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SLURRY FIRED HEATER COLD-FLOW MODELLING

By
S. F. Moujaes

Work Performed Under Contract No. AC05-78OR03054

International Coal Refining Company
Allentown, Pennsylvania

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EXECUTIVE SUMMARY

This report summarizes the experimental and theoretical work leading to the scale-up of the SRC-I Demonstration Plant slurry fired heater. The scale-up involved a theoretical model using empirical relations in the derivation, and employed variables such as flow conditions, liquid viscosity, and slug frequency. Such variables have been shown to affect the heat transfer characteristics of the system. The model assumes that if all other variables remain constant, the heat transfer coefficient can be scaled up proportional to $D^{-2/3}$ (D = inside diameter of the fired heater tube).

All flow conditions, liquid viscosities, and pipe inclinations relevant to the demonstration plant have indicated a slug flow regime in the slurry fired heater. The annular and stratified flow regimes should be avoided to minimize the potential for excessive pipe erosion and to decrease temperature gradients along the pipe cross section leading to coking and thermal stresses, respectively.

Cold-flow studies in 3- and 6.75-in.-inside-diameter (ID) pipes were conducted to determine the effect of scale-up on flow regime, slug frequency, and slug dimensions. The developed model assumes that conduction heat transfer occurs through the liquid film surrounding the gas slug and laminar convective heat transfer to the liquid slug. A weighted average of these two heat transfer mechanisms gives a value for the average pipe heat transfer coefficient.

The cold-flow work showed a decrease in the observed slug frequency between the 3- and 6.75-ID pipes. Data on the ratio of gas to liquid slug length in the 6.75-in. pipe are not yet complete, but are expected to yield generally lower values than those obtained in the 3-in. pipe; this will probably affect the scale-up to demonstration plant conditions.

INTRODUCTION

The design of the coal slurry fired heater is one of the more difficult problems in direct coal liquefaction. The difficulty arises from the constraints placed on the design by the nature of the three-phase flow system (coal/solvent/hydrogen) in the fired heater pipes. The potential for maldistribution in multiphase flow required ICRC to adopt a design that utilizes a single fired heater tube per slurry charge pump. Multiple tube circuits per pump would require a manifolding system with coal slurry control valves and would present flow distribution problems. The useful life of such control valves has been severely limited at coal liquefaction pilot plants due to the erosive nature of the solid coal particles dispersed in the slurry. The perceived penalties for valve maintenance and the potential for fired heater downtime due to valve failure outweighed the additional capital expense of adopting a single fired heater tube circuit per slurry charge pump approach. An 8-in. fired heater tube outside diameter (OD) was then selected in order to decrease the number of circuits from six to four, and thus reduce the capital cost required. The maximum design conditions for the demonstration plant fired heater are 1,050°F outside wall tube temperature and 2,600 psig.

The erosive nature of the coal slurry determined the number of fired heater circuits required for the demonstration plant. Pilot plant experience indicated that unacceptable levels of erosion occurred at superficial slurry velocities above 25 ft/sec. ICRC selected four parallel circuits for the demonstration plant fired heater, which limited the maximum superficial slurry velocity to 18 ft/sec.

It was also necessary to address the coking potential of the coal slurry in the design of the demonstration plant fired heater. Wilsonville and Fort Lewis pilot plant experience has shown fired heater coking to be particularly sensitive to tube wall temperature. Most severe coking has been noted when inside tube wall temperatures exceed 900-950°F. In order to minimize the potential for coke formation in the demonstration plant fired heater, the design inside tube wall temperature has been limited to 850°F.

The complex nature of the three-phase flow system necessitated the selection of a horizontal configuration for the slurry fired heater. As the slurry is heated, the coal reacts and dissolves in the recycle solvent. The slurry exhibits a high-viscosity "gel" phase, which depends on composition, temperature, coal grind, and residence time in the heater. This non-Newtonian three-phase flow with varying viscosity cannot be accommodated by conventional multiphase heat transfer procedures. Thus, ICRC elected to rely on the prudent scale-up of pilot plant data in order to establish design criteria for the SRC-I Demonstration Plant fired heater. Both the Wilsonville and Fort Lewis pilot plants employed horizontal fired heaters. A review of multiphase flow correlations (using the Taitel and Dukler flow map, Froude number vs. h_L/D ; Brown and Thorogood, 1981) indicated the horizontal fired heaters operated in the "intermittent" or "slug" flow regime. Based on these correlations, horizontal configuration scale-up would not affect the fluid flow regime. However, scale-up to a vertical configuration would alter the regime to a combination of slug and annular flow. Thus, in order to limit risk in the demonstration plant design, ICRC retained the horizontal configuration of the pilot plant fired heaters.

The objective of this report is to highlight all information leading to the design of the SRC-I Demonstration Plant fired heater. This information includes experimental data from the Fort Lewis Pilot Plant, a theoretical model developed by ICRC, and cold-flow studies performed by ICRC and Creare, Inc. on two different pipe diameters.

In this report, ICRC attempts to use information obtained from the cold-flow studies on variations in slug frequency, the ratio of slug length to pipe diameter, and flow regime with variations in pipe diameter to confirm the assumptions used in developing the theoretical model. ICRC then formulates an expression that can be used with confidence to scale up available data to demonstration plant conditions.

TECHNICAL DESIGN CONSIDERATIONS

Heat Duty

The determination of the coal slurry fired heater heat duty is complicated by the simultaneous occurrence of chemical reactions and

vaporization of recycle solvent. As the coal dissolves, an increased slurry heat capacity is noted between 500 and 600°F. With continued heating, process solvent vaporizes, further affecting the apparent slurry heat capacity. The amount of process solvent that vaporizes depends upon the solvent boiling point distribution, the gas flow rate, and temperature. In order to provide an accurate estimate of slurry fired heater heat duty, as well as temperature/enthalpy correlations for the determination of heat transfer coefficients, ICRC contracted with Wiltec Research Company to perform calorimetric measurements of slurry heat capacity.

Wiltec produced temperature/enthalpy correlations for various coal/process solvent/gas mixtures in a batch autoclave (Freeman and Wilson, 1982). ICRC then compared the change in enthalpy for the gas and solvent fractions, as determined by the ChemShare[®] computer program, with the total measured enthalpy change. The difference represented the apparent heat capacity of the coal, including any chemical reaction (see Appendix A). Thus, a temperature/enthalpy correlation for the coal was determined and used in conjunction with the temperature/enthalpy data for the solvent and gas to predict the heat duty for a variety of operating conditions at the Fort Lewis Pilot Plant, as well as for demonstration plant design conditions. On this basis, the demonstration plant fired heater heat duty was established as 287 MM Btu/hr for a temperature rise in the heater from 500 to 780°F (Figure 1). It was concluded by Wiltec (Freeman and Wilson, 1982) that the flat portion observed in the first trial of their experiments is due to inadequate mixing conditions. Later runs were performed which eliminated this plateau by improving the mixing characteristics of their apparatus. The slurry heaters are expected to have better mixing of the three-phase system.

Pressure Drop

As stated earlier, investigation of the probable flow regime at the Wilsonville and Fort Lewis pilot plants and the SRC-I Demonstration Plant using multiphase flow correlations suggested that flow is not homogeneous and is in the slug flow regime. Measurement of pressure

drop for slurry/gas systems indicated that two-phase gas/liquid correlations are applicable to the slurry flow, in which the two-phase (coal/recycle solvent) slurry is treated as a single liquid phase. ICRC used Dukler's constant slip correlation to evaluate the two-phase flow pressure drop, since this correlation was found to reasonably predict the observed pressure drop (Appendix B). The assumption for using the turbulent flow relation of f vs. Re stems from the observed experimental fact that by using this correlation and the total mass flow one would get a good fit between the measured and calculated pressure drops if the liquid Re is in the turbulent regime.

In order to predict the pressure drop for the demonstration plant fired heater, a series of tests was performed in the Fort Lewis fired heater at various coal/process solvent/gas flow rates (Thew, 1980). The measured pressure drop data were then analyzed using the Dukler two-phase pressure drop correlation, with viscosity as an unknown variable (see Appendix C). The slurry viscosity was selected to match the observed pressure drop. In this manner, Brown was able to develop an apparent viscosity for the coal slurry under a variety of operating conditions, and could scale the pressure drop up to demonstration plant conditions directly from the Dukler correlation. Figure 2 presents the apparent slurry viscosities determined from the Fort Lewis data, and Table 1 presents the corresponding pressure drops used in the design of the demonstration plant fired heater. The Creare, Inc. (Patel and Sam, 1983) cold-flow experiments using a 400-cP viscosity for the slurry have shown that pressure drop with a 6 ft/sec slurry velocity (demonstration plant design) gives a pressure drop of about 54 psi/1,000 ft, which is within the range of predicted pressure drop values (Table 1). Table 1 shows two ranges of temperatures, i.e., 560-590°F and 670-760°F where the pressure drop/1,000 ft is higher than 60 psi. The first range represents the region of gel formation which involves an increase in apparent viscosity. The second range denotes an increase in vaporization of the solvent which in turn increases the total superficial mixture velocity and hence, pressure drop.

Heat Transfer Coefficients

The heat transfer coefficients for the demonstration plant design were scaled up from the data obtained during the Fort Lewis Pilot Plant tests. The Fort Lewis fired heater was extensively instrumented to provide accurate bulk fluid and tube wall temperatures (see Appendix A). It was arranged in a horizontal rectangular coil configuration with an upward slope of 1° and radiant firing on two sides of the heater coil. To determine the heat transfer coefficients, the bulk and skin temperatures were measured using thermocouples located on each coil. This method allowed reasonable determination of the heat transfer coefficients over narrow bulk fluid temperature ranges.

The design of the Fort Lewis fired heater produced a relatively constant heat flux over the entire length of the heater. This allowed heat transfer coefficients to be calculated in two ways. The first utilized the temperature/enthalpy correlation produced from Wiltec data (Freeman and Wilson, 1982) to determine enthalpy changes over a given length of the fired heater. The second performed an energy balance based on the heater firing rate, and assumed a constant heat flux over the entire fired heater surface. In general, the slurry enthalpy change predicted by the temperature/enthalpy correlation was within $\pm 5\%$ of that determined by the heater energy balance (Appendix A). However, the temperature/enthalpy correlation predicted higher heat transfer coefficients at the heater outlet (Table 2). The more conservative heat transfer coefficients predicted by the assumption of constant heat flux were then scaled up for the demonstration plant design.

Figure 3 presents typical heat transfer coefficients for the coal slurry system as measured at the Fort Lewis Pilot Plant. The slurry temperature corresponds to the distance the slurry has traveled along the fired heater tube. As the coal slurry is heated, the slurry viscosity approaches a maximum between 500 and 600°F, and the heat transfer coefficients approach a minimum at roughly the same slurry temperatures. Over the entire range of the Fort Lewis fired heater data, the slurry Reynolds number would predict laminar flow. However, as demonstrated in Figure 3, the Fort Lewis data were inconsistent with laminar heat transfer behavior with a minimum heat transfer coefficient of more than

50 Btu/hr-ft²-°F. Because of the inconsistency between the Fort Lewis data and conventional multiphase heat transfer correlation predictions, ICRC developed an empirical model to fit the data:

$$H = A (Ml)^a (Mg)^b \quad (1)$$

where H = heat transfer coefficient
 Ml = slurry mass flow rate (lb/hr)
 Mg = gas mass flow rate (lb/hr)
 A,a,b = curve fitting constants

Table 3 presents the values of the curve fitting parameters for the heat transfer data, assuming constant heat flux along the pipe (Appendix C).

Scale-up of these heat transfer coefficients to the demonstration plant fired heater was based on ICRC's development of a slug flow heat transfer model (Appendix C). Figure 4 presents a physical picture of slug flow. Alternating slugs of gas and slurry flow along the pipe at approximately the same velocity. At the viscosities found in coal slurries, the liquid slug would be in laminar flow except for the forward mixing region. A complete gas/liquid slug is composed of the following regions: 1) a gas slug moving with velocity V_G ; 2) a liquid mixing region of length l_m ; 3) a fully developed liquid region of velocity V_s , and 4) a liquid carpet at the bottom moving with a velocity $V_L(t)$. The overall heat transfer coefficient for slug flow can be evaluated as the weighted average of the various flow cells within the slug. The mixing region can be treated as turbulent heat transfer with a characteristic length of pipe diameter divided by liquid slug length. In the gas portion of the slug, the coefficient is based on heat transfer to the thin film of liquid on the pipe walls. The basic assumption used in developing the model is that the gas slug and liquid film around it were symmetrical with respect to the pipe axis.

In developing the scale-up relations using this model, the following assumptions were made:

- ° The axial temperature profile along the fired heater in the demonstration plant is similar to that at the Fort Lewis Pilot Plant.
- ° The ratio of gas volumetric flow to slurry volumetric flow is constant with scale-up.
- ° The slug frequency remains unchanged with scale-up.
- ° The slug length to fired heater diameter remains unchanged with scale-up.
- ° An intermittent or slug flow regime still exists after scale-up.

With these assumptions, the treatment of laminar slug flow heat transfer predicts the following diameter dependence for the heat transfer coefficient, H (Appendix C):

$$H \propto (K_1(D)^{-1/2} + K_2(D)^{-2/3}) \quad (2)$$

where K_1 and K_2 are constants and D is the inside diameter of the fired heater tube. $D^{-2/3}$ represented the most conservative scale-up factor and was used to scale the Fort Lewis data to demonstration plant conditions. Table 4 presents parameters for the empirical model calculation of the heat transfer coefficient for the demonstration plant.

The highest mixture velocity encountered during the Fort Lewis tests was 12 ft/sec, while design of the demonstration plant fired heater was based on 18 ft/sec. The empirical model in equation (1) would predict a velocity dependence for heat transfer coefficients between the 0.9 and 0.4 exponent (by adding exponents a and b) based on slurry bulk temperature; however, development of the slug flow heat transfer model would suggest a velocity dependence of no more than the 0.5 exponent (Brown, 1982). ICRC was reluctant to rely on the empirical model outside the range of velocities over which it was developed. Therefore, the demonstration plant fired heater design was based on the heat transfer coefficients calculated by the empirical model for 12 ft/sec and then multiplied by 1.5. The demonstration plant design heat transfer coefficients are presented in Table 5.

