1.2E-02	9.95E-02	S
5.45E-01	1.035E 00	8.59E-01	1.66E 03	1.01E 02	3.8E 01	3.1E-01	9.96E-02	1.60E 03	3.96E 00	5.8E 01	1.2E-02	9.94E-02	S
5.52E-01	1.036E 00	8.60E-01	1.67E 03	1.01E 02	3.8E 01	3.1E-01	9.95E-02	1.58E 03	3.96E 00	5.8E 01	1.2E-02	9.93E-02	S
5.59E-01	1.037E 00	8.61E-01	1.68E 03	1.01E 02	3.8E 01	3.1E-01	9.94E-02	1.56E 03	3.96E 00	5.8E 01	1.2E-02	9.92E-02	S
5.66E-01	1.038E 00	8.62E-01	1.69E 03	1.01E 02	3.8E 01	3.1E-01	9.93E-02	1.54E 03	3.96E 00	5.8E 01	1.2E-02	9.91E-02	S
5.73E-01	1.039E 00	8.63E-01	1.70E 03	1.01E 02	3.8E 01	3.1E-01	9.92E-02	1.52E 03	3.96E 00	5.8E 01	1.2E-02	9.90E-02	S
5.80E-01	1.040E 00	8.64E-01	1.71E 03	1.01E 02	3.8E 01	3.1E-01	9.91E-02	1.50E 03	3.96E 00	5.8E 01	1.2E-02	9.89E-02	S
5.87E-01	1.041E 00	8.65E-01	1.72E 03	1.01E 02	3.8E 01	3.1E-01	9.90E-02	1.48E 03	3.96E 00	5.8E 01	1.2E-02	9.88E-02	S
5.94E-01	1.042E 00	8.66E-01	1.73E 03	1.01E 02	3.8E 01	3.1E-01	9.89E-02	1.46E 03	3.96E 00	5.8E 01	1.2E-02	9.87E-02	S
6.01E-01	1.043E 00	8.67E-01	1.74E 03	1.01E 02	3.8E 01	3.1E-01	9.88E-02	1.44E 03	3.96E 00	5.8E 01	1.2E-02	9.86E-02	S
6.08E-01	1.044E 00	8.68E-01	1.75E 03	1.01E 02	3.8E 01	3.1E-01	9.87E-02	1.42E 03	3.96E 00	5.8E 01	1.2E-02	9.85E-02	S
6.15E-01	1.045E 00	8.69E-01	1.76E 03	1.01E 02	3.8E 01	3.1E-01	9.86E-02	1.40E 03	3.96E 00	5.8E 01	1.2E-02	9.84E-02	S
6.22E-01	1.046E 00	8.70E-01	1.77E 03	1.01E 02	3.8E 01	3.1E-01	9.85E-02	1.38E 03	3.96E 00	5.8E 01	1.2E-02	9.83E-02	S
6.29E-01	1.047E 00	8.71E-01	1.78E 03	1.01E 02	3.8E 01	3.1E-01	9.84E-02	1.36E 03	3.96E 00	5.8E 01	1.2E-02	9.82E-02	S
6.36E-01	1.048E 00	8.72E-01	1.79E 03	1.01E 02	3.8E 01	3.1E-01	9.83E-02	1.34E 03	3.96E 00	5.8E 01	1.2E-02	9.81E-02	S
6.43E-01	1.049E 00	8.73E-01	1.80E 03	1.01E 02	3.8E 01	3.1E-01	9.82E-02	1.32E 03	3.96E 00	5.8E 01	1.2E-02	9.80E-02	S
6.50E-01	1.050E 00	8.74E-01	1.81E 03	1.01E 02	3.8E 01	3.1E-01	9.81E-02	1.30E 03	3.96E 00	5.8E 01	1.2E-02	9.79E-02	S
6.57E-01	1.051E 00	8.75E-01	1.82E 03	1.01E 02	3.8E 01	3.1E-01	9.80E-02	1.28E 03	3.96E 00	5.8E 01	1.2E-02	9.78E-02	S
6.64E-01	1.052E 00	8.76E-01	1.83E 03	1.01E 02	3.8E 01	3.1E-01	9.79E-02	1.26E 03	3.96E 00	5.8E 01	1.2E-02	9.77E-02	S
6.71E-01	1.053E 00	8.77E-01	1.84E 03	1.01E 02	3.8E 01	3.1E-01	9.78E-02	1.24E 03	3.96E 00	5.8E 01	1.2E-02	9.76E-02	S
6.78E-01	1.054E 00	8.78E-01	1.85E 03	1.01E 02	3.8E 01	3.1E-01	9.77E-02	1.22E 03	3.96E 00	5.8E 01	1.2E-02	9.75E-02	S
6.85E-01	1.055E 00	8.79E-01	1.86E 03	1.01E 02	3.8E 01	3.1E-01	9.76E-02	1.20E 03	3.96E 00	5.8E 01	1.2E-02	9.74E-02	S
6.92E-01	1.056E 00	8.80E-01	1.87E 03	1.01E 02	3.8E 01	3.1E-01	9.75E-02	1.18E 03	3.96E 00	5.8E 01	1.2E-02	9.73E-02	S
6.99E-01	1.057E 00	8.81E-01	1.88E 03	1.01E 02	3.8E 01	3.1E-01	9.74E-02	1.16E 03	3.96E 00	5.8E 01	1.2E-02	9.72E-02	S
7.06E-01	1.058E 00	8.82E-01	1.89E 03	1.01E 02	3.8E 01	3.1E-01	9.73E-02	1.14E 03	3.96E 00	5.8E 01	1.2E-02	9.71E-02	S
7.13E-01	1.059E 00	8.83E-01	1.90E 03	1.01E 02	3.8E 01	3.1E-01	9.72E-02	1.12E 03	3.96E 00	5.8E 01	1.2E-02	9.70E-02	S
7.20E-01	1.060E 00	8.84E-01	1.91E 03	1.01E 02	3.8E 01	3.1E-01	9.71E-02	1.10E 03	3.96E 00	5.8E 01	1.2E-02	9.69E-02	S
7.27E-01	1.061E 00	8.85E-01	1.92E 03	1.01E 02	3.8E 01E								

INLET TUBE DIAMETER = 0.1000-INCH
 AVERAGE PRESSURE = 15.0-PSIG
 TOTAL GAS SUPERF. VEL. = 20.0000 FT/SEC
 TOTAL LIQUID SUPERF. VEL. = 0.2000 FT/SEC