Cold-Flow Studies

Several assumptions were necessary to scale the heat transfer coefficients to demonstration plant conditions. The fluid flow regime was assumed to be the same for the demonstration and pilot plants. Although slug flow was confirmed by tracer studies in the Fort Lewis fired heater, the two-phase correlations used to predict the flow regime for the demonstration plant relied heavily on low-viscosity liquid, low-density gas, and small-pipe-diameter data. In addition, there was uncertainty in the boundaries between the various flow regimes. Finally, slug frequency could not be determined from the Fort Lewis data. ICRC therefore conducted its own small-scale study (i.e., using a 3-in. pipe diameter) and contracted with Creare, Inc. to conduct a full-scale cold-flow experiment using pipe a diameter of 6.75 in. In order to minimize the potential for error in the demonstration plant fired heater design, ICRC designed the slurry fired heater with a 1° upward incline, which promoted slug flow. The demonstration plant slurry viscosities were matched with air, used as the gas, in the 3-in. cold-flow studies; Freon gas was used in the low-pressure cold-flow test at Creare, Inc. (1983) to simulate the gas density in the demonstration plant fired heater.

The cold-flow studies confirmed intermittent or slug flow in the two pipe sizes inclined at 1° . The data were plotted against the Taitel/Dukler flow map; Figures 5-8 show that for the liquid viscosities of interest (i.e., 1-400 cP) most flow conditions clearly exhibit slug flow. Note that with a 1° upward inclination and higher viscosity, the data points are more centered in the region of intermittent flow, as opposed to a 0° inclination and lower viscosity, with which the points move closer to the annular/intermittent transition line. These observations confirm that the demonstration plant flow conditions would exhibit intermittent flow, and that the Taitel/Dukler map can be used with confidence to predict flow regimes within the SRC-I Demonstration Plant slurry fired heater.

Figures 9 and 10 show the measured transition between stratified and intermittent regimes in the 3- and 6.75-in.-ID pipes. Note that the pipe configuration inclined at 1° exhibits a greater tendency to change

from stratified to intermittent flow, at lower liquid velocities, than does the pipe configuration inclined at 0° . All of these data indicate that, as designed, the demonstration plant slurry fired heater will produce the desired slug flow regime.

Once the first assumption of an existing slug flow regime in the heater was satisfied, the second assumption required to scale up the heat transfer coefficient proportional to $D^{-2/3}$ was addressed, i.e., that the ratio of pipe diameter to slug length is constant with scale-up. In order to confirm this assumption, Creare, Inc. conducted their cold-flow experiments in a 6.75-in.-ID pipe. The results from the test performed at ICRC on a 3-in. pipe at ambient conditions (Table 6) show that, in general, an increase in liquid viscosity will result in a decrease in the average total length of the slug, i.e., liquid + gas, and an increase in liquid velocity will decrease total slug length. A representative value for L_T/D (L_T = total slug length) for the 3-in. pipe at demonstration plant flow conditions is about 95.0. Creare, Inc. is currently reducing data for the 6.75-in. pipe study to determine how this ratio is affected by diameter. However, observation of the videotapes indicates that the L_T/D will be smaller for the larger sized pipe.

The third assumption for scale-up that must be addressed is whether slug frequency remains constant with change in diameter. Figures 11-15 show the effect of flow conditions, liquid viscosity, and pipe inclination on variations in slug frequency. In general, slug frequency (ν_s) increases with an increase in liquid viscosity and liquid superficial velocity (V_{LS}). In addition, ν_s increases with an upward inclination from 0 to 1° . Increasing the gas superficial velocity (V_{GS}) seems to have a minimal effect on ν_s . Finally, slug frequency decreases with an increase in pipe diameter.

SCALE-UP OF THE SLURRY FIRED HEATER FOR THE SRC-I DEMONSTRATION PLANT

The theoretical development of the heat transfer coefficient (Brown, 1982) to allow scale-up to demonstration plant size, assuming a constant ratio of volumetric gas flow rate to liquid flow rate, constant slug frequency, constant slug length to pipe diameter on scale-up, and

similar axial temperature profile to that in the Fort Lewis fired heater, resulted in the following relationship:

$$H \propto (K_1 D^{-1/2} + K_2 D^{-2/3}) \quad (2)$$

where H = heat transfer coefficient, and
 K_1 and K_2 are constants.

This result was obtained under the assumption that the liquid film around the gas slug and the liquid slug itself contribute to heat transfer. In order to be conservative, ICRC based scale-up on the second term, i.e., $K_2 D^{-2/3}$ (Brown, 1982).

The first two assumptions can be addressed from the cold-flow studies, in which it was shown that slug frequency decreases with diameter, and it is expected, at least qualitatively, that the slug-length-to-diameter ratio will differ for different diameters. The ratio of volumetric gas flow to liquid flow is given in the design (i.e., 2:1), and hence is not a parameter. The temperature profile cannot be delineated to any extent until data are obtained from the demonstration plant itself.

One way to address the scale-up question in light of the currently available data is to compare the suggested basis for scale-up, i.e., $H \propto D^{-2/3}$, to an expression that contains the effect of u_s for a variety of flow conditions in the 3- and 6.75-in.-ID pipes.

With the developed theoretical model (Brown, 1982), the following conservative relationship is derived:

$$Nu \propto K_3 Pe_{mod}^{1/3} \quad (3)$$

where Nu = Nusselt number - $\frac{HD}{k}$

K_3 = constant

Pe_{mod} = modified Peclet number or

$$Pe_{mod} = \frac{C_p \rho_L u_s D^2}{k} \quad (4)$$

where C_p = specific heat

ρ_L = slurry density

u_s = slug frequency

D = inside diameter of heater pipe

k = slurry thermal conductivity.

Equation (3) becomes

$$\text{and} \quad Nu \propto K_3 u_s^{1/3} D^{2/3} \quad (5)$$

$$H \propto K_4 u_s^{1/3} D^{-1/3} \quad (6)$$

In equation (6), the only assumption is that the gas-to-liquid volumetric flow ratio is constant with scale-up.

Table 7 is a comparison of calculations of different scale-up rationales for the heat transfer coefficients as follows:

$$\frac{H_{6.75 \text{ in.}}}{H_{3 \text{ in.}}} = A \text{ (derived from } H \propto D^{-2/3} \text{)}$$

$$\frac{H_{6.75 \text{ in.}}}{H_{3 \text{ in.}}} = B \text{ (derived from } H \propto u_s^{1/3} D^{-1/3} \text{)}$$

The data used for the calculations in Table 7 are taken from Figures 11-15 for the 3- and 6.75-in. pipes, respectively.

Table 7 covers a range of flow conditions and viscosities that could exist in the fired heater. The comparison between scale-up methods C and B shows reasonably close agreement. The data base for slug frequency versus flow conditions, viscosity, and pipe diameter size is currently limited (i.e., only two pipe diameters have been investigated), which makes it difficult to determine a correlation for slug frequency as a function of pipe inside diameter. From the percentage differences between C and B (column 4, Table 7), it seems that the change in the ratio of slug length to pipe diameter and slug frequency

as a function of pipe diameter balance out in their effect on the scale-up of the heat transfer coefficient. Hence, for convenience and simplicity, it is recommended that the simpler scale-up expression, i.e., $H \propto D^{-2/3}$, be used, since it contains the least number of parameters needed.

CONCLUSIONS

Cold-flow studies conducted by ICRC confirmed that intermittent (slug) flow exists in the SRC-I Demonstration Plant fired heater if the pipe is inclined upward at 1° . The studies also revealed that although slug frequency and the ratio of slug length to pipe diameter vary with pipe diameter, their combined effect on scale-up is not very significant. Hence, the scale-up factor of the heat transfer coefficient, $H \propto D^{-2/3}$, is still valid.

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NOMENCLATURE

- D = pipe diameter
- H = heat transfer coefficient
- A, a, b = curve fitting constants
- M_l = slurry mass flow rate
- M_g = gas mass flow rate
- K_1, K_2, K_3, K_4 = constants
- Nu = Nusselt number
- Pe = Peclet number
- C_p = specific heat at constant pressure
- ρ = density
- u_s = slug frequency
- k = slurry thermal conductivity
- C, B = ratios of heat transfer coefficients for scale-up using different approaches
- V_{LS} = superficial slurry velocity
- V_{GS} = superficial gas velocity
- l_u = total length of slug (i.e., liquid + gas)
- l_s = liquid slug length
- l_f = length of gas slug
- l_m = length of recirculating region behind gas slug
- V_G = velocity of gas slug
- $V_{L(t)}$ = velocity of liquid carpet
- V_s = velocity of liquid slug
- V_t = velocity of gas/liquid interface
- V_{L0} = velocity of developing section of liquid carpet
- $R_f(t)$ = profile of gas slug

Table 1

Predicted Pressure Gradients for SRC-I Demonstration Plant
(Four Fired Heaters)

Temperature range (°F)	Pressure gradient (psi/1,000 ft)
300-350	44
350-400	44
400-450	45
450-500	46
500-530	48
530-560	53
560-590	62
590-620	54
620-650	53
650-670	44
670-700	67
700-760	63

Table 2

Heat Transfer Coefficients for Run SL-44R

Coil section	H_i (Wiltec enthalpies) (Btu/hr °F ft ²)	H_i (constant heat flux) (Btu/hr °F ft ²)
1 - 2	137	158
2 - 3	188	188
3 - 4	181	212
4 - 5	187	166
5 - 6	103	128
6 - 7	109	112
7 - 8	119	121
8 - 9	129	145
9 - 10	199	175
10 - 11	294	231
11 - 12	350	252
12 - 13	456	252

Table 3

Parameters for Empirical Correlation of Ft. Lewis Heat Transfer Data
(Brown, 1982)

Temperature range (°F)	A	a	b
375-410	0.0908	0.597	0.309
410-440	0.4097	0.429	0.325
440-470	1.0131	0.300	0.373
470-500	0.5834	0.353	0.354
500-530	0.3577	0.422	0.304
530-560	0.4424	0.398	0.289
560-590	0.6922	0.336	0.307
590-620	1.6127	0.244	0.323
620-650	6.0415	0.153	0.279
650-670	26.3444	0.042	0.247
670-700	11.7971	0.192	0.174
700-740	5.1969	0.355	0.006

Table 4

Parameters for SRC-I Demonstration Plant Empirical Heat Transfer Correlation

Temperature range (°F)	A	a	b
375-410	0.0023	0.597	0.309
410-440	0.0156	0.429	0.325
440-470	0.0482	0.300	0.373
470-500	0.0253	0.353	0.354
500-530	0.0147	0.422	0.304
530-560	0.0202	0.398	0.289
560-590	0.0357	0.336	0.307
590-620	0.1032	0.244	0.323
620-650	0.5639	0.153	0.279
650-670	3.6702	0.042	0.247
670-700	1.3238	0.192	0.174
700-740	0.4997	0.355	0.066

Table 5

Slurry Fired Heater Heat Transfer Coefficients

Temperature range (°F)	H (outside area) ^a (Btu/hr °F ft ²)	Q/A ^b (Btu/hr ft ²)
450-500	48	11,000
500-530	43	11,000
530-560	39	11,000
560-590	39	11,000
590-620	43	11,000
620-650	54	11,000
650-670	70	11,000
670-700	83	11,000
700-760	87	11,000

^aH = heat transfer coefficient in which the theoretical model assumes constant heat flux along the heater coil.

^bQ = heat input into fired heater in Btu/hr; A = surface area of fired heater in ft² using the outside diameter of the heater.

Table 6

Comparison of Predicted and Observed Liquid Slug Lengths
(McDermott and Ying, 1981)

v_{LS}^a (ft/sec)	v_{GS}^b (ft/sec)	l_u^c (ft)	l_s (obsd) ^d (ft)	l_s (calcd) (ft)
<u>350 cP, 1° Inclination</u>				
1.00	2.02	10.6	3.5	2.5
	-9.4			
3.00	2.06	6.3	3.1	3.5
3.00	8.33	17.3	3.3	2.3
3.00	14.33	29.8	3.6	3.8
5.00	8.55	13.1	3.2	4.1
5.00	15.10	23.6	3.1	3.7
<u>65 cP, 1° Inclination</u>				
1.00	2.01	10.0	2.3	2.5
1.00	14.01	96.4	3.2	3.4
1.00	20.01	151.8	2.4	19.3
3.00	8.27	23.6	3.9	1.1
3.00	14.27	42.6	3.4	7.7

^a v_{LS} = superficial liquid velocity.

^b v_{GS} = superficial gas velocity.

^c l_u = total length of slug (liquid + gas).

^d l_s = liquid slug length.

Table 7

Comparison of Two Scale-Up Approaches

Flow conditions	C ($H \propto D^{-2/3}$)	B ($H \propto u_s^{1/3} D^{-1/3}$)	$\frac{C-B}{C}$ (%)
$V_L = 2.4$ ft/sec $V_{gS} = 12$ ft/sec 1° inclination 400 cP	0.58	0.52	+10
Same as above except 1 cP	0.58	0.62	-6.9
$V_L = 2.4$ ft/sec $V_{gS} = 6$ ft/sec 1° inclination 400 cP	0.58	0.54	+6.9
Same as before except 1 cP	0.58	0.53	+6.9
$V_L = 5.2$ ft/sec $V_g = 7.0$ ft/sec 0° inclination 1 cP	0.58	0.69	-18.0
$V_L = 3.0$ ft/sec $V_g = 6.0$ ft/sec 0° inclination 1 cP	0.58	0.64	-10.0

FIGURE 1
PROCESS DUTY PER HEATER

FIGURE 2
SLURRY VISCOSITY VS. TEMPERATURE

FIGURE 3
CALCULATED COEFFICIENTS SL-60'S

FIGURE 4 SLUG FLOW MODEL

(Brown, 1982)

- l_g - length of gas slug
- l_s - length of liquid slug
- l_u - total length of liquid and gas slug
- l_m - length of recirculating region behind gas slug
- V_G - velocity of gas slug
- $V_L(t)$ - velocity of liquid carpet

- V_s - velocity of liquid slug
- V_I - velocity of gas-liquid interface
- V_{LO} - velocity at developing section of liquid carpet

FIGURE 5
COMPARISON OF FLOW TRANSITION WITH
TAITEL/DUKLER FLOW MAP

(McDermott & Ying 1981)

FIGURE 6
COMPARISON OF FLOW TRANSITION WITH
TAITEL/DUKLER FLOW MAP

(Creare Inc., 1983)

FIGURE 7
COMPARISON OF FLOW TRANSITION WITH
TAITEL/DUKLER FLOW MAP

(Creare Inc., 1983)

FIGURE 8
COMPARISON OF FLOW TRANSITION WITH
TAITEL/DUKLER FLOW MAP

(Creare Inc., 1983)

FIGURE 9
MEASURED TRANSITION BETWEEN SEPARATED
AND INTERMITTENT REGIME IN 1-DEGREE
PIPE INCLINATION

1 - DEGREE INCLINATION

(McDermott & Ying 1981)

FIGURE 10
TRANSITION BETWEEN SLUG AND
STRATIFIED REGIMES

(Creare Inc., 1983)

FIGURE 11
EFFECTS OF INCLINATION AND LIQUID SUPERFICIAL
VELOCITY ON MEASURED SLUG FREQUENCIES

(McDermott & Ying 1981)

FIGURE 12
EFFECT OF LIQUID VISCOSITY ON MEASURED
SLUG FREQUENCIES
(1-DEGREE PIPE INCLINATION)

(McDermott & Ying 1981)

FIGURE 13: EFFECT OF LIQUID AND GAS VELOCITY ON SLUG FREQUENCY

(Creare Inc., 1983)

FIGURE 14: EFFECT OF LIQUID AND GAS VELOCITY ON SLUG FREQUENCY

(Creare INC., 1983)

**FIGURE 15: EFFECT OF LIQUID AND GAS
VELOCITY ON SLUG FREQUENCY**

(Creare Inc., 1983)

APPENDIX A
Analysis of Wilco Coal Enthalpy-Temperature Data

International Coal Refining Company

Date 23 May 1980

INTEROFFICE
MEMORANDUM

Subject ANALYSIS OF WILCO COAL ENTHALPY-
TEMPERATURE DATA

To R. M. Thorogood

(ORGANIZATION OR DEPARTMENT)

From B. O. Brown

(ORGANIZATION OR DEPARTMENT)

cc: J. M. Geist
E. M. Phillips
S. M. Morris
F. P. Stein
C. L. Yeh

Summary

Six temperature-enthalpy runs on hydrogen-coal slurry mixtures were carried out using a rocked calorimeter by the Wilco Research Company. Runs were made using the mixture components at weight ratios and heating rates similar to the SRC-I Process. The coal slurry mixtures were composed of 200 mesh Pyro Kentucky #9 coal and Wilsonville process solvent in the presence of H₂ gas.

Analysis of these data produced a base curve of coal enthalpy vs. temperature with a correction term for the endotherm of coal dissolution as a function of heating rate.

The Slurry Preheater enthalpy temperature relation has been calculated from the correlation and this predicts a net duty of 251 MMBTU/Hr. Values of 240, 270 and 318 MMBTU/Hr. were also found if half, two or three times the design gas flow rates are used respectively.

The attached outlines the analysis procedure and presents pertinent process parameters.

B. O. Brown

B. O. Brown

BOB/plb
Attachments

METHOD OF ANALYSIS

Procedure:

Six temperature-enthalpy runs on hydrogen-coal slurry mixtures were carried out by the Wilco Research Company via a rocked calorimeter (Figure 1). All runs were made using the mixture components at weight ratios similar to the SRC-I Process, except for one made on a coal free basis.

Using only the recycle solvent and gas, a series of flash calculations were undertaken for each run at the temperatures and pressures reported by Wilco. This was achieved by use of the Flash Module of the CYCSYN Program, which predicts enthalpies, flow rates and other important phase parameters. The enthalpies of the gas and solvent fractions were then subtracted from the enthalpy of the three-phase mixture (given by Wilco) and the remainder determined as the apparent enthalpy of the coal.

Results and Discussions:

Data received from Wilco are shown in Tables 1-6 and Figures 2-4. Tables 7-11 and Figure 5 present the derived values for the coal enthalpy as a function of temperature for each run.

Each curve (Figure 5) shows an increase in coal enthalpy with temperature. At least one positive deviation from a smooth curve (for each run) is noted. This occurs in the region of 550-625°F and peaks at about 600°F. Higher heating rates tend to give larger deviations.

Previous kinetic studies have indicated that the coal, in the presence of hydrogen and process solvent, follows the following reaction path:

Coal — Dissolved Coal — Preasphaltenes — Asphaltenes — Oils

The first step (solvation) is endothermic, the others (hydrogenation) are exothermic in nature. It is also known that the latter steps become significant at temperatures above 750°F and residence times of 1/2 - 2 hours.

It is postulated that the positive enthalpy deviation observed in each plot is due to the initial endotherm (coal solvation) followed by an exotherm (hydrogenation of the dissolved coal). Some support for this is obtained by examination of the pressure-time behavior of the calorimeter (Figure 4). A reduction of the pressure-time slope occurs in all coal slurry runs at about 70% of the run intervals corresponding to about 600°F, which matches the peak position of the major enthalpy deviations. This may reflect an increase in hydrogen consumption. Run 3 (no coal) does not show a similar effect. Other apparent peaks are not similarly confirmed nor are they reproduced in repeated runs. These may be due to experimental error.

Since the magnitude of the enthalpy peak decreases with a reduced heating rate (consistent with the above reaction model), formulation of a base enthalpy-temperature curve by elimination of the peak data points corresponds to an infinite run time (zero heating rate). Such a plot is shown in Figures 6 and 7. Correlation of the peak enthalpy deviations with heating rate (Figure 8) shows a linear behavior. Extrapolation to higher rates will be incorrect since at zero residence times no coal solvation would occur and hence no peak in enthalpy. It may be possible to predict the relationship from kinetic rate constants from coal dissolution (conversion) and hydrogenation reactions, but this has not been attempted in this present study.

Application and Results:

An estimation of the net duty of the Slurry Preheater for the Demonstration Plant was calculated using the following procedure:

1. With present design flow rates, and estimates of pressure and temperature profiles for the fired heater, a series of flash calculations using the CYCSYN Program Flash Module, were carried out.
2. From enthalpy, flowrates and other phase properties predicted by the flash module, the duty due to gas and solvent can be obtained.
3. Coal enthalpies are then calculated by use of Figure 6. Enthalpy peak values are incorporated by using Figure 7 with the appropriate slurry residence time (8 minutes) and extrapolating linearly from 600°F to a zero increment at 550°F and 625°F.
4. Addition of the values obtained in steps 2 and 3 gives the net duty of the heater.

Values obtained for the heater duty as a function of temperature are presented in Table 12. Shown also are those for a half, two and three times the initial gas flow rate of the design case. That for the design case is plotted in Figure 9.

The net duty calculated for the design rates was 251 MMBTU/Hr. For 1/2, two and three times initial gas rate the values were 240, 270 and 318 respectively.

For the sake of completeness, pertinent parameters generated in this application are also included. These can be listed as follows:

Table 13 - Percent Vaporization Vs. Temperature (Design Case)
Table 14 - Vapor and Liquid Flow Rate Vs. Temperature for
1/2, Design, 2 and 3 Times Design Gas Flow Rates

Figure 10 - Slurry, and Solvent Density Vs. Temperature
Figure 11 - Gas Density Vs. Temperature

Figure 1, Schematic of Rocked Calorimeter Assembly for H₂-Slurry Temperature-Enthalpy Measurements

Figure 2, Enthalpy Versus Temperature, Comparison of H₂-Slurry Runs (Curves Shifted to Arbitrary Basis)

Figure 3, Enthalpy Versus Temperature Without Coal, Run no. 3

Figure 4 , Pressure Versus Percent of Run-Time-Interval,
Comparison of All Six Runs

Figure 5
DERIVED COAL ENTHALPY VS. TEMPERATURE

Figure 6
COAL ENTHALPY BTU/LBM BASE CURVE VS. TEMPERATURE

- - Run 1
- △ - Run 2
- - Run 5
- X - Run 6

Figure 7

DESIGN BASE CURVE (COAL ENTHALPY VS. TEMPERATURE)
KENTUCKY 9 (PYRO) IN WILSONVILLE SOLVENT

FIGURE 8

(MAX. POINT-BASE CASE) VS. RESIDENCE TIME

MAX. Point
- BASE
curve

K-E SEMI-LOGARITHMIC 46 5493
3 CYCLES X 70 DIVISIONS MADE IN U.S.A.
KEUFFEL & ESSER CO.

Figure 9

TOTAL STREAM ENTHALPY VS. TEMPERATURE

Coal = 500,000 lb/hr (Ky #9 Pyro)
 Solvent = 813,016 lb/hr
 Gas = 35,586 lb/hr

Figure 10 - Slurry and Solvent Densities

Figure 11 - Gas Density Vs. Temperature

Density
(lb/ft³)

K-E 10 X 10 TO THE INCH 46 0703
7 X 10 INCHES
MADE IN U.S.A.
KUEPPEL & EMMER CO.

TABLE 1
FULL CELL CHARGE AT SLOW HEATING RATE,
ENTHALPY OF H₂ - COAL SLURRY RUN No. 1

Grams in 950cc cell	<u>H₂</u> 12.66	<u>Solvent</u> 456.1	<u>Coal</u> 280.0	<u>Total</u> 737.8
<u>Temp.</u> °C	<u>Total Time</u> <u>Seconds</u>	<u>Enthalpy</u> <u>cal/gram</u>	<u>Pressure</u> <u>psia</u>	
35	0	2.4	920	
50	69	1.7	960	
65	163	4.4	1040	
80	277	9.8	1110	
95	397	15.9	1175	
110	518	21.9	1235	
125	643	28.4	1300	
140	771	35.0	1365	
155	902	42.0	1430	
170	1037	49.3	1495	
185	1175	56.9	1570	
200	1314	64.5	1635	
215	1455	72.0	1710	
230	1592	79.1	1790	
245	1735	86.7	1875	
260	1880	94.3	1945	
275	2017	100.7	2005	
282.2	2074	102.6	2020	
290	2153	106.9	2050	
297.2	2232	111.5	2080	
305	2309	115.4	2120	
312.2	2434	125.9	2190	
320	2497	127.9	2210	
327.2	2533	126.8	2220	
335	2630	133.0	2210	
342.2	2720	138.8	2195	
350	2808	143.9	2190	
357.2	2894	149.1	2200	
365	2980	153.8	2210	
372.2	3061	158.2	2230	
380	3139	161.9	2230	
387.2	3234	168.0	2220	
398.9	3373	176.1	2190	
412.8	3538	185.5	2155	

TABLE 2
FULL CELL CHARGE AT FAST HEATING RATE,
ENTHALPY OF H₂ - COAL SLURRY RUN No. 2

Grams in 933cc cell	<u>H₂</u>	<u>Solvent</u>	<u>Coal</u>	<u>Total</u>
	1.66	456.1	280.2	738.0

<u>Temp.</u> °C	<u>Total Time</u> <u>Seconds</u>	<u>Enthalpy</u> <u>cal/gram</u>	<u>Pressure</u> <u>psia</u>
35	0	1.7	1070
80	71	1.4	1210
95	108	5.6	1270
110	148	10.6	1330
125	192	16.9	1380
140	239	24.1	1450
155	287	31.6	1510
170	328	36.5	1590
185	388	47.7	1650
200	439	55.8	1730
215	489	63.4	1820
230	542	71.9	1910
245	621	88.7	2060
260	672	96.2	2160
275	711	99.6	2220
290	772	109.9	2360
305	834	120.2	2550
320	922	138.5	2580
335	951	138.3	2580
350	1062	163.1	2670
365	1087	161.5	2690
380	1112	159.8	2710
398.9	1177	167.9	2730
412.8	1252	182.1	2775

TABLE 3
WITHOUT COAL BUT WITH FULL CELL CHARGE AT FAST HEATING RATE,
ENTHALPY OF H₂ - SOLVENT RUN No. 3

Grams in 933cc cell	<u>H₂</u>	<u>Solvent</u>	<u>Coal</u>	<u>Total</u>
	1.72	642.0	0.0	643.7
<u>Temp.</u> <u>°C</u>	<u>Total Time</u> <u>Seconds</u>	<u>Enthalpy</u> <u>cal/gram</u>	<u>Pressure</u> <u>psia</u>	
35	0	3.4	1050	
50	24	2.5	1110	
65	55	4.5	1170	
80	93	8.8	1220	
95	136	15.1	1290	
110	183	23.0	1320	
125	230	30.9	1390	
140	276	38.5	1450	
155	323	46.5	1500	
170	371	54.8	1570	
185	420	63.6	1630	
200	469	72.4	1680	
215	518	81.2	1720	
230	568	90.4	1810	
245	619	100.0	1870	
260	672	110.3	1940	
275	725	120.7	2010	
305	831	141.5	2140	
320	888	153.5	2200	
335	945	165.4	2280	
350	1001	177.0	2350	
365	1057	188.6	2430	
380	1111	199.4	2490 ^{a)}	
398.9	1186	215.6	2810 ^{a)}	
412.8	1234	224.8	3090 ^{a)}	

a) The cell was probably hydrostatic with liquid at these last two temperatures.

TABLE 4
REDUCED CELL CHARGE AT FAST HEATING RATE,
ENTHALPY OF H₂ - COAL SLURRY RUN No. 4

<u>Grams in</u> 930cc cell	<u>H₂</u>	<u>Solvent</u>	<u>Coal</u>	<u>Total</u>
	1.27	428.3	261.6	691.2

<u>Temp.</u> °C	<u>Total Time</u> <u>Seconds</u>	<u>Enthalpy</u> <u>cal/gram</u>	<u>Pressure</u> <u>psia</u>
35	0	3.2	790
50	22	2.1	820
65	47	1.9	850
80	80	4.2	890
95	119	8.5	920
110	164	14.1	970
125	212	20.8	1000
140	262	28.1	1050
155	314	36.4	1110
170	367	45.1	1150
185	419	53.3	1210
200	472	61.6	1260
215	524	69.6	1330
230	578	78.2	1410
245	632	86.6	1490
260	677	92.2	1550
275	735	102.0	1620
290	800	114.2	1690
305	876	130.2	1850
320	921	135.5	1910
335	949	135.3	1920
350	978	135.4	1920
365	1033	144.0	1920
380	1086	152.0	1970
398.9	1158	163.8	2040
412.8	1219	175.1	2100

TABLE 5
REDUCED CELL CHARGE AT SLOW HEATING RATE,
ENTHALPY OF H₂ - COAL SLURRY RUN No. 5

Grams in 930cc cell	<u>H₂</u>	<u>Solvent</u>	<u>Coal</u>	<u>Total</u>
	1.34	426.7	262.0	690.0

<u>Temp.</u> <u>°C</u>	<u>Total Time</u> <u>Seconds</u>	<u>Enthalpy</u> <u>cal/gram</u>	<u>Pressure</u> <u>psia</u>
35	0	4.0	770
50	59	3.7	800
65	135	6.1	830
80	236	12.2	860
95	340	19.0	900
110	450	26.5	940
125	557	33.5	985
140	664	40.6	1030
155	775	48.3	1080
170	888	56.1	1130
185	1001	63.7	1190
200	1117	71.8	1250
215	1234	79.8	1310
230	1351	87.6	1380
245	1464	94.6	1450
260	1591	104.3	1510
275	1721	114.0	1560
282.2	1778	117.8	1585
290	1877	127.4	1630
297.2	1925	129.8	1650
305	1952	128.7	1670
312.2	2015	133.2	1710
320	2098	140.3	1740
327.2	2132	140.4	1750
335	2183	142.7	1750
342.2	2244	146.9	1755
350	2330	154.3	1760
357.2	2427	164.0	1770
365	2502	169.9	1790
372.2	2569	174.8	1805
380	2636	179.5	1820
387.2	2714	186.1	1830
398.9	2832	195.4	1830
412.8	2961	204.7	1840