DISTRIBUTION FACTORS		TUBE #1		TUBE #2		TUBE #3		TUBE #4		TUBE #5		TUBE #6		TUBE #7		TUBE #8		TUBE #9		TUBE #10		TUBE #11		TUBE #12		TUBE #13		TUBE #14		TUBE #15		TUBE #16		TUBE #17		TUBE #18		TUBE #19		TUBE #20		TUBE #21		TUBE #22		TUBE #23		TUBE #24		TUBE #25		TUBE #26		TUBE #27		TUBE #28		TUBE #29		TUBE #30		TUBE #31		TUBE #32		TUBE #33		TUBE #34		TUBE #35		TUBE #36		TUBE #37		TUBE #38		TUBE #39		TUBE #40		TUBE #41		TUBE #42		TUBE #43		TUBE #44		TUBE #45		TUBE #46		TUBE #47		TUBE #48		TUBE #49		TUBE #50		TUBE #51		TUBE #52		TUBE #53		TUBE #54		TUBE #55		TUBE #56		TUBE #57		TUBE #58		TUBE #59		TUBE #60		TUBE #61		TUBE #62		TUBE #63		TUBE #64		TUBE #65		TUBE #66		TUBE #67		TUBE #68		TUBE #69		TUBE #70		TUBE #71		TUBE #72		TUBE #73		TUBE #74		TUBE #75		TUBE #76		TUBE #77		TUBE #78		TUBE #79		TUBE #80		TUBE #81		TUBE #82		TUBE #83		TUBE #84		TUBE #85		TUBE #86		TUBE #87		TUBE #88		TUBE #89		TUBE #90		TUBE #91		TUBE #92		TUBE #93		TUBE #94		TUBE #95		TUBE #96		TUBE #97		TUBE #98		TUBE #99		TUBE #100		TUBE #101		TUBE #102		TUBE #103		TUBE #104		TUBE #105		TUBE #106		TUBE #107		TUBE #108		TUBE #109		TUBE #110		TUBE #111		TUBE #112		TUBE #113		TUBE #114		TUBE #115		TUBE #116		TUBE #117		TUBE #118		TUBE #119		TUBE #120		TUBE #121		TUBE #122		TUBE #123		TUBE #124		TUBE #125		TUBE #126		TUBE #127		TUBE #128		TUBE #129		TUBE #130		TUBE #131		TUBE #132		TUBE #133		TUBE #134		TUBE #135		TUBE #136		TUBE #137		TUBE #138		TUBE #139		TUBE #140		TUBE #141		TUBE #142		TUBE #143		TUBE #144		TUBE #145		TUBE #146		TUBE #147		TUBE #148		TUBE #149		TUBE #150		TUBE #151		TUBE #152		TUBE #153		TUBE #154		TUBE #155		TUBE #156		TUBE #157		TUBE #158		TUBE #159		TUBE #160		TUBE #161		TUBE #162		TUBE #163		TUBE #164		TUBE #165		TUBE #166		TUBE #167		TUBE #168		TUBE #169		TUBE #170		TUBE #171		TUBE #172		TUBE #173		TUBE #174		TUBE #175		TUBE #176		TUBE #177		TUBE #178		TUBE #179		TUBE #180		TUBE #181		TUBE #182		TUBE #183		TUBE #184		TUBE #185		TUBE #186		TUBE #187		TUBE #188		TUBE #189		TUBE #190		TUBE #191		TUBE #192		TUBE #193		TUBE #194		TUBE #195		TUBE #196		TUBE #197		TUBE #198		TUBE #199		TUBE #200		TUBE #201		TUBE #202		TUBE #203		TUBE #204		TUBE #205		TUBE #206		TUBE #207		TUBE #208		TUBE #209		TUBE #210		TUBE #211		TUBE #212		TUBE #213		TUBE #214		TUBE #215		TUBE #216		TUBE #217		TUBE #218		TUBE #219		TUBE #220		TUBE #221		TUBE #222		TUBE #223		TUBE #224		TUBE #225		TUBE #226		TUBE #227		TUBE #228		TUBE #229		TUBE #230		TUBE #231		TUBE #232		TUBE #233		TUBE #234		TUBE #235		TUBE #236		TUBE #237		TUBE #238		TUBE #239		TUBE #240		TUBE #241		TUBE #242		TUBE #243		TUBE #244		TUBE #245		TUBE #246		TUBE #247		TUBE #248		TUBE #249		TUBE #250		TUBE #251		TUBE #252		TUBE #253		TUBE #254		TUBE #255		TUBE #256		TUBE #257		TUBE #258		TUBE #259		TUBE #260		TUBE #261		TUBE #262		TUBE #263		TUBE #264		TUBE #265		TUBE #266		TUBE #267		TUBE #268		TUBE #269		TUBE #270		TUBE #271		TUBE #272		TUBE #273		TUBE #274		TUBE #275		TUBE #276		TUBE #277		TUBE #278		TUBE #279		TUBE #280		TUBE #281		TUBE #282		TUBE #283		TUBE #284		TUBE #285		TUBE #286		TUBE #287		TUBE #288		TUBE #289		TUBE #290		TUBE #291		TUBE #292		TUBE #293		TUBE #294		TUBE #295		TUBE #296		TUBE #297		TUBE #298		TUBE #299		TUBE #300		TUBE #301		TUBE #302		TUBE #303		TUBE #304		TUBE #305		TUBE #306		TUBE #307		TUBE #308		TUBE #309		TUBE #310		TUBE #311		TUBE #312		TUBE #313		TUBE #314		TUBE #315		TUBE #316		TUBE #317		TUBE #318		TUBE #319		TUBE #320		TUBE #321		TUBE #322		TUBE #323		TUBE #324		TUBE #325		TUBE #326		TUBE #327		TUBE #328		TUBE #329		TUBE #330		TUBE #331		TUBE #332		TUBE #333		TUBE #334		TUBE #335		TUBE #336		TUBE #337		TUBE #338		TUBE #339		TUBE #340		TUBE #341		TUBE #342		TUBE #343		TUBE #344		TUBE #345		TUBE #346		TUBE #347		TUBE #348		TUBE #349		TUBE #350		TUBE #351		TUBE #352		TUBE #353		TUBE #354		TUBE #355		TUBE #356		TUBE #357		TUBE #358		TUBE #359		TUBE #360		TUBE #361		TUBE #362		TUBE #363		TUBE #364		TUBE #365		TUBE #366		TUBE #367		TUBE #368		TUBE #369		TUBE #370		TUBE #371		TUBE #372		TUBE #373		TUBE #374		TUBE #375		TUBE #376		TUBE #377		TUBE #378		TUBE #379		TUBE #380		TUBE #381		TUBE #382		TUBE #383		TUBE #384		TUBE #385		TUBE #386		TUBE #387		TUBE #388		TUBE #389		TUBE #390		TUBE #391		TUBE #392		TUBE #393		TUBE #394		TUBE #395		TUBE #396		TUBE #397		TUBE #398		TUBE #399		TUBE #400		TUBE #401		TUBE #402		TUBE #403		TUBE #404		TUBE #405		TUBE #406		TUBE #407		TUBE #408		TUBE #409		TUBE #410		TUBE #411		TUBE #412		TUBE #413		TUBE #414		TUBE #415		TUBE #416		TUBE #417		TUBE #418		TUBE #419		TUBE #420		TUBE #421		TUBE #422		TUBE #423		TUBE #424		TUBE #425	

INLET TUBE DIAMETER = 0.1000-INCH TOTAL GAS SUPERF. VEL. = 5.0000 FT/SEC TOTAL LIQUID SUPERF. VEL. = 0.2000 FT/SEC
 AVERAGE PRESSURE = 15.0-PSIG AREA FRACTION FOR TUBE NO. 1 = 0.500

DISTRIBUTION FACTORS		PKS DROP	<-----TUBE #1----->						<-----TUBE #2----->					
(G IN #1	LIQ IN #1	(PSI/FT)	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*
0.000	7.798E-02	4.35E-01P	0.00	9.28E 00	0.0	2.9E-02	1.00E 00	8	3.64E 02	1.10E 02	9.3E 00	3.4E-01	2.40E-01 S	
0.842E-03	1.300E-01	4.19E-01P	2.49E 00	1.55E 01	3.4E-02	4.8E-02	9.55E-01	8	3.62E 02	1.04E 02	9.2E 00	3.2E-01	2.40E-01 S	
1.308E-02	1.006E-01	4.09E-01P	4.99E 00	1.91E 01	1.3E-01	6.0E-02	9.25E-01	8	3.60E 02	9.99E 01	9.2E 00	3.1E-01	2.40E-01 S	
2.053E-02	1.707E-01	4.05E-01	7.48E 00	2.03E 01	1.9E-01	8.4E-02	9.11E-01	8	3.57E 02	9.87E 01	9.1E 00	3.1E-01	2.40E-01 S	
2.757E-02	1.075E-01	4.05E-01	9.98E 00	1.99E 01	2.5E-01	6.2E-02	9.08E-01	8	3.55E 02	9.91E 01	9.1E 00	3.1E-01	2.41E-01 S	
3.421E-02	1.577E-01	4.07E-01	1.25E 01	1.88E 01	3.2E-01	5.9E-02	9.11E-01	8	3.52E 02	1.00E 02	9.0E 00	3.1E-01	2.42E-01 S	
4.105E-02	1.454E-01	4.10E-01	1.50E 01	1.73E 01	3.8E-01	5.4E-02	9.17E-01	8	3.50E 02	1.02E 02	8.9E 00	3.2E-01	2.43E-01 S	
4.789E-02	1.325E-01	4.13E-01	1.75E 01	1.58E 01	4.5E-01	4.9E-02	9.24E-01	8	3.47E 02	1.03E 02	8.9E 00	3.2E-01	2.44E-01 S	
5.474E-02	1.196E-01	4.15E-01	2.00E 01	1.43E 01	5.1E-01	4.5E-02	9.30E-01	8	3.45E 02	1.05E 02	8.8E 00	3.3E-01	2.45E-01 S	
6.158E-02	1.077E-01	4.18E-01	2.24E 01	1.28E 01	5.7E-01	4.0E-02	9.36E-01	8	3.42E 02	1.06E 02	8.7E 00	3.3E-01	2.47E-01 S	
6.842E-02	9.007E-02	4.20E-01	2.49E 01	1.14E 01	6.4E-01	3.6E-02	9.41E-01	8	3.40E 02	1.08E 02	8.7E 00	3.4E-01	2.48E-01 S	
7.526E-02	8.489E-02	4.22E-01	2.74E 01	1.01E 01	7.0E-01	3.2E-02	9.45E-01	8	3.37E 02	1.09E 02	8.6E 00	3.4E-01	2.49E-01 S	
8.211E-02	7.406E-02	4.24E-01P	2.99E 01	8.81E 00	7.6E-01	2.8E-02	9.50E-01	8	3.35E 02	1.10E 02	8.5E 00	3.4E-01	2.50E-01 S	
8.895E-02	6.349E-02	4.26E-01	3.24E 01	7.56E 00	8.3E-01	2.4E-02	9.53E-01	8	3.32E 02	1.11E 02	8.5E 00	3.5E-01	2.51E-01 S	
9.579E-02	5.309E-02	4.28E-01P	3.49E 01	6.32E 00	8.9E-01	2.0E-02	9.56E-01	8	3.30E 02	1.13E 02	8.4E 00	3.5E-01	2.52E-01 S	
1.026E-01	4.272E-02	4.29E-01P	3.74E 01	5.08E 00	9.6E-01	1.6E-02	9.59E-01	8	3.27E 02	1.14E 02	8.4E 00	3.6E-01	2.53E-01 S	
1.093E-01	3.247E-02	4.31E-01P	3.99E 01	3.86E 00	1.0E 00	1.2E-02	9.62E-01	8	3.25E 02	1.15E 02	8.3E 00	3.6E-01	2.55E-01 S	
1.163E-01	2.202E-02	4.32E-01P	4.24E 01	2.62E 00	1.1E 00	8.2E-03	9.64E-01	8	3.22E 02	1.16E 02	8.2E 00	3.6E-01	2.56E-01 S	
1.232E-01	1.147E-02	4.34E-01P	4.49E 01	1.37E 00	1.1E 00	4.3E-03	9.66E-01	8	3.20E 02	1.18E 02	8.2E 00	3.7E-01	2.57E-01 S	
1.300E-01	7.324E-04	4.35E-01P	4.74E 01	8.72E-02	1.2E 00	2.7E-04	9.68E-01	8	3.17E 02	1.19E 02	8.1E 00	3.7E-01	2.58E-01 S	