TABLE 6
REDUCED CELL CHARGE AT INTERMEDIATE HEATING RATE,
ENTHALPY OF H₂ - COAL SLURRY RUN No. 6

Grams in 920cc cell	<u>H₂</u>	<u>Solvent</u>	<u>Coal</u>	<u>Total</u>
	1.24	426.9	262.4	690.5

<u>Temp.</u> <u>°C</u>	<u>Total Time</u> <u>Seconds</u>	<u>Enthalpy</u> <u>cal/gram</u>	<u>Pressure</u> <u>psia</u>
35	0	3.9	750
50	32	2.0	790
65	75	2.9	830
80	129	6.0	880
95	193	11.6	920
110	261	18.2	960
125	333	25.6	1010
140	404	32.8	1030
155	479	40.7	1110
170	557	49.3	1160
185	633	57.4	1215
200	711	65.7	1280
215	789	73.9	1340
230	869	82.2	1410
245	945	89.4	1490
260	1023	97.0	1550
275	1100	104.5	1610
290	1200	116.7	1680
305	1339	137.5	1800
320	1417	144.8	1870
335	1465	145.7	1870
350	1566	158.1	1870
365	1658	168.4	1910
380	1737	175.5	1945
398.9	1871	191.9	2010
412.8	1967	203.3	2070

Table 7

COAL ENTHALPY VS. TEMPERATURE (RUN 1)

For H₂ = 1.66g
Solvent = 456.1g
Coal = 280.0g

TEMPERATUREBTU/LBM COAL

365	0.000
392	12.697
419	24.817
446	33.425
473	44.927
500	56.793
527	61.877
540	58.677
554	65.982
567	75.422
581	80.567
594	117.350
608	112.683
621	94.770
635	111.178
648	125.272
662	135.014
675	146.892
689	154.423
702	162.188
716	164.669
729	180.727
750	197.275
775	213.962

Table 8

COAL ENTHALPY VS. TEMPERATURE (RUN 2)

For H₂ = 1.66g
Solvent = 456.1g
Coal = 280.0g

TEMPERATUREBTU/LBM COAL

365	0.000
392	14.593
419	27.313
446	43.598
473	98.666
500	109.136
527	100.058
554	123.190
581	145.784
608	206.222
635	178.953
662	269.183
689	233.886
716	197.137
750	200.510
775	241.262

Table 9

COAL ENTHALPY VS. TEMPERATURE (RUN 4)For H₂ = 1.27g, Solvent = 428.3g, Coal = 261.6g

<u>TEMPERATURE</u>	<u>BTU/LB COAL (FOR COAL)</u>
365	0.000
392	13.845
419	25.762
446	40.211
473	53.389
500	52.976
527	72.071
554	102.226
581	149.787
608	146.293
635	116.347
662	87.410
689	98.430
716	105.937
750	122.839
775	147.478

Table 10

COAL ENTHALPY VS. TEMPERATURE (RUN 5)

H₂ = 1.34g
Solvent = 426.7g
Coal = 262.0g

<u>TEMPERATURE</u>	<u>BTU/LBM (COAL)</u>
365	0.000
392	14.942
419	28.923
446	42.115
473	50.981
500	72.358
527	93.400
540	99.276
554	131.575
567	130.642
581	112.072
594	120.853
608	140.920
621	128.717
635	125.888
648	132.928
662	154.037
675	186.922
689	200.636
702	210.522
716	217.984
729	235.786
750	257.887
775	274.963

Table 11

COAL ENTHALPY VS. TEMPERATURE (RUN 6)

H₂ = 1.24g
Solvent = 426.9g
Coal = 262.4g

TEMPERATUREBTU/LBM (COAL)

365	0.000
392	15.717
419	31.106
446	46.475
473	56.308
500	67.767
527	78.401
554	110.912
501	183.648
608	192.149
635	170.108
662	202.122
689	223.621
716	229.175
750	271.376
775	298.346

Table 12

TOTAL STREAM ENTHALPYDesign
Case

Coal = 500,000 lb/hr
 Solvent = 813,016 lb/hr
 Gas Design Flow = 35,586 lb/hr

M. Wt. = 5.51

<u>TEMPERATURE</u>	ENTHALPY BTU/LBM COAL INLET GAS FLOW			
	<u>1/2 DESIGN</u>	<u>DESIGN</u>	<u>2 X DESIGN</u>	<u>3 X DESIGN</u>
500	0.000	0.0	0.0	0.0
526	38.1	39.5	42.6	45.5
552	76.6	79.6	85.8	91.8
578	164.5	169.3	178.8	187.9
604	235.8	242.4	255.4	267.9
630	205.7	214.2	330.9	247.0
656	252.1	262.9	283.3	303.2
682	303.0	316.1	340.8	364.5
708	359.6	375.2	404.2	431.9
734	418.7	437.1	471.0	502.7
760	<u>480.4</u>	<u>502.1</u>	<u>540.8</u>	<u>635.1</u>
TOTAL DUTY (MMBTU/HR)	240.2	251.1	270.4	317.6

Table 13

PERCENT VAPORIZATION VS. TEMPERATUREDESIGN CASE

<u>TEMPERATURE</u>	<u>% VAPORIZATION (COAL FREE)</u>
500	0.000
526	0.170
552	0.398
578	0.697
604	1.082
630	1.569
656	2.181
682	2.940
708	3.873
734	5.009
760	6.387

Table 14
GAS AND LIQUID FLOW RATES

<u>TEMPERATURE</u> <u>°F</u>	<u>DESIGN RATES</u> <u>lb/hr</u>	<u>(ACTUAL RATE/DESIGN RATE)</u>		
		<u>1/2 DESIGN</u>	<u>2 X DESIGN</u>	<u>3 X DESIGN</u>
500	V=34238.4	0.4168	2.1908	3.3900
	L=814362.3	1.0027	0.9936	0.9869
526	V=35625.3	0.4115	2.2021	3.4117
	L=812975.3	1.0039	0.9911	0.9819
552	V=37481.5	0.4060	2.2132	3.4324
	L=811119.1	1.0055	0.9878	0.9753
578	V=39914.0	0.4006	2.2237	3.4512
	L=808686.6	1.0076	0.9836	0.9670
604	V=43045.8	0.3953	2.2332	3.4669
	L=805554.7	1.0102	0.9783	0.9565
630	V=47019.2	0.3904	2.2410	3.4778
	L=801581.4	1.0136	0.9716	0.9434
656	V=51997.6	0.3860	2.2465	3.4823
	L=796603.1	1.0177	0.9633	0.9273
682	V=58181.5	0.3822	2.2482	3.4775
	L=790419.2	1.0230	0.9531	0.9077
708	V=65779.0	0.3793	2.2463	3.4623
	L=782821.6	1.0294	0.9407	0.8840
734	V=75032.6	0.3776	2.2382	3.4324
	L=773568.1	1.0374	0.9259	0.8561
760	V=86250.4	0.3772	2.2218	3.3833
	L=762350.3	1.0471	0.9084	0.8237

APPENDIX B

Dukler Two-Phase Flow Pressure Drop Correlation

$$\Delta P_{TP} = 4\beta f_{TP} G_T^2 L / (2 g_c \rho D)$$

where

$$\rho = \lambda \rho_L + \rho_G (1-\lambda)$$

$$\beta = \frac{\rho_L \lambda^2 + \rho_G}{\rho R_L} [(1-\lambda)^2 / (1-R_L)]$$

G_T = Total mass velocity (gas and liquid)

L = Pipe length

D = Pipe diameter

f_{TP} = Two-phase friction factor

ρ_G, ρ_L = Gas and liquid densities

λ = Liquid inlet volume fraction

R_L = Liquid holdup

The two-phase friction factor, f_{TP} , has been shown to be related to the factor f_0 by

$$\frac{f_{TP}}{f_0} = 1 - \ln \lambda / [1.281 + 0.478 \ln \lambda + 0.444 (\ln \lambda)^2 + 0.094 (\ln \lambda)^3 + 0.00843 (\ln \lambda)^4]$$

f_0 is derived from conventional friction factor vs. Reynolds number plots using a Reynolds number defined as

$$Re_{TP} = \beta D G_T / \mu N_S$$

$$\text{with } \mu N_S = \mu_L \lambda + \mu_G (1-\lambda)$$

APPENDIX C

Analysis of Ft. Lewis Fired Heater Tests

INTEROFFICE
MEMORANDUM

Date 21 January 1982

Subject ICRC Fired Heater Tests Analysis

To Distribution

From [REDACTED]

ICRC

(ORGANIZATION OR DEPARTMENT)

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Attached is a detailed report of the analysis carried out on data received during the June-August 1980 SRC-I Fired Heater Tests. These tests were conducted at the Fort Lewis Pilot Plant in Tacoma, Washington.

Prior to the issuance of this report concerns have been raised about the magnitude of the increase in the heat transfer coefficients (based on the Wilco enthalpy measurements) at the last two turns (last two values) of the heater coil. Calculations based on a uniform heat flux derived from Fort Lewis energy balances have shown good agreement with the Wilco based results except at the above mentioned sections. An increase in 'h' is shown but not to the extent indicated by the Wilco values. Work is underway to incorporate the lower values into the Demo Plant design.

B.O. Beoun

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Attachment

ANALYSIS OF FORT LEWIS

FIRED HEATER TESTS

BY

B. O. BROWN

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SUMMARY

During the June-August 1980 interim a series of experimental runs were undertaken at the Fort Lewis facilities to model and better understand the fluid dynamics to be expected within the demonstration plant preheater. A run matrix was developed so that the effects of slurry and gas rates, slurry coal concentration, mix tank temperatures and residence time could be separately assessed. Analysis of the data resulted in the development of a procedure for the prediction of heat transfer coefficients and pressure drop for the demonstration plant heater.

INTRODUCTION

The SRC-I process involves the preheating of a three-phase mixture (H_2 gas-coal-solvent) from $350^\circ F$ - $760^\circ F$ at pressures exceeding 2,000 psi. Characterization of the process stream hydrodynamics is essential to an economic preheater design. Current attempts at such have resorted to treatment of the coal and solvent as a pseudo-single-phase mixture (herein called slurry) with accompanying fluid properties. This necessitates superficial liquid velocities high enough (about 2 ft/sec) to avoid particle saltation.

It is presently known that the slurry phase has non-Newtonian characteristics (pseudo-plastic shear-thinning behavior) over the temperature range of interest. Moreover, at around 550 - $600^\circ F$, where it is believed that coal dissolution occurs initiating the various liquefaction reactions, a marked increase in slurry viscosity is observed. These, combined with the known complexities faced in the study of two-phase systems have hampered the development of fundamental equations describing the process fluid behavior, thus setting the need for relying on empirical correlations.

During the June-August interim, a series of experimental runs were undertaken at the Fort Lewis facilities in Tacoma, Washington. These runs were specifically designed to model and thus best understand the fluid dynamics to be expected within the SRC-I Demonstration Plant slurry fired heater. A run matrix was developed so that the effects of gas and slurry rates, slurry coal concentration, slurry mix tank temperature and residence time could be separately assessed. Data collected were reduced to heat transfer coefficients and apparent slurry viscosities and later scaled up to the demonstration plant conditions. The analysis procedure and results are described hereafter.

EXPERIMENTAL EQUIPMENT AND INSTRUMENTATION

The Fort Lewis slurry preheater is sized for a throughput of 50 tons/day of coal. The process coil is a 2" sch 160 321 SS pipe, 467 ft long, wound through 13-3/4 turns in an obround configuration with a continuous upward slope of approximately 1° .

Bulk fluid temperatures are measured at each odd-numbered turn and the inlet and outlet positions of the heater coil. The pressure drop profile is derived from eight DP cells; six connected between each odd-numbered turn and one each at the inlet-coil #1 and coil #13-outlet sections, respectively. The sum of the individual readings is checked by an overall cell placed between the inlet and outlet turns of the heater. Outside skin temperatures are monitored per coil turn by two sets of thermocouples, the fifth generation and ICRC types.

Firing is achieved by six floor-mounted burners, so positioned that a double fired region exists at over 70% of the coil area. Figure 1 is a plan view of the heater with the relative locations of the various measurement points. A more detailed description of the heater and its accompanying instrumentation can be found in Ref. (1).

PROCEDURE AND RUN DESCRIPTION

The Fort Lewis test program can be essentially broken up into three sections, namely:

- 1) Fired heater tests
- 2) High-temperature slurry mix tank tests
- 3) Radioactive tracer tests

Within the fired heater test sequence the type runs complete were:

- a) Solvent only - Baseline series
- b) Solvent and gas - Baseline series
- c) Slurry only - SL -70's
- d) Slurry and gas - SL-40's, 50's, 60's

Run types a and b were conceived as checks on the thermocouple accuracies, indicators of significant coke buildup and certifying the applicability of various two-phase pressure drop correlations. Types c and d provided data on the heat transfer and pressure drop characteristics of the SRC-I process mixture.

The high-temperature slurry mix tank tests (HT-series) were run to determine the operability of the slurry mix system at elevated temperatures. This test series later grew to include runs that examined the effects of increased coal concentration and mix tank residence time. Since the plant fired heater was in operation during the mix tank series, valuable heat transfer and pressure drop data were obtained.

With firing rate adjusted so as to achieve the respective bulk temperature rise (generally 350-760°F) and plant line-out achieved, bulk temperature, pressure drop, coil skin temperature, phase flowrates and compositions were constantly monitored and averaged over run durations ranging from 2-24 hours. Table 1 outlines the various run series and pertinent operating conditions.

Paramount to the understanding of the flow characteristics present in the process heater is a knowledge of the average phase residence times and thus average phase holdups. These were provided by the radioactive tracer tests. Specifically two sodium iodide scintillation detectors were located on the process lines to monitor the flow of the tracer material. One was placed 2 ft upstream of the heater inlet, the other 32 ft from the outlet. The signal from each was electronically processed and finally fed to a dual pen chart recorder and multichannel height analyzer for analog and digital recording. The instrumentation system is shown in Figure 2.

Injection of the radioactive tracer into the process stream was accomplished by means of pressurized nitrogen. Liquid phase measurements were carried out using radioactive bromine in bromophenanthrene. Argon-41 was used for the gas phase.

Inlet gas and slurry rates were varied so as to encompass those of the aforementioned series. Table 2 lists the various test conditions. A total of 12 runs were carried out.

METHOD OF ANALYSIS

Data obtained from the baseline runs (a and b of the fired heater sequence were, as mentioned previously, for verification of thermocouple accuracies, etc. The analysis of these are presented later (Appendix 2).

Data obtained with slurry phase present (fired heater test series c and d and high temperature series) were analyzed in order to extract information on the variation of local heat transfer coefficient and slurry phase viscosities with temperature. General assumptions made were:

- 1) Major coal reactions do not occur appreciably within the fired heater. Thus the presence of H_2 gas or change in slurry coal concentration does not affect the coal enthalpy curve used (see Appendix 1).
- 2) Coal and solvent phases can be combined and treated as a single homogeneous liquid phase (slurry).
- 3) Point-by-point equilibrium is achieved as one progresses down the heater.

Assumptions 1 and 3 enable the use of the computer program CYCSYN via the NGPA thermodynamic system to predict vapor and slurry flow rates at the respective measured temperatures and pressures throughout the heater coil. The significance of this was examined and is presented later (see Appendix 1). Data analysis was carried out as follows:

Heat Transfer Analysis

Skin temperatures were measured at each coil turn, while bulk temperature readings were taken at each odd-numbered turn. However, close observation of the bulk temperature profile showed an almost linear relationship

If a heat balance is written for one coil turn, neglecting axial conduction, then:

$$\text{Heat conducted out} - \text{Heat conducted in} = \text{Heat gained by process fluid} \quad (1)$$

or

$$(H_{\text{coal}} + H_{\text{gas}} + H_{\text{solvent}})_{\text{Coil \#N} + 1} - (H_{\text{coal}} + H_{\text{gas}} + H_{\text{solvent}})_{\text{Coil \#N}} = Q \quad (1a)$$

The contribution to "Q" by the gas and solvent can be obtained from the NGPA thermo-system mentioned above. That for the coal can be calculated from a coal enthalpy vs. temperature curve previously reported (Ref. 3).

The heat duty per coil turn, "Q", can also be represented by:

$$Q = U_o \pi D_o L \Delta T_{LM} \quad (2)$$

where

$$\Delta T_{LM} = \frac{(T_w - T_b)_{N+1} - (T_w - T_b)_N}{\ln \frac{(T_w - T_b)_{N+1}}{(T_w - T_b)_N}} \quad (3)$$

$$U_o = \frac{1}{\frac{D_o}{D_i h_i} + \frac{D_o \ln(D_o/D_i)}{2K}} \quad (4)$$

T_w, T_b = wall and bulk temperatures, respectively

D_o, D_i = outside and inside coil diameter

h_L = inside heat transfer coefficients

K = thermal conductivity of tube metal

N = coil number

L = length of coil turn

The ICRC skin thermocouple readings were used in the above equation, since

- 1) It is believed that they are more accurate because of their means of attachment to the tube surface (see Ref. 1).
- 2) Their locations are within the double fired region of the coil (over 70% of the coil area is double fired) and it was felt that they would be more representative of the true coil temperature.

Local heat transfer coefficients h_i as a function of the average bulk temperature over the coil interval in question could be derived from the above equations.

Pressure Drop Analysis - Slurry Viscosities

Maintaining slurry phase velocities above coal particle sortation velocities permits the treatment of the solvent and coal as a pseudo-single phase liquid, and thus the overall system as two phases. Pressure drop data obtained at Fort Lewis were used to determine apparent slurry viscosities as a function of bulk fluid temperature.

A computer program was previously developed (Ref. 4) and used in an analysis of pressure drop data received from the Wilsonville Pilot Plant. This was later modified and used in this analysis. Essentially, the program using the two-phase correlation of Dukler (Ref. 5) converges on a liquid phase viscosity that matches the observed pressure drop.

The two-phase pressure drop correlation was developed via similarity analysis assuming a constant slip between the two phases. The Dukler constant slip two phase pressure drop is defined as

$$P_{TP} = 4 \beta f_{TP} \frac{G^2 L}{2g_c D} \quad (5)$$

where $\rho = \lambda \rho_L + \rho_G (1 - \lambda) \quad (6)$

$$\beta = \frac{\rho_L}{\rho} \frac{\lambda^2}{R_L} + \frac{\rho_G}{\rho} \left\{ \frac{(1 - \lambda)^2}{(1 - R_L)} \right\} \quad (7)$$

G_T = Total mass velocity (gas + liquid)

L = Pipe length

D = Pipe diameter

f_{TP} = Two-phase friction factor

ρ_G, ρ_L = Gas and liquid densities

λ = Liquid inlet volume fraction

R_L = Liquid holdup (Hughmark correlation)

With a two-phase Reynolds Number (Re_{TP}) defined as

$$Re_{TP} = \frac{DG_G}{\mu_{NS}} \quad (8)$$

$$\text{Where } \mu_{NS} = \mu_L + \mu_G (1 - \lambda) \quad (9)$$

the conventional friction factor vs. Reynolds Number plot can be used to obtain a friction factor value f_o . This value f_o , has been shown by Dukler to be related to f_{TP} by

$$\frac{f_{TP}}{f_o} = 1 - \frac{\ln}{\{1.281 + 0.478 \ln \lambda + 0.444 (\ln \lambda)^2 + 0.094 (\ln \lambda)^3 + 0.00843 (\ln \lambda)^4\}} \quad (10)$$

Flow Regime

Three, two-phase flow pattern maps were used to determine existing flow regimes within the Fort Lewis preheater. These were:

- 1) Taitel and Dukler Flow Map (Ref. 6)
- 2) Baker Map (Ref. 7)
- 3) HTRI Map (Ref. 8)

Where applicable, the liquid phase viscosity obtained from the pressure drop analysis was used. The flow regime maps were all incorporated within the above mentioned computer program.

RESULTS AND DISCUSSION

Fired Heater Tests

Slurry Only Runs - SL-70's

The calculated heat transfer coefficients, h , and apparent slurry viscosities are plotted as functions of fluid bulk temperature in Figures 3 and 4. With increasing bulk temperatures, the calculated coefficients remain constant up to about 550-580°F, then increase sharply in the outlet turns. There is a general increase in the value of h as the slurry flow rate is increased from 5,600-11,000 lb/hr.

Within a run, slurry viscosities are observed to decrease as the fluid bulk temperature is increased from 350-470°F. A sharp increase is observed to a maximum value of about 900 lb/ft-hr (between 570-620°F) and a likewise sharp decrease towards the coil outlet. The calculated values for Run SL-71 ($M = 5,600$ lb/hr) do not follow the trend as observed in Runs SL-72 and SL-73. It is believed that the apparent viscosities for SL-72 are in error due to the fact that the sum of the individual "DP" cell readings is 60% above that of the overall "DP" cell (48 psi as opposed to 30 psi). There is an indication of a reduction in slurry viscosity with increasing slurry rates up to bulk temperatures of 570°F; however, a reverse trend is noted at the outlet.

For Runs SL-72 and SL-73, the maximum viscosity (maximum pressure drop region) tends to follow the region of lowest h . Thus, in some sections both the apparent viscosity and heat transfer coefficient are increasing. A possible explanation is presented later.

Slurry and Gas - SL-40s, 50s and 60s

Run series SL-40s, 50s, and 60s examined the effects of increasing phase flow rates on heat transfer coefficients and apparent slurry viscosities. The SL-40 series was to determine the effect of increasing gas rate, with slurry, velocities at the proposed Demo Plant Value of approximately 4 ft/sec; SL-50s did likewise, however, with slurry velocities at 2 ft/sec. Series SL-60, with gas velocity held at 8 ft/sec (proposed Demo Plant Value) examined the effect of increasing slurry rate.

From the values presented in Figures 5, 6, and 7, it is seen that increasing phase flow rate, increases the heat transfer coefficient. Within a run, the value of h increases from the coil inlet (350°F) to an intermediate maximum at around 450-500°F, decreases to a minimum between 550-600°F, then sharply increases towards the coil outlet (760°F). For the SL-50 series the outlet values of h approach a second maximum at 670-700°F and decreases at the outlet. Due to the high ratio of gas/liquid rates such decrease might be as a result of a flow regime transition near the outlet sections of the coil.

Derived viscosities are shown in Figures 8-10. There is an indication of pseudo-plastic, shear-thinning behavior especially within the temperature

range of 600-700°F where the highest apparent viscosities are observed. The highest values are achieved with the lower gas rate (lowest slurry rate for the SL-60 series) runs; however, these maximum values of apparent viscosities are still lower than those of the slurry-only runs. In the early section of the coil (350-500°F), few values of apparent viscosity are reported. This was due to the inability of the program to converge on a suitable viscosity to match the measured pressure gradient. For both the high gas rate runs (SL-45 and SL-55) the trend indicated by the derived apparent viscosities, although similar, differ from that of the other runs within their respective series. Since there is no similar disparity in the heat transfer coefficients, this might be an artifact of the pressure drop analysis. It is noted that the pressure drop maximum again normally lag the heat transfer minimum, but it is interesting to note the agreement (for SL-45 + SL-55) of the viscosity maximum and heat transfer minimum.

High Temperature Tests

HT-11 through H-14 - Effect of Mix Tank Temperature

The effect of increasing the mix tank temperature (inlet bulk temperature to the fired heater) from 350-410°F, on the process fluid hydrodynamics were assessed. From Figures 11 and 12 it can be seen that only within the early portion of the heater coil (inlet - 550°F) are any effects noted. Higher inlet bulk temperatures increase slurry apparent viscosity and decreases the inside heat transfer coefficient.

It is suspected that this apparent change in hydrodynamics properties is a result of differing radial temperature profiles, at the same measured bulk temperature. The measured skin temperature profiles (and bulk temperatures) also differ for all four runs up to 570°F, then agree thereafter. With differing radial profiles each run would exhibit dissimilar inside wall temperatures and thus different apparent viscosities, which inevitably leads to the apparent differences in fluid flow behavior.

HT-15-1 through HT-15-12 - Effect of Mix Tank Residence Time and Coal Concentration

With gas and slurry rate held constant at 250 lbs/hr and 9,110 lb/hr respectively, the effects of mix tank residence time and coal concentration were assessed. Slurry coal concentration was varied from 38-45 wt.%, while at each concentration mix tank residence time varied between 3.5-24.6 minutes.

From the results depicted in Figures 13-18, no dependency on residence time is observed. As previously noted in the other gas-slurry runs, the calculated heat transfer decreases from the inlet of the coil, then sharply increases towards the outlet. The minimum value of 'h' is shifted from approximately 550 to 450°F with increasing coal concentration. Apparent slurry viscosities are observed to decrease, then increase sharply and later decrease toward the coil outlet. The region of highest apparent viscosity is approached earlier (shifting from 650°F to 530°F) with increasing coal concentration. At a coal concentration of 45 wt.%, two minimum values for h and two maximum values for $\mu(\text{app})$ are observed. The first viscosity maximum is observed to lag behind the first heat transfer minimum; the second agrees with the second heat transfer minimum which occurs at around 650-670°. Respective values calculated for Run HT-15-9 are observed to follow a trend separate from that of the other

corresponding runs at 45 wt.% coal. The disparity was attributed to non-steady-state conditions.

The results plotted in Figures 13-18 are replotted in Figures 19-26, where the effects of coal concentration can be more readily seen.

Flow Regimes

Predicted flow patterns, by the flow regime maps of Taitel and Dukler, Baker and HTRI, are shown in Figures 27-29 for the various run series. The predictions of Taitel and Dukler are in agreement with the HTRI map, showing a transition from a slug (intermittent) regime to annular flow with increasing gas rate. This differs from that of the Baker map which shows a transition from bubble or froth flow to a dispersed regime.

It was intended to support the various flow map predictions with high speed pressure drop readings in the manner of Weisman, et al (ref. 9), however, this proved unsuccessful due to the spacings of the pressure taps. However, the trend indicated by the overall pressure drop with increasing gas rate by run series SL-40 and SL-50 proved instructive. It has been shown (Ref. 10, 11) that a pressure drop reduction can occur in gas-liquid systems when the liquid phase is non-Newtonian (shear-thinning) and flows in a laminar slug regime. The coal slurry within the temperature range of interest, 350-760°F, has been shown to be non-Newtonian (Ref. 12). Figures 30 and 31 show the overall coil pressure drop as a function of gas rate. For both curves a minimum value of the pressure drop is indicated.

Evidence for Possible Gas Generation from Coal Dissolution

In the majority of the data mentioned above, it was noted that the derived viscosity maximum occurs downstream of the minimum value of the calculated heat transfer coefficients. If, for the moment, attention is paid to the slurry-only runs, then the occurrence and possible explanation can be better grasped.

The measured pressure drop profiles for the slurry only runs (SL-70's) suggest, via derived viscosities with the assumption of a single-phase system, a laminar flow behavior throughout the heater. Present understanding of laminar single-phase systems would require at constant mass flow rates, the heat transfer coefficient 'h', to remain constant with increasing or decreasing viscosity, $\mu(\text{app})$. However, as shown in Figures 3 and 4 (note apparent viscosities for run SL-71 are suspect) this is only obeyed up to 500°F; whereafter sharp increases in the value of 'h' occur, simultaneously with increasing $\mu(\text{app})$. Above 600°F $\mu(\text{app})$ starts decreasing, but 'h' is still increasing. Similar behavior in 'h' was reported by Oak Ridge (Ref. 13); however, corresponding pressure drop measurements were not taken.

A plausible explanation for the increasing value of 'h' in an apparent laminar system, is that whereby vapor is produced during and/or after coal dissolution. The increased volumetric flow rate would increase 'h' and the measured pressure drop. Due to the assumption of a single-phase system, the increased pressure drop would be erroneously treated as a viscosity increase. If, as presently suspected, the slurry viscosity does increase somewhat and then decrease with increasing temperature, then the combined effect of such with gas generation

would explain the increasing 'h' within a region of increasing and then decreasing pressure drop.

Another indication of possible gas generation is seen in the fifth generation thermocouple skin temperature profiles. These thermocouples, although not used in the heat transfer calculations, provided valuable insight into the possible flow mechanisms occurring within the fired heater. If, for the moment, attention is focused on Runs SL-73, SL-41R and SL-51, Figures 32-34, it can be seen that within certain regions of the preheater coil there is a discrepancy between the two fifth generation T.C.s (at odd-numbered turns), that are placed at different angular positions within a cross-sectional area. As shown in Figure 1, one thermocouple faces the flame at the east end of the heater, while the other is in the same plane away from the flame. Initially it was felt that the profile shown in Figure 32 for SL-73 (up to coil #7) was due to heat flux variation (varying view factors) around the coil circumference and that at the outlet sections due to an even flux distribution caused by high gas turbulence outside the coil. However, with gas injection (SL-41R and SL-51R) the discrepancies noticed in SL-73 at the earlier turns disappear.

With a slurry rate of approximately 11,000 lbs/hr (/4 ft/sec) in SL-73 and SL-41R, and a gas rate in SL-41R of 150 lbs/hr (2 ft/sec), the added heat duty in the latter is insignificant. This can be seen in the increase in natural gas firing from 262-265 lbs/hr. Indeed, if the discrepancy in SL-73 was due to heat flux variation, then the increase in inside coefficient by gas injection should aggravate the difference in the T.S.'s readings. Wall conduction tends to smooth wall temperatures profiles in furnaces with circumferential variation in heat fluxes; however, with increasing inside coefficient, at essentially similar firing rates and bulk temperatures, the smoothing ability of the wall conduction diminishes. Thus, it is felt that the discrepancy is due to occurrences in and not outside the coil, and leads to the inference that there is no circumferential heat flux variations within the furnace.