INLET TUBE DIAMETER = 3.1000-INCH TOTAL GAS SUPERF. VEL. = 5.0000 FT/SEC TOTAL LIQUID SUPERF. VEL. = 0.2000 FT/SEC
 AVERAGE PRESSURE = 15.0-PSIG AREA FRACTION FOR TUBE NO. 1 = 0.500

DISTRIBUTION FACTORS																			
TUBE #1					TUBE #2					TUBE #3					TUBE #4				
LAS IN #1	LIQ IN #1	PKS GRP	G-REYNCS	L-REYNCS	G-VEL-#	L-VEL-#	HOLDUP	* G-REYNCS	L-REYNCS	G-VEL-#	L-VEL-#	HOLDUP	* G-REYNCS	L-REYNCS	G-VEL-#	L-VEL-#	HOLDUP	* G-REYNCS	L-REYNCS
6.700E-01	9.995E-01	4.35E-01	3.17E-02	1.19E-02	8.10-00	3.7E-01	2.58E-01	4.74E-01	8.77E-02	1.2E-00	2.7E-04	9.68E-01	8	4.74E-01	8.77E-02	1.2E-00	2.7E-04	9.68E-01	8
6.700E-01	9.995E-01	4.34E-01	3.20E-02	1.18E-02	8.2E-00	3.7E-01	2.57E-01	4.44E-01	1.37E-00	1.1E-00	4.3E-03	9.68E-01	8	4.44E-01	1.37E-00	1.1E-00	4.3E-03	9.68E-01	8
6.837E-01	9.760E-01	4.32E-01	3.22E-02	1.16E-02	8.2E-00	3.6E-01	2.56E-01	4.24E-01	1.26E-00	1.1E-00	8.2E-03	9.68E-01	8	4.24E-01	1.26E-00	1.1E-00	8.2E-03	9.68E-01	8
6.903E-01	9.670E-01	4.31E-01	3.25E-02	1.15E-02	8.3E-00	3.6E-01	2.55E-01	3.99E-01	1.36E-00	1.0E-00	1.2E-02	9.68E-01	8	3.99E-01	1.36E-00	1.0E-00	1.2E-02	9.68E-01	8
6.974E-01	9.572E-01	4.29E-01	3.27E-02	1.14E-02	8.4E-00	3.6E-01	2.53E-01	3.74E-01	1.45E-00	9.6E-01	1.6E-02	9.59E-01	8	3.74E-01	1.45E-00	9.6E-01	1.6E-02	9.59E-01	8
6.042E-01	9.465E-01	4.28E-01	3.30E-02	1.13E-02	8.4E-00	3.5E-01	2.52E-01	3.49E-01	1.54E-00	8.5E-01	2.0E-02	9.50E-01	8	3.49E-01	1.54E-00	8.5E-01	2.0E-02	9.50E-01	8
6.11E-01	9.365E-01	4.26E-01	3.32E-02	1.11E-02	8.5E-00	3.5E-01	2.51E-01	3.24E-01	1.63E-00	7.6E-01	2.4E-02	9.41E-01	8	3.24E-01	1.63E-00	7.6E-01	2.4E-02	9.41E-01	8
6.179E-01	9.259E-01	4.24E-01	3.35E-02	1.10E-02	8.5E-00	3.4E-01	2.50E-01	2.99E-01	1.72E-00	7.0E-01	2.8E-02	9.31E-01	8	2.99E-01	1.72E-00	7.0E-01	2.8E-02	9.31E-01	8
6.247E-01	9.151E-01	4.22E-01	3.37E-02	1.09E-02	8.6E-00	3.4E-01	2.49E-01	2.74E-01	1.81E-00	6.4E-01	3.2E-02	9.21E-01	8	2.74E-01	1.81E-00	6.4E-01	3.2E-02	9.21E-01	8
6.310E-01	9.039E-01	4.20E-01	3.40E-02	1.08E-02	8.7E-00	3.4E-01	2.48E-01	2.49E-01	1.90E-00	5.7E-01	3.6E-02	9.11E-01	8	2.49E-01	1.90E-00	5.7E-01	3.6E-02	9.11E-01	8
6.384E-01	8.923E-01	4.18E-01	3.42E-02	1.06E-02	8.7E-00	3.3E-01	2.47E-01	2.24E-01	2.00E-00	5.1E-01	4.0E-02	9.01E-01	8	2.24E-01	2.00E-00	5.1E-01	4.0E-02	9.01E-01	8
6.455E-01	8.802E-01	4.15E-01	3.45E-02	1.05E-02	8.8E-00	3.3E-01	2.45E-01	2.00E-01	2.10E-00	4.5E-01	4.4E-02	8.91E-01	8	2.00E-01	2.10E-00	4.5E-01	4.4E-02	8.91E-01	8
6.524E-01	8.675E-01	4.13E-01	3.47E-02	1.03E-02	8.9E-00	3.2E-01	2.44E-01	1.75E-01	2.20E-00	4.0E-01	4.8E-02	8.81E-01	8	1.75E-01	2.20E-00	4.0E-01	4.8E-02	8.81E-01	8
6.592E-01	8.546E-01	4.10E-01	3.50E-02	1.02E-02	9.0E-00	3.2E-01	2.43E-01	1.50E-01	2.30E-00	3.5E-01	5.2E-02	8.71E-01	8	1.50E-01	2.30E-00	3.5E-01	5.2E-02	8.71E-01	8
6.658E-01	8.425E-01	4.07E-01	3.52E-02	1.00E-02	9.0E-00	3.1E-01	2.42E-01	1.25E-01	2.40E-00	3.0E-01	5.6E-02	8.61E-01	8	1.25E-01	2.40E-00	3.0E-01	5.6E-02	8.61E-01	8
6.726E-01	8.305E-01	4.05E-01	3.55E-02	9.9E-03	9.1E-00	3.1E-01	2.41E-01	9.9E-02	2.50E-00	2.5E-01	6.0E-02	8.51E-01	8	9.9E-02	2.50E-00	2.5E-01	6.0E-02	8.51E-01	8
6.795E-01	8.185E-01	4.03E-01	3.57E-02	9.8E-03	9.1E-00	3.1E-01	2.40E-01	7.48E-02	2.60E-00	1.9E-01	6.4E-02	8.41E-01	8	7.48E-02	2.60E-00	1.9E-01	6.4E-02	8.41E-01	8
6.863E-01	8.065E-01	4.01E-01	3.60E-02	9.7E-03	9.2E-00	3.1E-01	2.39E-01	4.99E-02	2.70E-00	1.3E-01	6.8E-02	8.31E-01	8	4.99E-02	2.70E-00	1.3E-01	6.8E-02	8.31E-01	8
6.932E-01	7.945E-01	4.00E-01	3.62E-02	1.04E-02	9.3E-00	3.2E-01	2.38E-01	2.49E-02	2.80E-00	7.0E-01	7.2E-02	8.21E-01	8	2.49E-02	2.80E-00	7.0E-01	7.2E-02	8.21E-01	8
1.000E-00	5.200E-01	4.35E-01	3.64E-02	1.10E-02	9.3E-00	3.4E-01	2.37E-01	2.17E-02	2.90E-00	5.5E-01	7.6E-02	8.11E-01	8	2.17E-02	2.90E-00	5.5E-01	7.6E-02	8.11E-01	8