It has been shown (Ref. 14) that for laminar flow within a helical coil secondary flow induced from the circulatory motion of the fluid, will result in circumferential skin temperature various even for constant wall heat fluxes. Due to the almost constant value of 'h' and the suggestion of laminar flow from the pressure drop profile at the earlier sections of the coil together with the fact that the fifth generation thermocouples are located just after a bend in the process coil, it is felt that secondary flow is responsible for the observed deviation. Turbulent flow tends to dampen the effects of secondary fluid motion, thus the observed agreement at the outlet of the coil for the slurry only runs would be explained by gas-induced turbulence.

For Run SL-51 where the gas rate is approximately the same as in SL-41R (130 lbs/hr) but slurry rate at 5,500 lbs/hr, and Run SL-41R, similar profiles are noted. This shows how effective the gas is in making the process stream behave in a turbulent manner.

Radioactive Tracer Tests

Previously obtained data from the Fort Lewis Tracer Tests were analyzed and the results reported. Tables 3 and 4 lists the calculated phase residence times and holdups. In general the results can be summarized as follows:

- 1) Gas phase residence times varied from 0.5-0.88 minutes; liquid phase from 1.19-2.18 minutes.
- 2) Liquid phase holdups varied from 0.32-0.61.
- 3) Increasing gas rate at a constant liquid rate decreases both gas and liquid residence times.
- 4) Increasing liquid rate at a constant gas rate tends to decrease liquid residence time, however, negligible effect is observed on gas residence time.
- 5) Liquid holdup increases with increasing liquid rate or decreasing gas rate.

Due to the significant solubility of argon within the slurry phase, the gas residence times and, thus, phase holdups, are questionable. Further details can be found in the Appendix 5.

CORRELATION OF FORT LEWIS RESULTS AND SCALE-UP TO DEMO PLANT

Heat Transfer

A theoretical treatment of the problem of heat transfer within a laminar slug flow regime (see Appendix 4) indicated that the heat transfer coefficient 'h' can be represented by

$$H = K_1 D^{-1/2} + K_2 D^{-2/3} \quad (11)$$

where K_1 , K_2 are constants,

and D = pipe diameter.

This was based on the assumption that phase velocities, temperature profiles and the ratio of slug length (gas and liquid) to pipe diameter, lu/D , would remain constant with pipe diameter changes. Inherent in this development also is that a slug flow regime will prevail within the Demonstration Plant heater. This had been previously assessed and reported (Ref. 15). Due to the duplication of proposed Demo Plant Heater Flow Conditions (phase velocities, etc.), the Fort Lewis calculated heat transfer coefficients could thus be used via (Equation 11) to predict expected flow Demo Plant Values. The values of heat transfer coefficients (Fort Lewis calculated) shown for run series SL-40, SL-50, SL-60 were correlated in the form of

$$H_{FL} = A M_L^a M_G^b \quad (12)$$

where A, a, b - are constants (shown in Table A),

M_L, M_G - mass and liquid flow rates in lbs/hr,

$h = \text{Btu/hr. ft}^2 \text{ } ^\circ\text{F}$ (outside area).

This form of correlating the data was chosen above the general Reynolds number, Prandtl type correlation due to:

- 1) Inability to quantify Re, Pr from uncertain physical properties (μ_L, k_L).
- 2) Its simplicity and direct application to Demo Plant Design.

For the run series used (SL-40s, 50s and 60s) the bulk temperature profile were similar (350-760°F), thus enabling the determination of the constants A, a, and b as functions of temperature. These are listed in Table 5. The predicted vs. measured values of 'h' are shown in Figure 35A. It is seen that over 97% of the data is predicted within $\pm 30\%$ of the measured value. An average value of 0.789 was found for the correlation coefficient, R^2 .

In order to be as conservative as possible within reason, it was decided to scaleup bases on

$$h \propto D^{-2/3} \quad (13)$$

as opposed to that shown in Equation 11.

Thus, remembering the general assumptions made in the development of Equation A, the Demonstration Plant Heater coefficient can be predicted from

$$H_{\text{Demo}} = A' M_L^a M_G^b$$

where M_L, M_G are Demo Plant Mass Flow Rates are lbs/hr,

a, b are constants as shown in Table A.

$$A' = A \left(\text{from Table 1} \right) \times \left(\frac{D_{FL}}{D_{\text{Demo}}} \right) \{ 2(a + B) + 2/3 \} \times \frac{(\text{Demo})_{\text{inside}}}{(\text{Demo})_{\text{outside}}}$$

D = Pipe diameter,

FL, Demo = Fort Lewis and Demo Plant, respectively,

The values of A' are listed in Table 6.

Slurry Viscosity - Pressure Drop

Derived viscosities from runs having similar slurry compositions as that proposed for the demonstration plants were shown previously in Figures 4, 8, 9, and 10. Due to the lack of sufficient data for the extraction of fundamental equations, it was decided to plot a smooth curve connecting the maximum calculated values at each temperature. The resultant curve is shown in Figure 36.

In predicting demonstration plant pressure drop it is important that the procedure used in generating this curve (Figure 36) be reversed. That is, CYCSYN should be used to generate vapor and liquid rates to be used in the Dukler Constant Slip correlation for Two-Phase Pressure Drop Predictions.

CONCLUSIONS AND RECOMMENDATIONS

The above analysis has revealed the following:

- 1) Increasing gas and slurry rate increases slurry heat transfer coefficient.
- 2) Slug flow occurs throughout the heater based on the various map predictions and the reduction of pressure drop with increasing gas rate in the central temperature zone.
- 3) For the slurry only runs there is a strong indication of gas formulation possibly from the initial stages of coal reactions, at temperatures approaching 650°F.
- 4) The presence of a small quantity of gas (1.5 wt %) results in the turbulent behavior of the process stream.
- 5) There is a strong indication of the presence of secondary flow at the coil bends especially within the region of high viscosity.
- 6) Increasing slurry coal concentration shows the region of highest viscosity to be approached earlier (shift from 650-530°F with a 38-40 wt % change in coal concentration).
- 7) Correlation for heat transfer coefficients and pressure drop in the SRC-I fired heater are presented and recommended for design purposes.

With the above in mind, it is recommended that:

- 1) A repeated set of runs be carried out at a different coil diameter so as to assess the scaling procedure used.
- 2) Vapor-slurry equilibrium data be obtained throughout the temperature range of interest.
- 3) Reinstrumentation of fired heater so as to obtain measurements over smaller increments and, if possible, for pressure and temperatures to be recorded over the same size increment.

- 4) A more detailed study of the flow characteristics of the coal-slurry should be carried out in view of the fact of possible gas generation around 600°F. Measurement by pipeline viscometers could be much in error.

Unfortunately, the shutdown of the Fort Lewis pilot plant prevents any further preheater tests.

FIGURE 1 - FIRED HEATER PLAN VIEW

FIGURE 2

ELECTRONIC EQUIPMENT AND SCHEMATIC OF TRACER FLOW

FIG. 3 - CALCULATED HEAT TRANSFER
COEFFICIENTS (SLURRY ONLY RUNS)

**FIG. 4 - APPARENT VISCOSITIES
(SLURRY ONLY RUNS)**

FIG. 5 - CALCULATED COEFFICIENTS SL-40'S

FIG. 6 - CALCULATED COEFFICIENTS SL-50'S

FIG. 7 - CALCULATED COEFFICIENTS SL-60'S

FIG. 8 - APPARENT VISCOSITIES SL-40's

FIG. 9 - APPARENT VISCOSITIES SL-50'S

FIG. 10 - APPARENT VISCOSITIES SL-60'S

**FIG. 11 - CALCULATED COEFFICIENTS
(EFFECT OF RESIDENCE TIME) HT-11-14**

FIG. 12 - APPARENT VISCOSITIES HT-11-14

FIG. 13 - CALCULATED COEFFICIENTS HT-15-1-4

FIG. 14 - CALCULATED COEFFICIENTS HT-15-5-8

FIG. 15 - CALCULATED COEFFICIENTS HT-15-9-12

FIG. 16 - APPARENT VISCOSITIES HT-15-1-4

FIG. 17 - APPARENT VISCOSITIES HT-15-5-8

FIG. 18 - APPARENT VISCOSITIES HT-15-9-12

**FIG. 19 - CALCULATED COEFFICIENTS
(EFFECT OF COAL CONCENTRATION)**

FIG. 20 - CALCULATED COEFFICIENTS
(EFFECT OF COAL CONCENTRATION)

**FIG. 21 - CALCULATED COEFFICIENTS
(EFFECT OF COAL CONCENTRATION)**

**FIG. 22 - CALCULATED COEFFICIENTS
(EFFECT OF COAL CONCENTRATION)**

**FIG. 23 - APPARENT VISCOSITIES
(EFFECT OF COAL CONCENTRATION)**

**FIG. 24 - APPARENT VISCOSITIES
(EFFECT OF COAL CONCENTRATION)**

**FIG. 25 - APPARENT VISCOSITIES
(EFFECT OF COAL CONCENTRATION)**

FIG. 26 - APPARENT VISCOSITIES
(EFFECT OF COAL CONCENTRATION)

FIG. 27 - TAITEL AND DUKLER FLOW REGIME PREDICTIONS

FIG. 28 - BAKER FLOW REGIME PREDICTIONS

$$By = 2.16 W_V / A \sqrt{\rho_V \rho_L} \quad Bx = 396 (W_L / W_V) \mu_L^{1/3} (\rho_V \rho_L)^{1/2} / \rho_L^{2/3} \sigma_L$$

WHERE: A = AREA, FT^2
 W_V = FLOW RATE OF VAPOR, LB/HR
 W_L = FLOW RATE OF LIQUID, LB/HR

μ_L = VISCOSITY OF LIQUID, $LB/FT/HR$
 ρ_V = DENSITY OF VAPOR, LB/FT^3
 ρ_L = DENSITY OF LIQUID, LB/FT^3
 σ_L = SURFACE TENSION OF LIQUID, $DYNES/CM$

FIG. 29 HTRI PREDICTIONS

FIG. 30 - OVERALL PRESSURE DROP VS. GAS RATE

FIG. 31 - OVERALL PRESSURE DROP VS. GAS RATE

FIG. 32 - FIRED HEATER TEMPERATURE PROFILE

FIG. 33 - FIRED HEATER TEMPERATURE PROFILE

FIG. 34 - FIRED HEATER TEMPERATURE PROFILE

FIG. 35 - H PREDICTED VS. H MEASURED (SL-40'S, 50'S, 60'S)

FIG. 36 - SLURRY VISCOSITY VS. TEMPERATURE

TABLE 1 - FIRED HEATER TESTS RUN HISTORY

Series	Run Name	Date Run	Duration (hrs)	Sum of Indiv. DP's (PSI)	Overall DP (PSI)	Temp. In/Temp. Out °F	Gas Rate lb/hr	Coal Rate (MF) lb/hr	Solvent Rate lb/hr	Mol % H ₂	Wt % Coal in Slurry
Baseline	B-15R	7/23	2	52.0	53.0	355/352	397.4	--	10,030	86.2	--
	B-22R	7/24	2	14.5	13.0	347/761	82.7	--	10,110	83.4	--
	B-25R	7/24	2	56.5	56.5	361/761	401.2	--	10,172	83.9	--
	B-31R	7/25	2	11.0	7.0	327/761	84.0	--	5,040	83.8	--
	B-41R	7/23	2	14.0	0.0	270/650	--	--	10,960	--	--
	B-45	8/8	3	33.5	5.5	276/653	--	--	11,220	--	--
Slurry & Gas	SL-41R	7/28	6	60.2	59.3	345/768	148.2	4,220	6,810	86.4	38.3
	SL-42R	8/5	8	53.1	46.4	347/760	288.5	4,060	6,750	85.8	37.6
	SL-43R	7/27	8	70.1	70.1	352/764	602.5	4,150	6,770	85.9	38.0
	SL-44R	8/3	8	70.0	65.3	356/761	791.1	4,140	6,740	85.0	38.1
	SL-45R	7/27	8	100.5	100.1	362/764	1222.1	4,160	6,770	85.7	38.1
	SL-46	8/7	8	71.0	66.0	351/763	706.0	4,150	6,770	N/A	38.0
Slurry & Gas	SL-51	7/30	8	42.3	37.5	322/761	131.7	2,100	3,430	88.5	38.0
	SL-53R	7/31	8	44.2	40.1	341/760	607.2	2,130	3,420	87.4	38.4
	SL-54R	8/3	10	47.6	43.3	344/761	796.0	2,070	3,410	85.6	37.8
	SL-55R	7/30	8	69.0	64.4	349/760	1191.7	2,090	3,420	86.3	37.9
Slurry & Gas	SL-61	7/31	8	38.1	32.9	333/761	303.5	2,130	3,410	85.7	38.5
	SL-62	7/31	8	35.9	30.5	339/760	289.0	2,610	4,260	86.2	38.0
	SL-63	8/1	9	42.0	37.6	344/761	292.8	3,170	5,100	87.2	38.3
	SL-64	8/1	8	44.7	38.6	348/761	301.0	3,620	5,900	84.9	38.0
	SL-65	8/2	8	50.4	45.0	348/760	312.7	4,150	6,760	84.6	38.0
Slurry Only	SL-71	8/6	2	48	30	316/767	--	2,090	3,430	86.9	37.9
	SL-72	8/6	1	50	42	351/763	--	3,240	4,960	86.9	39.5
	SL-73	8/6	2	68	64	350/757	--	4,360	6,760	86.9	39.5

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Table I (continued)

Series	Run Name	Date Run	Duration (hrs)	Sum of Indiv. DP's (PSI)	Overall DP (PSI)	Temp. In/Temp. Out °F	Gas Rate lb/hr	Coal Rate (MF) lb/hr	Solvent Rate lb/hr	Mol % H ₂	Wt % Coal in Slurry
High Temp.	HT-11	8/9	24	62.3	58.2	344/763	218.4	3,464	5,633	86.0	38.1
	HT-12	8/11	24	58.4	55.4	369/759	238.4	4,451	5,642	86.0	38.0
	HT-13	8/13	24	63.2	58.8	388/759	264.4	3,437	5,485	86.0	38.5
	HT-14	8/16	24	80.6	76.0	416/760	257.0	3,449	5,455	86.0	38.7
High Temp.	HT-15-1	8/14	2	50.0	50.5	348/760	253	3,460	5,650	N/A	38.0
	HT-15-2	8/14	2	53.0	54.0	346/759	257	3,460	5,650	N/A	38.0
	HT-15-3	8/14	2	61.0	56.0	345/759	257	3,460	5,655	N/A	38.0
	HT-15-4	8/15	2	62.0	56.0	345/759	258	3,460	5,655	N/A	38.0
High Temp.	HT-15-5	8/15	2	79.5	74.5	349/760	251	3,960	5,640	N/A	41.2
	HT-15-6	8/15	2	83.5	79.0	345/759	253	3,960	5,640	N/A	41.2
	HT-15-7	8/15	2	82.5	79.0	347/759	250	3,960	5,650	N/A	41.2
	HT-15-8	8/15	2	76.5	73.5	346/759	246	3,960	5,445	N/A	42.1
High Temp.	HT-15-9	8/15	2	84.0	80.0	345/760	241	4,460	5,630	N/A	45.3
	HT-15-10	8/15	2	110.5	106.0	349/760	238	4,600	5,630	N/A	45.0
	HT-15-11	8/15	2	110.5	105.0	351/758	231	4,510	5,670	N/A	44.3
	HT-15-12	8/15	2	109.5	105.0	352/758	240	4,570	5,655	N/A	44.7
Radio-active Tracer Tests	RT-1	8/12					227	9,080*			
	RT-2	8/12					224	9,070			
	RT-3	8/12					574	9,050			
	RT-4	8/12					569	9,070			
	RT-5	8/12					537	10,960			
	RT-6	8/12					539	10,960			
	RT-7	8/12					269	10,970			
	RT-8	8/12					270	10,960			
	RT-9	8/12					264	5,600			
	RT-10	8/12					266	5,550			
	RT-11	8/12					532	5,490			
	RT-12	8/12					534	5,530			

*Slurry Rate

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TABLE II

<u>Tracer Test No.</u>	<u>V*</u>	<u>L*</u>
	<u>lb/hr</u>	
RT-1	227	9080
RT-2	224	9070
RT-3	574	9050
RT-4	569	9070
RT-5	537	10960
RT-6	539	10960
RT-7	269	10970
RT-8	270	10960
RT-9	264	5600
RT-10	266	5550
RT-11	532	5490
RT-12	534	5530

*V = Vapor

*L = Slurry

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Table III
CALCULATED PHASE RESIDENCE TIMES

<u>Tracer Test No.</u>	<u>V*</u>	<u>L*</u>	<u>Residence Times</u>
	<u>lb/hr</u>		<u>Min.</u>
RT-1	227	9080	V = 0.88
RT-2	224	9070	L = 1.83
RT-3	574	9050	V = 0.57
			L = 1.23
RT-4	569	9070	L = 1.24
RT-5	537	10960	V = 0.57
			L = 1.20
RT-6	539	10960	L = 1.19
RT-7	269	10970	V = 0.77
			L = 1.49
RT-8	270	10960	L = 1.50
RT-9	264	5600	L = 2.18
RT-10	266	5550	V = 0.79
RT-11	532	5490	V = 0.51
RT-12	534	5530	L = 1.62

*V = Vapor

*L = Slurry

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TABLE IV
LIQUID PHASE HOLDUPS

<u>Test No.</u>	<u>Hughmark^{(1)*}</u>	<u>Hughmark^{(2)*}</u>	<u>Tracer Residence Time Dat</u>
RT 1 & 2	0.62	0.72	0.61
RT-3 & 4	--	--	--
RT-5 & 6	0.42	0.60	0.40
RT-7 & 8	0.57	0.70	0.56
RT-9 & 10	0.45	--	0.46
RT-11 & 12	0.33	0.54	0.32

*Hughmark (1) - generalized correlation (Ref. 2)
(2) - slug flow correlation (Ref. 3)

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TABLE V
FORT LEWIS CORRELATION COEFFICIENTS

General Equation $H = A M_L^a M_G^b$

Coil #	Coefficients				Temp. Range
	A	a	b	R ²	
1-2	3.3175×10^{-2}	0.672	0.340	0.879	375 - 410
2-3	2.5845×10^{-2}	0.693	0.372	0.963	410 - 440
3-4	5.3609×10^{-1}	0.367	0.365	0.860	440 - 470
4-5	3.4973×10^{-2}	0.633	0.401	0.916	470 - 500
5-6	6.5440×10^{-2}	0.574	0.329	0.736	500 - 530
6-7	5.1405×10^{-2}	0.602	0.318	0.834	530 - 560
7-8	1.5175×10^{-2}	0.625	0.481	0.850	560 - 590
8-9	5.4616×10^{-2}	0.506	0.471	0.739	590 - 620
9-10	1.6491×10^0	0.282	0.312	0.532	620 - 650
10-11	1.2411×10^1	0.127	0.268	0.603	650 - 670
11-12	7.5279×10^{-1}	0.507	0.209	0.771	670 - 700
12-13	1.0534×10^{-1}	0.826	0.089	0.786	700 - 740

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TABLE VI
DEMO PLANT CORRELATION COEFFICIENTS

General Equation $H = A m_L^a m_G^b$

<u>Temp. Range °F</u>	<u>A</u>	<u>Coefficients</u>	
		<u>a</u>	<u>b</u>
375 - 410	6.1270×10^{-4}	0.672	0.340
410 - 440	4.1167×10^{-4}	0.693	0.372
440 - 470	2.1635×10^{-2}	0.367	0.365
470 - 500	6.0743×10^{-4}	0.633	0.401
500 - 530	1.6385×10^{-3}	0.574	0.329
530 - 560	1.2274×10^{-3}	0.602	0.318
560 - 590	2.1557×10^{-4}	0.625	0.481
590 - 620	1.1122×10^{-3}	0.506	0.471
620 - 650	9.7833×10^{-2}	0.282	0.312
650 - 670	1.2833×10^0	0.127	0.268
670 - 700	3.1768×10^{-2}	0.507	0.209
700 - 740	2.5506×10^{-3}	0.826	0.089

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APPENDIX I

Process Stream Physical Properties and Vapor Liquid Phase Equilibrium

The derivation of slurry phase viscosities and local heat transfer coefficients requires a knowledge of vapor/liquid phase flow rates and properties as functions of temperature. The following describes the development of each.

Phase Flow Rates

Each run at Fort Lewis was constantly monitored for gas and solvent inlet compositions and flow rates. Using the Bulk Temperature and Pressure Drop Profile for each run, the CYCSYN Flash Program was used via the NGPA Thermodynamic System to determine the VLE split throughout the heater. In this routine, coal was neglected, based on the assumption that it didn't affect the vapor-liquid equilibrium between the H_2 -rich gas and solvent. This was due to the fact:

- 1) Coal is suspected to go into solution around 600°F which occurs at a point about 60% the height of the preheater.
- 2) In order to develop a tractable procedure of calculating heat duty from available data, such was necessary.

This assumption was later checked by taking an arbitrary run and includes coal within the flash calculations by modeling it as SRC. It was observed that the major changes occur within the gas phase whereby mass and volumetric rates were reduced by 30% and 10% respectively.

The liquid portions from the CYCSYN calculations were added to the inlet coal rate and this constituted the slurry phase.

Vapor/Slurry Densities

Gas phase densities as predicted by CYCSYN were used. These range from 1.2 lb/ft³ to 2.6 lb/ft³ over the temperature ranges studied. Slurry densities were reported by Oak Ridge (Ref. 16) for coal concentration of 35 wt %. These were adjusted for a 38 wt % coal solvent mixture and used. The resulting curve is shown in Figure 37.

Gas Viscosities

A previous analysis of Wilsonville data (Ref. 4) had shown that within the range of expected values, 0.03 - 0.05 lb m/ft-hr negligible effect was observed in the predicted slurry viscosities. For this analysis a value of 0.04 lb m/ft-hr was used.

Vapor/Slurry Enthalpies - Heat Duty

Prior to the Fort Lewis tests, H_2 gas-coal slurry temperature-enthalpy measurements taken by the Wilco Research Company, were analyzed and reported (Ref. 3). Via CYCSYN a Coal-Enthalpy vs. Temperature was derived (Figure 38).

In the determination of the heat duty to a coil segment it is important that the procedure used in deriving the coal enthalpy be followed. That is, the H_2 -rich gas and solvent portions of the feed are used in the CYCSYN Flash Procedures, for the prediction of phase enthalpies and other stream properties. The difference in predicted enthalpies for the inlet and outlet sections of the coil segment under consideration is then added to that for the coal (derived from Figure 38).

**FIG. 37 - SLURRY DENSITY VS. TEMPERATURE
(FORT LEWIS RUNS JULY & AUG. 80)**

FIG. 38 - COAL ENTHALPY VS. TEMPERATURE

APPENDIX II

Baseline Runs - Check-In Pressure Drop Correlations and Thermocouple Accuracy

Pressure Drop Correlations

At the onset of the Fort Lewis test sequence, the gas-solvent runs of the Baseline series were conceived as checks on the accuracy of current Two-Phase Flow Pressure Drop correlations. Preliminary results, however, showed very low 'DP' readings (zero values were recorded for some coil sections) and thus, it was decided to use the Constant Slip Correlation of Dukler. This was developed from Similarity Analysis and later shown by the author to be reasonably accurate over a wide range of operating variables and flow regimes (Ref. 4).

Later in the program a high gas and liquid rate run (B-25R) was carried out which showed good agreement between the sum of the individual 'DP' readings and that of the overall 'DP' cell. Although no viscosity measurement was available for the solvent used in this run, an analysis as carried out for the coal slurry runs predicted values of solvent viscosity ranging from 1.4 to 11.4 lb/ft-hr (0.6 to 4.7 cp). Solvent viscosities ranging from 0.3 to 3.1 cp have been measured at Fort Lewis during the test duration. Table 7 outlines pertinent flow parameters for Run B-25R. The calculated values for the last three sections of the coil are considered suspect owing to the nonagreement with the trend set forth by the other coil sections.

Thermocouple Accuracies

The solvent-only runs of the baseline series were originally conceived as a possible means of correcting for suspected contact resistances of existing skin thermocouple. The idea, however, later proved fruitless, but the solvent runs soon proved to be an effective means of ascertaining thermocouple accuracies throughout the series.

As shown in Figures 39-40, the skin temperature profile, as measured by the fourth generation thermocouple for the solvent-only run and gas-slurry runs respectively, showed appreciable scatter within a respective coil position. Furthermore, for the slurry run no characteristic hump in the skin profile (indicating a region of highest viscosity) was noticed. It was then hypothesized that via the solvent-only runs whereby fluid properties are known, a contact resistance would be calculated for each thermocouple. However, further assessment of the problem revealed previously unseen complexities and the idea was aborted.

Expected skin temperatures were calculated via the Dittus-Boelter equation and the comparison with measured profiles (see Figure 41) presented to the pilot plant management. A plant shutdown was called for and modifications made to the thermocouples, resulting in a fifth generation of thermocouples. The results of a second solvent-only run with the predicted values are shown in Figure 42. Good agreement is observed.

After the installation of a new type thermocouple, to be later known as the ICRC Thermocouple, the various run series were carried out. The results of a similar solvent-only run carried out later in the test sequence are shown in Figure 43. Good agreement is again observed; however, it is seen that the fifth generation tends to read higher (up to +30°F) than the ICRC Thermocouple.

TABLE 7 - FLOW PARAMETERS FOR RUN B-25R

Inlet Gas Rate = 1,491 lb/hr

Inlet Solvent Rate = 10,172 lb/hr

Bulk Temperature Rise = 362-761°F

<u>Coil Section</u>	<u>Flow Rates^(*)</u>		<u>Density^(*)</u>		<u>Press. Gradient⁽²⁾</u>		<u>Calc. Liquid Viscosity</u>
	<u>Gas</u>	<u>Solv.</u>	<u>Gas</u>	<u>Solv.</u>	<u>Meas.</u>	<u>Calc.</u>	
	lb/hr		lb/ft ³		PSI/1000 ft		lb/ft-hr
Inlet Coil #1	1481	10184	1.42	58.0	123.1	123.7	11.4
1 - 3	1511	10153	1.38	57.4	134.1	134.8	15.0
3 - 5	1604	10060	1.37	55.5	115.8	125.0	7.9
5 - 7	1772	9892	1.42	54.6	107.4	113.5	3.6
7 - 9	2064	9600	1.56	52.7	104.3	114.6	2.6
9 - 11	2502	9162	1.78	51.5	96.9	90.2	0.1
11 - 13	3102	8562	2.08	49.9	178.8	179.3	14.3
13 - outlet	3717	7947	2.37	48.7	89.3	105.1	0.1

(*) Calculated via CYCSYN

(2) A constant gas viscosity of 0.04 lb/ft-hr has been used.

FIG. 39 - FIRED HEATER TEMPERATURE PROFILE

FIG. 40 - FIRED HEATER TEMPERATURE PROFILE

FIG. 41 - FIRED HEATER TEMPERATURE PROFILE

FIG. 42 - FIRED HEATER TEMPERATURE PROFILE

FIG. 43 FIRED HEATER TEMPERATURE PROFILE

APPENDIX III

Accuracy of Heat Duty Calculations

For each run at Fort Lewis, the heat duty on the furnace is calculated via two methods, namely:

- 1) Fort Lewis I - Based on the assumption of a constant specific heat for gas and slurry and an endotherm equivalent to 150 Btu/lb M for coal.
- 2) Fort Lewis II - Based on an energy balance for the heater.

The heat duty based off the ICRC procedure in Table 8 with that of the above methods. The calculated values of Fort Lewis (I) tends to be higher than those of the ICRC procedure by as much as 6%. Most of the ICRC values fall within $\pm 10\%$ of the Fort Lewis (II) duties. The highest deviation, 22%, is found for the lowest slurry rate Run SL-71. For the high temperature runs HT-13 through HT-15-12 erroneous values were calculated by the Fort Lewis II procedure. It is suspected that these were due to an error in the natural gas meter. The results are not presented here.