INLET TUBE DIAMETER = 0.1000-INCH TOTAL GAS SUPERF. VEL. = 2.0000 FT/SEC TOTAL LIQUID SUPERF. VEL. = 0.2000 FT/SEC
 AVERAGE PRESSURE = 15.0-PSIG AREA FRACTION FOR TUBE NO. 1 = 0.500

DISTRIBUTION FACTORS		PRS DROP (PSI/FT)	<-----TUBE #1----->						*	<-----TUBE #2----->						*
GAS IN #1	LIQ IN #1		G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	G-REYNOS		L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP			
4.700E-01	5.305E-01	5.01E-01	6.85E 01	6.31E 01	1.8E 00	2.0E-01	9.68E-01	B	7.73E 01	5.59E 01	2.0E 00	1.7E-01	9.73E-01	B		
4.732E-01	5.273E-01	5.01E-01	6.90E 01	6.27E 01	1.8E 00	2.0E-01	9.68E-01	B	7.68E 01	5.62E 01	2.0E 00	1.8E-01	9.72E-01	B		
4.763E-01	5.241E-01	5.01E-01	6.94E 01	6.24E 01	1.8E 00	2.0E-01	9.69E-01	B	7.64E 01	5.66E 01	2.0E 00	1.8E-01	9.72E-01	B		
4.795E-01	5.209E-01	5.01E-01	6.99E 01	6.20E 01	1.8E 00	1.9E-01	9.69E-01	B	7.59E 01	5.70E 01	1.9E 00	1.8E-01	9.72E-01	B		
4.826E-01	5.176E-01	5.01E-01	7.04E 01	6.16E 01	1.8E 00	1.9E-01	9.69E-01	B	7.54E 01	5.74E 01	1.9E 00	1.8E-01	9.72E-01	B		
4.858E-01	5.144E-01	5.01E-01	7.08E 01	6.12E 01	1.8E 00	1.9E-01	9.69E-01	B	7.50E 01	5.78E 01	1.9E 00	1.8E-01	9.72E-01	B		
4.889E-01	5.112E-01	5.01E-01	7.13E 01	6.08E 01	1.8E 00	1.9E-01	9.70E-01	B	7.45E 01	5.82E 01	1.9E 00	1.8E-01	9.71E-01	B		
4.921E-01	5.080E-01	5.01E-01	7.17E 01	6.05E 01	1.8E 00	1.9E-01	9.70E-01	B	7.41E 01	5.85E 01	1.9E 00	1.8E-01	9.71E-01	B		
4.953E-01	5.048E-01	5.01E-01	7.22E 01	6.01E 01	1.8E 00	1.9E-01	9.70E-01	B	7.36E 01	5.89E 01	1.9E 00	1.8E-01	9.71E-01	B		
4.984E-01	5.016E-01	5.01E-01	7.27E 01	5.97E 01	1.9E 00	1.9E-01	9.70E-01	B	7.31E 01	5.93E 01	1.9E 00	1.9E-01	9.71E-01	B		
5.016E-01	4.984E-01	5.01E-01	7.31E 01	5.93E 01	1.9E 00	1.9E-01	9.71E-01	B	7.27E 01	5.97E 01	1.9E 00	1.9E-01	9.70E-01	B		
5.047E-01	4.952E-01	5.01E-01	7.36E 01	5.89E 01	1.9E 00	1.8E-01	9.71E-01	B	7.22E 01	6.01E 01	1.8E 00	1.9E-01	9.70E-01	B		
5.079E-01	4.920E-01	5.01E-01	7.41E 01	5.85E 01	1.9E 00	1.8E-01	9.71E-01	B	7.17E 01	6.05E 01	1.8E 00	1.9E-01	9.70E-01	B		
5.111E-01	4.888E-01	5.01E-01	7.45E 01	5.82E 01	1.9E 00	1.8E-01	9.71E-01	B	7.13E 01	6.08E 01	1.8E 00	1.9E-01	9.70E-01	B		
5.142E-01	4.856E-01	5.01E-01	7.50E 01	5.78E 01	1.9E 00	1.8E-01	9.72E-01	B	7.08E 01	6.12E 01	1.8E 00	1.9E-01	9.69E-01	B		
5.174E-01	4.824E-01	5.01E-01	7.54E 01	5.74E 01	1.9E 00	1.8E-01	9.72E-01	B	7.04E 01	6.16E 01	1.8E 00	1.9E-01	9.69E-01	B		
5.205E-01	4.791E-01	5.01E-01	7.59E 01	5.70E 01	1.9E 00	1.8E-01	9.72E-01	B	6.99E 01	6.20E 01	1.8E 00	1.9E-01	9.69E-01	B		
5.237E-01	4.759E-01	5.01E-01	7.64E 01	5.66E 01	2.0E 00	1.8E-01	9.72E-01	B	6.94E 01	6.24E 01	1.8E 00	2.0E-01	9.69E-01	B		
5.268E-01	4.727E-01	5.01E-01	7.68E 01	5.62E 01	2.0E 00	1.8E-01	9.72E-01	B	6.90E 01	6.27E 01	1.8E 00	2.0E-01	9.68E-01	B		
5.300E-01	4.695E-01	5.01E-01	7.73E 01	5.59E 01	2.0E 00	1.7E-01	9.73E-01	B	6.85E 01	6.31E 01	1.8E 00	2.0E-01	9.68E-01	B		

LET TUBE DIAMETER = 0.1000-INCH
 AVERAGE PRESSURE = 15.0-PSIG

TOTAL GAS SUPERF. VEL. = 10.0000 FT/SEC
 AREA FRACTION FOR TUBE NO. 1 = 0.400

TOTAL LIQUID SUPERF. VEL. = 0.2000 FT/SEC

DISTRIBUTION FACTORS		PRS DROP	TUBE #1						TUBE #2					
GAS IN #1	LIQ IN #1	(PSI/FT)	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*	G-REYNOS	L-REYNOS	G-VEL-#	L-VEL-#	HOLDUP	*
0.000	2.389E-01	4.41E-01	0.00	2.54E 01	0.0	8.9E-02	1.00E 00	B	7.99E 02	9.92E 01	1.9E 01	2.8E-01	1.73E-01	S
5.263E-03	2.904E-01	4.22E-01	3.43E 00	3.09E 01	9.8E-02	1.1E-01	9.39E-01	B	7.94E 02	9.25E 01	1.9E 01	2.6E-01	1.73E-01	S
1.053E-02	3.035E-01	4.17E-01	6.86E 00	3.23E 01	2.0E-01	1.1E-01	9.14E-01	B	7.90E 02	9.08E 01	1.8E 01	2.6E-01	1.74E-01	S
1.579E-02	2.910E-01	4.20E-01	1.03E 01	3.10E 01	2.9E-01	1.1E-01	9.12E-01	B	7.86E 02	9.24E 01	1.8E 01	2.6E-01	1.74E-01	S
2.105E-02	2.712E-01	4.25E-01	1.37E 01	2.89E 01	3.9E-01	1.0E-01	9.20E-01	B	7.82E 02	9.50E 01	1.8E 01	2.7E-01	1.75E-01	S
2.632E-02	2.522E-01	4.31E-01	1.72E 01	2.68E 01	4.9E-01	9.4E-02	9.30E-01	B	7.78E 02	9.75E 01	1.8E 01	2.8E-01	1.75E-01	S
3.158E-02	2.356E-01	4.35E-01	2.06E 01	2.51E 01	5.9E-01	8.8E-02	9.58E-01	B	7.73E 02	9.96E 01	1.8E 01	2.8E-01	1.76E-01	S
3.684E-02	2.215E-01	4.39E-01	2.40E 01	2.36E 01	6.9E-01	8.2E-02	9.45E-01	B	7.69E 02	1.01E 02	1.8E 01	2.9E-01	1.76E-01	S
4.211E-02	2.091E-01	4.42E-01	2.75E 01	2.23E 01	7.8E-01	7.8E-02	9.51E-01	B	7.65E 02	1.03E 02	1.8E 01	2.9E-01	1.77E-01	S
4.737E-02	1.982E-01	4.45E-01	3.09E 01	2.11E 01	8.6E-01	7.4E-02	9.56E-01	B	7.61E 02	1.05E 02	1.8E 01	3.0E-01	1.77E-01	S
5.263E-02	1.882E-01	4.48E-01	3.43E 01	2.00E 01	9.6E-01	7.0E-02	9.50E-01	B	7.57E 02	1.06E 02	1.8E 01	3.0E-01	1.78E-01	S
5.789E-02	1.790E-01	4.50E-01	3.77E 01	1.91E 01	1.1E 00	6.7E-02	9.64E-01	B	7.52E 02	1.07E 02	1.8E 01	3.1E-01	1.78E-01	S
6.316E-02	1.704E-01	4.51E-01	4.12E 01	1.81E 01	1.2E 00	6.3E-02	9.67E-01	B	7.48E 02	1.08E 02	1.7E 01	3.1E-01	1.79E-01	S
6.842E-02	1.621E-01	4.53E-01	4.46E 01	1.73E 01	1.3E 00	6.0E-02	9.69E-01	B	7.44E 02	1.09E 02	1.7E 01	3.1E-01	1.79E-01	S
7.368E-02	1.542E-01	4.55E-01	4.80E 01	1.64E 01	1.4E 00	5.7E-02	9.72E-01	B	7.40E 02	1.10E 02	1.7E 01	3.2E-01	1.80E-01	S
7.895E-02	1.465E-01	4.56E-01	5.15E 01	1.56E 01	1.5E 00	5.5E-02	9.74E-01	B	7.36E 02	1.11E 02	1.7E 01	3.2E-01	1.81E-01	S
8.421E-02	1.389E-01	4.58E-01	5.49E 01	1.48E 01	1.6E 00	5.2E-02	9.75E-01	B	7.31E 02	1.12E 02	1.7E 01	3.2E-01	1.81E-01	S
8.947E-02	1.315E-01	4.59E-01	5.83E 01	1.40E 01	1.7E 00	4.9E-02	9.77E-01	B	7.27E 02	1.13E 02	1.7E 01	3.2E-01	1.82E-01	S
9.474E-02	1.242E-01	4.61E-01	6.16E 01	1.32E 01	1.8E 00	4.6E-02	9.78E-01	B	7.23E 02	1.14E 02	1.7E 01	3.3E-01	1.82E-01	S
1.000E-01	1.168E-01	4.62E-01	6.52E 01	1.24E 01	1.9E 00	4.4E-02	9.79E-01	B	7.19E 02	1.15E 02	1.7E 01	3.3E-01	1.83E-01	S

*TUBE DIAMETER = 0.1000-INCH
 AVERAGE PRESSURE = 15.0-PSIG
 AREA FRACTION FOR TUBE NO. 1 = 0.400
 TOTAL GAS SUPERF. VEL. = 10.000 FT/SEC
 TOTAL LIQUID SUPERF. VEL. = 0.2000 FT/SEC

DISTRIBUTION FACTORS PHD DROP TUBE #1 TUBE #2
 GAS IN #1 LIQ IN #1 (PSI/FT) G-REYNOS G-VEL-# L-VEL-# HOLDUP *
 G-REYNOS L-REYNOS G-VEL-# L-VEL-# L-VEL-# HOLDUP *
 <-----> <----->

[illegible]

APPENDIX B

Temperature Profiles without Internal Heat Generation: Preheater Design Considerations

It is desired to predict the temperature history of Synthoil slurry exiting from a packed, tubular preheater of specified design and operating conditions, based on data from two pilot plant runs made on units of smaller diameter, different throughput fluxes, and in one case, of different sized packing. Specifically, it is desired to predict the temperature conditions in an 8-inch preheater 15 ft. long, fed at 1046 pounds of slurry per hour. Generally, the problem is how to predict temperature conditions within a prototype preheater from data obtained in a pilot model.

Data available are insufficient to support direct correlation of the many variables. Therefore, two simplified mathematical models are considered. Both assume the reactor walls to be jacketed and heated to some temperature T_w , and an inlet temperature of T_o . Heat must be transferred first through the preheater walls and boundary layers. It must then be diffused through the packed bed and contents within the preheater. A first (idealized) model assumes wall transfer is limiting; the second that diffusion is limiting. In real preheaters, conditions will probably lie between these two extreme cases.

Wall Transfer Model

In both models the following assumptions are made:

A: Fluid moves in plug flow. This is certainly not highly valid, since there will be substantial axial dispersion through the packed bed.

And the amount of axial dispersion through a given length of bed will be affected by the packing size. Thus, different elements of Synthoil fluid emerging from the preheater will have different time-temperature histories. But at the present level of information, refinement beyond the plug flow approximation does not seem warranted.

B: Heat capacities and transfer coefficients are assumed not to vary between cases. In the mathematical models they will be assumed constant, but a less constrained functional relationship is developed in which they may be functions of temperature. This also does not seem too good an assumption since, for example, wall heat transfer might be expected to vary with mass velocity of fluid past the walls. This assumption will be discussed further in treating specific cases.

C: The reactor has a constant jacket on external wall temperature T_w , and an inlet temperature T_o .

D: There is no heat of reaction.

If a Lagrangian view of the fluid is adopted, it will be apparent that the problem becomes two-dimensional, analogous to the flow of heat into an infinitely long cylinder of some radius a , initial temperature T_o , and wall temperature T_w . That is, a given lamina of fluid sees wall temperature T_w as it passes through the preheater, and temperatures within it will be some function (to be determined) of its residence time θ . And residence time of the lamina is related to its distance from the entrance through the phase velocity v of the liquid phase. That is

$$x = v\theta$$

(B-1)

$$v = \frac{Q}{AF} = \frac{Q}{\pi a^2 \epsilon H} \quad (B-2)$$

Q = liquid flowrate

A = cross-section area of preheater

F = fraction of cross-section through which liquid flows, = ϵH

ϵ = void volume between packing

H = fraction of voids occupied by liquid

a = radius of preheater

In the wall transfer model the lamina at any time will have a radially uniform temperature T. Intuitively it is immediately apparent that if heat transfer through the walls dominate (with fixed values of T_w and T_o , and under the assumptions) the residence time θ needed to reach a specified temperature T will be proportional to the ratio:

$$(\text{lamina volume})/(\text{lamina wall area}).$$

This is demonstrated mathematically by making a heat balance for the lamina:

Heat in through wall = Increase in sensible heat

$$(\text{lamina wall area}) k(T_w - T)d\theta$$

$$= (\text{lamina volume}) \epsilon H \rho C dT$$

$$\frac{d\theta}{dT} = \frac{\epsilon H \rho C (\text{lamina volume})}{k(T_w - T) (\text{lamina wall area})} \quad (B-3)$$

or

$$\frac{d\theta}{dT} = \frac{a \epsilon_H \rho C}{2k(T_w - T)} \quad (B-4)$$

Thus to obtain a given differential temperature rise at dT at any given temperature will take twice as long as $d\theta$ if the radius a of the preheater is twice as large. And if each temperature increment takes twice as long, the integrated time for any specified temperature rise $(T - T_0)$ will be twice as long. (Note that the above requires the parameters such as heat capacity c and overall wall transfer coefficient k , be the same at any given temperature in model and prototype, but not that they be independent of temperature.)

To develop relationships in steady-state temperature profiles, it is noted from (1) and (2)

$$x = \frac{Q \theta}{\pi a^2 \epsilon_H} \quad (B-5)$$

$$d\theta = \frac{\pi a^2 \epsilon_H}{Q} dx \quad (B-6)$$

Substituting (6) in (4)

$$\frac{dx}{dT} = \frac{\rho C Q}{2\pi a k (T_w - T)} \quad (B-7)$$

(conveniently, ϵ_H drops out)

And for two different preheaters, at any specified temperature:

$$\frac{\left(\frac{dx}{dT}\right)_1}{\left(\frac{dx}{dT}\right)_2} = \frac{Q_1 a_2}{Q_2 a_1} \quad (B-8)$$

By the same arguments used for time, the integrated distance $\underline{x}$ needed to raise a flow from T_o to any given temperature $\underline{T}$ will be proportional to the flow and inversely proportional to the radius. And for given liquid fluxes G_L since $G_L = Q/\pi a^2$,

$$\frac{\left(\frac{dx}{dT}\right)_1}{\left(\frac{dx}{dT}\right)_2} = \frac{a_1}{a_2} \quad (B-9)$$

for fixed liquid flux or
phase velocity

In the above, we have developed a model that relates temperature profile in one preheater to that in another. Under the assumptions, $T = T\left(\frac{Qx}{a}\right)$. The form of the function is implicit in equation (7) but thus far we have allowed parameters $\underline{c}$ and $\underline{k}$ to be functions of T . In order to get an implicit solution we now assume that $\underline{c}$ and $\underline{k}$ will be constants, at least within any given preheater operation. Thus, integrating (7), subject to the boundary condition $T = T_o$ at $x = 0$

$$x = \frac{-\rho C Q}{2\pi a k} \ln \left(\frac{T_w - T}{T_w - T_o} \right) \quad (B-10)$$

Letting $\frac{T_w - T}{T_w - T_o} = \tau$; $\frac{\rho C}{\pi k} = B$; $2a = d$

$$x = - \frac{BQ}{d} \ln \tau \quad (B-11)$$

or:

$$\tau = e^{-\frac{dx}{BQ}} \quad (B-11a)$$

Temperature profile data now available to us compare two runs FB-39 and FB-53 in 3" and 1" units, respectively. The 3" unit was packed with 3/4 x 3/4 cylinders, the 1" unit with 1/4" x 1/4". Temperatures measured at various positions in these two preheaters are given by Figure B-1. Four temperatures are shown for each, but the two topmost ones in both cases were constant at 450°C. This, therefore, is taken to represent the wall temperature or T_w . Inlet temperatures are not specified, but were believed to be about 120°C. Thus, there are two significant temperature readings in each profile (ERDA Figures FB-39 and FB-53 are attached). In the 3" preheater of FB-39 there are temperatures of 300°C and 430°C two feet apart. They are used to determine the value of the group $\frac{BQ}{a}$ by taking the distance origin at the 300° reading, as follows. Substituting in equation (12)

$$2 \text{ ft} = - \frac{BQ}{a} \ln \left(\frac{450-430}{450-300} \right)$$

$$\frac{BQ}{a} = 0.993$$

$$B = \frac{0.993 \times 3 \text{ inches}}{25 \text{ lbs/hr}}$$

(mixed units)

Applying these parameters to the 3" preheater, the following profile is predicted:

FB-39

⊙ Thermocouple

Fig. B-1a

FB-53

⊙ Thermocouple

Fig. B-1b

<u>x, ft</u>	<u>Predicted T, °C</u>	<u>Recorded T</u>
0	-674 ^o !!	
0.5	-229 ^o !!	
1	39.2	
1.5	202	
2	300	300
2.5	359	
3	395	
3.5	417	
4.0	430	400
4.5	438	
5.0	443	
5.5	446	
6.0	447	450
6.5	448	
7.0	449	
7.5	449	
8.0	450	450

The model clearly does not look reasonable below 300^o. Using the data developed, it shows a temperature below absolute zero at the inlet. However, the model assumes a constant wall heat transfer coefficient. The viscosity is from two to three orders of magnitude higher between about 220^oC and 310^oC than elsewhere. Over this range heat transfer must in fact be much lower than elsewhere. The value developed for B was derived from a section of the preheater mostly above the high viscosity range. One could not expect it to be valid in the high viscosity range. Thus we might attribute the failure of the model function to variations of k with T. However, the opposite effect appears in the 1" preheater run.

In a completely analogous way a value for B and a corresponding temperature profile can be developed for the 1" preheater. One obtains:

$$\frac{BQ}{a} = 8.73$$

$$B = 0.349$$

A corresponding temperature profile is:

<u>x, ft.</u>	<u>Predicted T, °C</u>	<u>Observed T, °C</u>
0	328	
1	350	350
2	361	
3	371	
5	387	
7	400	
9	410	410
11	418	
13	425	
15	430	
17	435	
18.5	437	450
26.5	445	450

In this case two things are to be noted: First, the value of B derived from this 1" preheater is 0.349, which is 2.8 times as large as the value 0.119 found for the 3" preheater. B is by definition $\rho C / (2\pi k)$. ρ and c are not going to be significantly different in the two different preheaters. And k if anything, ought to be greater in the 1" unit, since Q 's were the same in both cases, and thus liquid mass velocity is nine times as high in the 1" unit. Thus B should, if anything, be lower in the 1" unit, rather than higher.

Second, the predicted inlet temperature is 328°C , which is unreasonably high. Furthermore, it is so in spite of the viscosity effect noted above.

One must conclude that the model is of dubious validity, or else that the data are faulty.

A second way of estimating parameters would be to assume an inlet temperature T_o of 120° , and determine the value of B for each of the significant readings. For the 3" model this gives the values $B = 0.304$ for the 2 ft. point and $B = 0.171$ for the 4 ft. point. The average is 0.238. This changes the 300° point to 330, and the 430° point to 410° . And this value is used to predict the wall transfer model curve of Figure 1 for the 8" preheater.

Diffusion Model

By heat balance for a Lagrangian lamina, as before

$$\frac{dT}{d\theta} = K \left[\frac{d^2T}{dr^2} + \frac{1}{r} \frac{dT}{dr} \right] \quad (\text{B-12})$$

A solution for this is given by Carslaw and Jaeger

$$\frac{T - T_o}{T_w - T_o} = \frac{2}{A} \sum e^{-K\alpha_n^2\theta} \frac{J_o(r\alpha_n)}{\alpha_n J_1(a\alpha_n)} \quad (\text{B-13})$$

α_1, α_2 -- are positive roots of $J_o(\alpha a) = 0$

They also give various numerical solutions in the form of plots. The most useful for our purposes are a plot of average $\frac{T - T_o}{T_w - T_o}$, and one of its value at the centerline ($r = 0$) against the group $K\theta/a^2$. The data are replotted in Figure B-1 as $\ln \tau$ vs. the group $K\theta/a^2$. Note that the plots become linear

at higher values of the group. The linear part of the curve for average τ is represented by the following function,

$$\tau_{av} = 0.666 e^{-5.716 K\theta/a^2} \quad K\theta/a^2 > 0.1 \quad (B-14)$$

And for the midline τ

$$\tau_{mid} = 1.57 e^{-5.716 K\theta/a^2} \quad K\theta/a^2 > 0.1 \quad (B-15)$$

To develop the relationship in steady-state temperature profiles, from (1)

$$\theta = x/v$$

We also assume that thermal diffusivity K will be proportional to phase velocity $\underline{v}$, i.e. $K = bv$. The rationale is that heat will be distributed laterally primarily by conduction. It therefore is transferred by lateral mixing or dispersion of the liquid phase. As a first approximation the rate of radial mixing will be proportional to the liquid phase velocity $\underline{v}$. Substituting in the group $(5.716 K\theta/a^2)$

$$-5.716 K\theta/a^2 = -5.716 bx/a^2 = -\beta x/d^2 \quad (B-16)$$

Thus,

$$\tau_m = \tau_m(-\beta x/d^2) \quad \text{all } \tau \quad (B-17)$$

and, from (17)

$$\tau_m = 1.57 e^{-\beta x/d^2} \quad \tau < 0.8 \quad (B-18)$$

Evaluation of parameter β is done much as in the wall transfer model. (It is assumed the data are for midline temperature.) Use of the two significant points for each run yield the following:

	<u>Dia</u>	<u>Q</u>	<u>β</u>	<u>T_o</u>
Run FB-53	1"	25 lbs/hr	0.1145	379°C
Run FB-39	3"	25 lbs/hr	9.07	-267°C

The discrepancies between β for the two runs is much larger even those for the wall transfer model. The two preheater models had different sized packing and different fluxes, but under the assumptions, one might expect the 1" model to show the greater radial dispersion, and hence the higher β . The difference is in the other direction.

Assuming, as for the wall transfer model, an inlet temperature of 120°, the two points in the 3" model give β 's of 5.59 and 7.32 for the 300° point and the 430° points, respectively. The average is 6.45. Use of 6.45 for β yields 326° for the 300° point, and 421° for the 430° point.

Although the β values for the 3" model are not too different, those for the 1" model are as follows: β for the 1 ft (350°) point = 1.645, for the 9 ft (410°) point, β = 0.285. The average is 0.965. Use of 965 for β yields 253° for the 350° point, and almost 450° for the 410° point -- a very poor fit.

To predict the average temperature profile in Figure B-1 the average β for the 3" data (6.45), is used. For example: From (16)

$$\frac{K\theta}{a^2} = \frac{\beta x}{5.716 d^2} = \frac{6.45 x}{5.716 8^2} = 0.0176 x$$

st 1 ft.

$$\frac{K\theta}{a^2} = 0.0176 \quad \tau_{av} = 0.73 \text{ from Figure B-1} \quad T_{av} = 209^\circ$$

The results are plotted in Figure 1.

The data in hand are so inconsistent with either model that it would be unsafe to draw any conclusions as to what, if either, is more nearly correct. But with reasonable probability the two curves of Figure 1 could be taken as bounds. The diffusion model gives the fastest approach, and this is a consequence of its independence of flow Q . Note that if the β from the 1" test were used, the approach would be much slower. Thus the upper curve probably represents something better than would be expected in practice. In view of this, it may be deemed more feasible to use a number of small diameter tubes in parallel for preheating.

Comparison of Models and Summary

Both models lead (under simplifying assumptions) to the conclusion that the time or distance along the preheater, to reach any specified temperature, will be some characteristic scale-up constant times that needed in pilot operation. Preheater temperature profiles will all be similar in shape, only the time or distance scales being appropriately stretched or contracted for the different cases. In other words, temperature, T , in the reactor is $T(\alpha\theta)$ or $T(\beta x)$, where α and β are parameters depending upon preheater diameter, a , also upon flowrates Q , packing sizes, and other factors; θ is residence time, and x is distance from preheater entrance.

The first or wall transfer model predicts that temperature T will be $\tau \left(\frac{ax}{BQ} \right)$, and under the simplifying assumptions, the following relationship is developed:

$$\frac{T_w - T}{T_w - T_o} = \tau = e^{-\frac{ax}{BQ}} \quad (B-19)$$

Here Q is the slurry flow rate, and B is a parameter equal to $\rho c / (2\pi k)$, where ρ is slurry density, c is its heat capacity, and k is the overall heat transfer coefficient for the wall and its boundary films.

In the second or diffusion-controlled model, there is a radial temperature gradient at all points past the entrance. Temperature next to the wall will be T_w at all x . Interior temperature will be some function of radius r and either detention time θ or distance x . However, it turns out that for temperatures at any specified radial location, or for temperature averaged across a preheater section

$$\tau = \tau \left(\frac{\beta x}{a} \right) \text{ or } \tau \left(\frac{\kappa \theta}{a} \right) \quad (B-20)$$

It also turns out that the temperature profile expected in the diffusion model is independent of flow rate (under the simplifying assumptions). Lateral heat transfer is conceived to be largely convective, and lateral dispersion of fluid should be, at least roughly and as a first approximation, proportional to other velocities in the system -- specifically the liquid phase velocity v . And this makes the spread of heat proportional to distance travelled rather than to time.

The practical difference between the models is that the distance x at which a specified temperature is attained is proportional to preheater radius a in the wall transfer model, liquid flux or flow per unit cross-section area being maintained constant. In the diffusion model it is proportional to a^2 . Thus, if it takes two feet of preheater to attain a temperature of 300°C in a 3" model, it would take $8/3 \times 2$ or 5.33 feet in an 8" preheater by the wall transfer model, and $(8/3)^2 \times 2$ or 14.2 feet by the diffusion model.

The pilot data are not consistent with either model, and are hard to rationalize on any basis. And data from the 3" run FB-39 deviates from the model in the opposite direction from that obtained in the 1" run FB-53. Variations of parameters with temperature that might be assumed to explain divergences in one test, would make the other fall even farther from the respective model used.

The fact that temperature profiles in a given preheater would spread directly with liquid feed rate Q in the wall transfer model, and negligibly in the diffusion model, gives a sensitive way of testing experimentally which, if either, is more nearly valid. But such experimental data are not now available.

In view of the wide spread between predictions of the two models, and apparent lack of internal consistency in the small amount of experimental data that are available, it is dangerous to put much faith in predictions for the 8" prototype. But Figure B-2 gives two curves for temperature vs. time in it, based on averaged data from run FB-39 (3") to evaluate parameters and on the two different models. They should, with reasonable probability, represent upper and lower bounds. Temperatures plotted for the diffusion model are average values.

FIG. 1

APPENDIX C

Effect of Temperature on Holdup

In Appendix B we discussed the temperature profile within the preheater where the temperature of the coal/oil slurry and the hydrogen gas is raised from about 100°C to the reaction temperature 450°C. While several other parameters are temperature-sensitive -- including gas density, gas velocity, and liquid surface tension -- the largest change with temperature occurs in the liquid viscosity. Figure C-1 shows that the liquid viscosity (P&M Report No. 9) varies by a factor of 3200 over the temperature range 100-450°C. Such a gross change in a physical property might be expected to have a large effect on the preheater operating variables, especially liquid holdup and residence time. The objective of this section is to estimate the importance of temperature changes on holdup.

As of this writing, no cold model studies had been conducted to directly observe the effect. Previous emphasis had been on flows within the reactor, where the temperature is maintained nearly constant by a jacket. However, since reactions are also expected to occur in the preheater, where the temperature varies, estimates of liquid residence time and holdup are again needed. In the absence of cold model studies of two-phase flow in packed beds, we relied upon an extension of the Ros correlation for two-phase upflow in pipes.

The sample calculation presented in the rest of this appendix was selected to correspond with conditions found in the preheater for Run FB-39. This preheater is 3-inch ID x 10 ft long, packed with 3/4-inch cylindrical ceramic packing, through which 25 lb/hr of liquid and 1300 SCFH of H₂ is passed. To evaluate the superficial velocities we shall need the

COAL-SOLVENT
VISCOSITY AS A FUNCTION OF TEMPERATURE

Fig. C-1

bed porosity and tortuosity. Clearly neither of these parameters could be directly measured. Instead, we assumed that the porosity of geometrically similar packing at steady state should be a unique dimensionless function of the ratio of packing size to the column diameter. Figure C-2 is a plot of the measured porosity as a function of this ratio for 1/8, 1/4, and 3/4-inch packings used in the 4-inch ID cold model. It is gratifying, but perhaps fortuitous, that this turns out to be a linear relation. Intuitively we expect the porosity to approach a constant as the ratio tends to zero (infinite column without wall effect), which is consistent with this linear relation. On this basis we selected $\epsilon = 0.496$ while $\tau = 1$ (arbitrary).

At reaction conditions (450°C, 4000 psi) we evaluated $\rho_G = 0.575$ lb/ft³ by using the corresponding states method (Hougan, Watson and Ragaty, Chemical Process Principles, 1948, p 591). Since the reduced temperature of H₂ is so high ($T_r = 21.7$), the compressibility was found to equal unity so the ideal-gas equation of state can subsequently be used. Using equations (C-3) and (C-11) from the Second Quarterly Report, we evaluate:

$$V_{SL} = \frac{\tau G_L}{\epsilon \rho_L} = 0.00457 \text{ ft/sec}$$

$$V_{SG} = \frac{\tau G_G}{\epsilon \rho_G} = 0.144 \text{ ft/sec}$$

$$2a_t = \frac{2}{3} \frac{\epsilon}{1-\epsilon} = 0.492 \text{ in}$$

$$N_D = 2a_t \sqrt{\rho_L g / \sigma} = 4.58$$

$$N_{VL} = V_{SL} (\rho_L / g \sigma)^{1/4} = 0.00850$$

Fig. C-2

1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18

$$N_{VG} = V_{SG}(\rho_L/g\sigma)^{1/4} = 0.267$$

$$N_L = \mu_L(g/\rho_L\sigma^3)^{1/4} = 0.00224$$

where $\mu_L = 0.01$ g/cm-sec, $\rho_L = 1$ g/cm³, $g = 981$ cm/sec² and $\sigma = 73$ dyne/cm was used.

According to equation (A-9) of this report, the gas flow rate is low enough so that Bubble Flow (Regime I) occurs. Since N_{VG} becomes even smaller when $T < 450^\circ\text{C}$, Bubble Flow will prevail throughout the pre-heater. Thus the holdup fraction must be evaluated from (A-3a) of this report, where N_S is defined by equation (C-10) of the previous report. The effect of lowering the temperature is to raise ρ_G and change μ_L . Using the ideal gas law (compressibility equals unity), N_{VG} can be corrected for temperature according to:

$$N_{VG} = 0.267(T/723^\circ\text{K}) \quad (\text{C-1})$$

This is the only dimensionless group affected by ρ_G . Liquid viscosity affects only N_L , which can be corrected for viscosity changes according to:

$$N_L = 0.00224(\mu_L/1 \text{ cp}) \quad (\text{C-2})$$

Changes in surface tension were ignored.

Table C-1 numerically gives the effect of temperature on the liquid holdup. Note that the effect is not monotonic. Decreasing from 723°K to 660°K actually increases the holdup due to decreases in V_{SG} at nearly constant μ_L . However, as T approaches 507°K , the viscosity sharply rises

TABLE C-1

Temperature Dependence in Liquid Holdup

T (°K)	N_{VG}	μ^{**} (cp)	N_L	f_1^*	f_2^*	f_3^*	f_4^*	f_3'	N_S	H_L
723	0.267	1.0	0.00224	1.25	0.24	0.83	-18	+4.76	1.59	0.833
660	0.244	1.0	0.00224	1.25	0.24	0.83	-18	+4.76	1.53	0.842
600	0.222	3.2	0.00717	1.25	0.24	1.0	+13	-1.84	1.16	0.810
595	0.220	10.0	0.0224	1.30	0.25	2.0	+34	-5.42	1.04	0.791
588	0.217	32.	0.0717	1.90	0.80	3.1	+53	-8.47	1.51	0.857
586	0.216	100	0.224	2.05	1.03	3.6	+55	-8.41	1.67	0.871
579	0.214	320	0.717	1.35	0.87	3.9	+55	-8.11	0.992	0.787
565	0.209	1000	2.24	0.86	0.68	4.2	+55	-7.81	0.530	0.601
507	0.187	3200	7.17	0.70*	0.58*	4.4*	+55*	-7.61	0.443	0.592
495	0.183	32	0.0717	1.90	0.80	3.1	+53	-8.47	1.63	0.883
448	0.165	10	0.0224	1.30	0.25	2.0	+34	-5.42	1.16	0.856
373	0.138	32	0.0717	1.90	0.80	3.1	+53	-8.47	1.75	0.920
495	0.183	32	0.0717	1.90	0.80	3.1	+53	-8.47	1.63	0.883

* f_1 are functions of only N_L and were read from graphs in Govier and Aziz,
The Flow of Complex Mixtures in Pipes, 1972, p 354.

** Read from curve labelled "P&M Report No. 9" in Figure C-1.

more than three orders of magnitude to 3200 cp. Despite this gross increase in μ_L and N_L , the holdup is only decreased from 0.84 to 0.59. Further decreases in temperature correspond to a gross decrease in viscosity which restores the holdup to its initially large value.

In conclusion, it seems that, except in the narrow zone of the preheater where the viscosity is very large, the liquid holdup can be considered essentially constant and equal to the value observed under similar conditions in the cold model. That is, temperature has only a slight effect on holdup. This conclusion should be checked in the cold model by using a fluid (say, glycerine) with a much larger viscosity than water.

APPENDIX D

Radial Temperature Distribution with Internal Heat Generation: Thermal Reactor Stability

Since the reaction occurring in the Synthoil reactor was exothermic, heat must be removed from the reactor to keep the temperature from rising. This is usually accomplished by placing a cooler fluid in contact with the outer wall of the reactor which can absorb the heat produced by the reaction. The heat is then transferred radially from the center of the reactor to the walls, with the center being hotter than the wall.

As the temperature increases the rate of reaction also increases. Unless this heat can be removed quickly, the heat produced by the reaction will raise the local temperature, which further increases the rate of reaction to produce more heat, which in turn raises the temperature, etc. A thermal runaway may result.

If the reactor diameter is small, the resistance to heat transfer to the wall will also be small so that rapid heat removal is achieved. However, as the diameter is increased the resistance increases and the center of the reactor becomes hotter. Eventually, at some critical diameter, the heat removal cannot keep up with the rate of production of heat and a thermal runaway occurs.

George Quarder of Dow Chemical Co. treated this problem in a letter dated February 9, 1976. He concluded that the reactor will be stable if:

$$\frac{1}{h} + \frac{a}{4k_r} < \frac{2RT^2}{a(-\Delta H_r) R E} \quad (D-1)$$

where h = heat transfer coefficient for reactor wall

a = reactor radius

k_r = radial thermal conductivity

R = gas constant

T = absolute temperature

R = rate of reaction

ΔH_r = heat of reaction

E = reaction's activation energy

Neglecting the wall resistance ($h \sim \infty$), and using $T = 723^\circ K$, $E = 25$ kcal/gmole, $R(-\Delta H_r) = 0.07$ cal/cm³-sec, and $k_r = GC_p d_p / 10 = 0.014$ cal/sec-cm-^oC, Quarderer estimated that the largest stable reactor has a diameter of 6 inches.

The conclusions are based on a substantial number of assumptions and/or approximations, so the critical diameter for stable operation of a reactor cannot be evaluated precisely. But certainly there would be substantial risk that a reactor as large as 6" diameter would be thermally unstable, again assuming his input data are valid.

Quarderer seems to have used a reaction heat of 80 BTU per standard cubic feet of hydrogen gas consumed. Dennis Smith provides a figure of 52 BTU/ft³ of H₂ adsorbed, based on Amoco Report, February 9, 1974. This would change the critical diameter by a factor of $(80/52)^{1/2}$ or about 1-1/4. This doesn't change the picture by that much. And note Quarderer's final "word of caution." Any hot spots, due to maldistribution, would increase the average rate of heat generation (due to the exponential order of the rate equations).

In the following, we make some rather academic comments on the literature cited by Quarderer. They do not in any way invalidate the soundness of his deductions.

First (page 10, Notes), the boundary conditions of Wehner and Wilhelm (Danckwertz boundary conditions), are conventional but not "correct" for dispersion in porous media due to repeated redistribution of flow around packing. (See Gumm, D.J., Trans. Instr. Chem. Engr. 47, T351 (1969).) Thus, $\partial c / \partial z = 0$ at $Z = L$ is not a strictly valid boundary condition. But its use doesn't introduce significant error for most purposes.

Second, $d^2T/dZ^2 \leq 0$ is, strictly speaking, neither a sufficient nor necessary condition for thermal stability unless certain other conditions are imposed. This is readily shown by counterexamples: (1) Assume reactants enter a reactor cooled by some medium at any T_c such that more heat is removed than is produced by the reaction at its entrance temperature T_j . As the reactants move through the reactor they cool, approaching asymptotically some balanced reaction temperature at which heat removed equals heat generated. The temperature gradient dT/dZ is negative, but becomes even less negative; i.e. d^2T/dZ^2 is positive. But the reaction is not (at least in the ordinary sense) unstable. Thus, $d^2T/dZ^2 > 0$ is not, in itself, a complete criterion for instability. (2) Also consider any reactor maintained, however unstably, at constant temperature by adjusting cooling temperature T_c so that the heat removed everywhere equals that generated. Then dT/dZ is everywhere zero (for the moment) and so therefore is d^2T/dZ^2 . But if an upward perturbation in temperature would increase the heat generated faster than it does that removed, the reactor will be unstable. If it increases the heat removed faster, then the reactor will be stable. Thus, $d^2T/dZ^2 = 0$ is not, in itself, a valid criterion of stability.

The latter example gives, we feel, an intuitively clearer picture of reactor instability than the d^2T/dZ^2 criterion. The practical stability question is: Will a reactor maintain itself at some specified T over a finite Z ($dT/dZ = 0$, $d^2T/dZ^2 = 0$), given an appropriate cooling temperature T_c , or will any perturbation in temperature grow? (This is a standard version of stability.) More specifically, will an upward perturbation in temperature cause a further rise in temperature. It will do so if $dH_{\text{generated}}/dT > dH_{\text{removed}}/dT$, both being evaluated at the specified reactor temperature.

In the notation of the Quarderer reports,

$$H_{\text{generated}} = R(-\Delta H_r)$$

$$H_{\text{removed}} = \frac{2h(T - T_c)}{a \left(1 - \frac{ha}{4k_r}\right)}$$

and

$$\frac{dH_{\text{generated}}}{dT} = \frac{(-\Delta H_r) R E}{R_g T^2}$$

$$\frac{dH_{\text{removed}}}{dT} = \frac{2h}{a \left(1 - \frac{ha}{4k_r}\right)}$$

Reactor is stable if:

$$\frac{2h}{a \left(1 - \frac{ha}{4k_r}\right)} > \frac{(-\Delta H_r) R E}{R_g T^2}$$

The Quarderer mathematical procedure is illegitimate in that for stable operation at T over a finite Z, both $\partial T / \partial x$ and $\partial^2 T / \partial x^2$ must be zero, and his equation

$$G C_p \frac{\partial^2 T}{\partial Z^2} = (-\Delta H_r) \frac{\partial T}{\partial T} \cdot \frac{\partial T}{\partial Z} - \frac{2h}{a \left(1 + \frac{ha}{4k_r} \right)} \frac{\partial T}{\partial z}$$

will hold for all values of T. But he drops out G, C_p , and Z in a way that is complementary to his introducing them, so they never have any mathematical effect on the results. In the end he equates $dH_{\text{generated}}/dT$ to dH_{removed}/dT to determine the bound of stability.