TABLE 8 - CALCULATED HEAT DUTIES, MMBtu/HR

<u>Run No.</u>	<u>Fort Lewis (I)</u>		<u>Fort Lewis (II)</u>		<u>ICRC</u>
SL-71	1.60	(-1.88)	2.01	(-21.9)	1.57
SL-72	2.21	(-8.14)	2.29	(-11.35)	2.03
SL-73	3.02	(-7.95)	3.01	(-7.64)	2.78
SL-41R	3.22	(-4.04)	2.85	(+8.42)	3.09
SL-42R	3.22	(-4.66)	3.02	(+1.66)	3.07
SL-43R	3.47	(-2.88)	3.29	(2.43)	3.37
SL-44R	3.55	(-3.94)	3.31	(3.02)	3.41
SL-45R	3.88	(-2.32)	3.48	(8.91)	3.79
SL-46					3.45
SL-51	1.74	(-2.87)	1.81	(-6.63)	1.69
SL-53R	2.01	(-2.99)	2.05	(-4.88)	1.95
SL-54R	2.12	(-2.36)	2.11	(-1.90)	2.07
SL-55R	2.42	(-1.65)	2.64	(-9.85)	2.38
SL-61	1.81	(-4.42)	1.85	(-6.49)	1.73
SL-62	2.16	(-4.63)	2.11	(-2.37)	2.06
SL-63	2.55	(-5.88)	2.45	(-2.04)	2.40
SL-64	2.87	(-5.57)	2.56	(+5.86)	2.71
SL-65	3.26	(-4.91)	2.96	(+4.73)	3.10
HT-11	2.73	(-4.76)	2.95	(-11.86)	2.60
HT-12	2.68	(-6.34)	2.70	(-7.04)	2.51
HT-13	2.48	(-6.05)			2.33
HT-14	2.35	(-6.38)			2.20
HT-15-1	2.72	(-4.78)			2.59
HT-15-2	2.73	(-4.76)			2.60
HT-15-3	2.74	(-5.11)			2.60
HT-15-4	2.73	(-5.13)			2.59
HT-15-5	2.88	(-5.90)			2.71
HT-15-6	2.90	(5.52)			2.74
HT-15-7	2.89	(-5.17)			2.74
HT-15-8	2.85	(-5.61)			2.69
HT-15-9	3.10	(-10.32)			2.78
HT-15-10	3.08	(-5.52)			2.90
HT-15-11	3.04	(-6.25)			2.85
HT-15-12	3.04	(-6.25)			2.85

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APPENDIX IV

Theoretical Development of Heat Transfer in a Laminar Slug Flow Regime

To aid in the scaling of results to the demonstration plant a theoretical treatment of the problem of heat transfer in a laminar slug regime was carried out by Corporate Engineering. The idealized model treated the gas and liquid slugs as symmetrical cylinders with respect to the pipe axis, with a stagnant liquid film within the gas region.

Based on the assumptions of constant slug velocities and gas/liquid ratios and similar bulk temperature profiles, it was deduced that the heat transfer coefficient 'h' can be represented by

$$h = K_1 D^{-1/2} + K_2 D^{-2/3}$$

where K_1 , K_2 are constants,

D = pipe diameter,

It was found that the factor $K_1 D^{-1/2}$ was the major contribution to the heat transfer coefficient; however, it was elected to scale-up based on $D^{-2/3}$, since this offered a more conservative approach. Further details of the model are attached.

Demonstration plant preheater model

Fig. 1

Fig. 2

D = pipe diameter, m
 δ = liquid film thickness, m
 l_g = gas slug length, m
 l_L = liquid slug length, m
 l_u = slug unit length, m

V_t = slug translational velocity, m/s

T_w = wall temperature, $^{\circ}\text{C}$

T_b = bulk temperature, $^{\circ}\text{C}$

t = time, see

x = distance from the wall, m

Q_L = slurry flowrate, m^3/s

Q_G = gas flowrate, m^3/s

V_{LS} = slurry superficial velocity, m/s

V_{GS} = gas superficial velocity, m/s

$V_{TS} = V_{LS} + V_{GS}$ = two phase superficial velocity, m/s

$t_G = \frac{L_G}{V_t}$ = period of the liquid film exposure to the gas slug, sec.

$f = \frac{V_t}{L_G}$ = slug frequency, $1/s$

μ = viscosity, $\frac{kg}{m \cdot s}$

ρ_L = slurry density, $\frac{kg}{m^3}$

C_p = slurry specific heat capacity, $\frac{cal}{kg \cdot ^\circ C}$

K = thermal conductivity, $\frac{cal}{s \cdot m \cdot ^\circ C}$

$\alpha = \frac{K}{\rho_L C_p}$ = thermal diffusivity, m^2/s

h_f = heat transfer coefficient due to liquid film contribution, $\frac{cal}{m^2 \cdot s \cdot ^\circ C}$

h_G = heat transfer coefficient in a liquid slug, $\frac{cal}{m^2 \cdot s \cdot ^\circ C}$

$h_{TOT} = h_f + \frac{L_L}{L_G} h_G$ = total equivalent heat transfer coefficient for the pipe, $\frac{cal}{m^2 \cdot s \cdot ^\circ C}$

BASIC ASSUMPTION

1. Slug flow is assumed. For gas flow, the regime is turbulent, and for liquid flow it is laminar.
2. Gas and liquid slugs are cylinders symmetrical with respect to the pipe axis.
3. There is no pressure drop along the length of the slug.
4. There is stagnant liquid film between the wall and the slug cross-section.
5. Both liquid film and liquid slug contribute to the overall heat transfer.
6. Liquid film is then in proportion to the film length; therefore heat transfer along the film is negligible.
7. There is a strong interaction between the stagnant liquid film and liquid slug; the latter has a flat velocity profile at the surfaces adjacent to the turbulent gas slugs. As a result, the liquid film is destroyed on contact with the liquid slug and the heat accumulated by the film is dissipated within the liquid slug by convection (rapid process).
8. As a result of thorough mixing between the film and liquid slug, the temperature of the film, which is left behind the liquid slug, is assumed to be equal to that of the slug.
9. The translational velocities of liquid and gas slugs are equal and both are higher than the average velocity of liquid in the slug (the sum of superficial gas and liquid velocities).

GENERAL STRATEGY

Provided the slug frequency is known, the total heat transfer coefficient for the slug unit (paired liquid-gas slugs) is sought as a sum of two weighted heat transfer coefficients; the first one is a result of the contribution of the liquid film (gas slug), and the second one is due to the developing flow entering and leaving the short pipe with a flat velocity profile (liquid slug). The stagnant liquid film is heated over a period $t_G = \frac{l_G}{V_t}$. At the time

$t = 0$ (immediately after a liquid slug passes), the entire temperature of the film equals that of a liquid slug (T_G). The maximum temperature to which the film can be heated for the time $t = \infty$ equals that of the pipe wall (T_w). At this point the amount of heat that can be accumulated by the film is at its maximum.

Within the finite period of time ($0 < t < \infty$), there is a temperature profile across the film. Assuming that there is no heat transfer from the film into the gas slug, one can calculate the amount of heat that is accumulated by the film for the period t_G of the film existence.

HEATING OF THE LIQUID FILM

For a film of thickness δ in the x direction and infinite in the y and z directions, and assuming an adiabatic condition on the face $x = \delta$, one obtains the heat transfer by conduction on the face $x = 0$ (see Fig. 2, page 1), as follows:

$$\frac{\partial T}{\partial t} = \alpha \frac{\partial^2 T}{\partial x^2} \quad (1)$$

$$\alpha = \frac{k}{\rho_L c_p}$$

The boundary and initial conditions are:

$$t = 0 \quad 0 < x < \delta \quad T = T_G \quad (2)$$

$$t > 0 \quad x = 0 \quad T = T_w \quad (3)$$

$$t > 0 \quad x = \delta \quad \frac{\partial T}{\partial x} = 0 \quad (4)$$

The solution of Equation (1) is

$$T = \sum_{n=1}^{\infty} e^{-\lambda_n^2 \alpha t} (A_n \cos \lambda_n x + B_n \sin \lambda_n x) \quad (5)$$

To satisfy the Boundary condition (3), Equation (5) should be

$$T = T_w + \sum_{n=1}^{\infty} e^{-\lambda_n^2 \alpha t} B_n \sin \lambda_n x \quad (6)$$

From the boundary condition (4), it follows:

$$\frac{\partial T}{\partial x} = \sum_{n=1, \text{ odd}}^{\infty} \lambda_n B_n e^{-\lambda_n^2 x t} \cos \lambda_n x$$

$$\left. \frac{\partial T}{\partial x} \right|_{x=\delta} = 0 \quad \text{for } \lambda_n \delta = \frac{n\pi}{2}, \quad \text{therefore } \lambda_n = \frac{n\pi}{2\delta} \quad (7)$$

For $t = 0$ (initial condition (2)), Equation (6) becomes

$$T_0 - T_w = \sum_{n=1, \text{ odd}}^{\infty} B_n \sin \lambda_n x \quad (8)$$

$$\int_0^{\delta} (T_0 - T_w) \sin \lambda_m x d\lambda_m x = \sum_{n=1, \text{ odd}}^{\infty} \int_0^{\delta} B_n \sin \lambda_n x \sin \lambda_m x d\lambda_m x \quad (9)$$

$$\text{where } \lambda_m = \frac{m\pi}{2\delta}$$

Only one term on the right-hand side of (g) survives at $m = n$ (due to orthogonality):

$$\int_0^{\delta} (T_0 - T_w) \sin \frac{n\pi x}{2\delta} d\frac{n\pi x}{2\delta} = \int_0^{\delta} B_n \sin^2 \frac{n\pi x}{2\delta} d\frac{n\pi x}{2\delta} \quad (10)$$

$$B_n = \frac{\int_0^{\delta} (T_0 - T_w) \sin \frac{n\pi x}{2\delta} d\frac{n\pi x}{2\delta}}{\int_0^{\delta} \sin^2 \frac{n\pi x}{2\delta} d\frac{n\pi x}{2\delta}} =$$

$$= \frac{(T_0 - T_w) \left[\cos \frac{n\pi x}{2\delta} \right]_0^{\delta}}{\left[\frac{1}{2} \frac{n\pi x}{2\delta} - \frac{1}{4} \sin \frac{n\pi x}{\delta} \right]_0^{\delta}} = \frac{(T_0 - T_w) \left[1 - \cos \frac{n\pi \delta}{2\delta} \right]}{\frac{1}{4} \frac{n\pi}{\delta} [\delta - 0] - \frac{1}{4} [\sin n\pi - \sin 0]} = \frac{T_0 - T_w}{\frac{n\pi}{4}} \quad (11)$$

By substituting (7) and (11) into (6) the equation for the temperature distribution can be written

$$\frac{T_w - T}{T_w - T_c} = \sum_{n=1,3,\dots}^{\infty} \frac{2}{(n\pi/2)} e^{-\left(\frac{n\pi}{2}\right)^2 \frac{\alpha t}{\delta^2}} \sin\left(\frac{n\pi}{2}\right) \frac{x}{\delta} \quad (12)$$

We are solving only for the amount of heat absorbed by the film for the time t_G ; that is, Equation (12) can be rewritten to solve for the maximum fraction of heat absorbed by the film against time:

$$\frac{T_w - T_c}{T_w - T_c} + \frac{T_c - T}{T_w - T_c} = \sum_{n=1,3,\dots}^{\infty} \frac{2}{(n\pi/2)} e^{-\left(\frac{n\pi}{2}\right)^2 \frac{\alpha t}{\delta^2}} \sin\left(\frac{n\pi}{2}\right) \frac{x}{\delta}$$

$$\frac{T - T_c}{T_w - T_c} = 1 - \sum_{n=1,3,\dots}^{\infty} \frac{2}{(n\pi/2)} e^{-\left(\frac{n\pi}{2}\right)^2 \frac{\alpha t}{\delta^2}} \sin\left(\frac{n\pi}{2}\right) \frac{x}{\delta} \quad (13)$$

Integrating (13) over the film thickness δ the expression for F can be obtained as follows:

$$F = \int_0^{\delta} \frac{T - T_c}{T_w - T_c} d\frac{x}{\delta} = 1 - \int_0^{\delta} \sum_{n=1,3,\dots}^{\infty} \frac{2}{(n\pi/2)} e^{-\left(\frac{n\pi}{2}\right)^2 \frac{\alpha t}{\delta^2}} \sin\left(\frac{n\pi}{2}\right) \frac{x}{\delta} d\frac{n\pi x}{2\delta}$$

$$= 1 - \sum_{n=1,3,\dots}^{\infty} \frac{2}{(n\pi/2)} e^{-\left(\frac{n\pi}{2}\right)^2 \frac{\alpha t}{\delta^2}} \left[\cos \frac{n\pi x}{2\delta} \right]_0^{\delta} = 1 - \sum_{n=1,3,\dots}^{\infty} \frac{2}{(n\pi/2)} e^{-\left(\frac{n\pi}{2}\right)^2 \frac{\alpha t}{\delta^2}} \quad (14)$$

The dimensionless variable $\frac{\alpha t}{\delta^2}$ is known as the Fourier number in unsteady

heat conduction problems.

The numerical summation of (14) can be performed for certain numerical values of the Fourier number, and the plot of F against $\frac{\alpha t}{\delta^2}$ can be obtained.

Because for $0 \leq F \leq 1$ the value of $\frac{\alpha t}{\delta^2}$ varied from 0 to ∞ , the appropriate

range for this value should be estimated. If one assumes the following average values for $K =$

$$= 0.07 \frac{\text{Btu}}{\text{ft} \cdot \text{hr} \cdot ^\circ\text{F}} = 0.029 \frac{\text{cal}}{\text{m} \cdot \text{s} \cdot ^\circ\text{C}}, \quad C_p = 0.6 \frac{\text{Btu}}{\text{lb} \cdot ^\circ\text{F}} = 600 \frac{\text{cal}}{\text{kg} \cdot ^\circ\text{C}}$$

$$\text{and } \rho_L = 63 \frac{\text{lb}}{\text{ft}^3} \approx 1000 \frac{\text{kg}}{\text{m}^3}, \text{ then the average value for } \alpha = 5 \cdot 10^{-8} \frac{\text{m}^2}{\text{s}}.$$

The average film thickness $\delta \cong 10 \text{ mm} = 10^{-2} \text{ m}$. Therefore if the time between 0.1 and 10 sec, the value of $\frac{\alpha t}{\delta^2}$ varies between $5 \cdot 10^{-5}$ and $5 \cdot 10^{-3}$.

In Fig. 3, the plot of F against $\frac{\alpha t}{\delta^2}$ is depicted for $10^{-5} \leq \frac{\alpha t}{\delta^2} \leq 1$. It is seen

from Fig. 3 that up to the value $\frac{\alpha t}{\delta^2} \leq 10^{-1}$, the above dependence is described

by the straight line with the slope equaling 0.5. Therefore for $t_G = \frac{l_G}{V_t} \leq 200 \text{ sec}$

for $\delta = 10^{-3} \text{ m}$, the dependence of F on $\frac{\alpha t}{\delta^2}$ can be described as follows:

$$F = \beta \left(\frac{\alpha t}{\delta^2} \right)^{0.5} \quad (15)$$

$$\text{where } \beta = 1.128 = \frac{2}{\sqrt{\pi}}$$

The amount of heat absorbed by the film for the time t therefore can be expressed as $Q_f = V_f \rho_L C_p \Delta T F$, (16)

where V_f = film volume.

Fig. 3.

HEAT TRANSFER COEFFICIENTS

The amount of heat that can be accumulated by the liquid film of length l_g and thickness δ for the period $t_g = l_g/V_g$ (see Equation (16)) is as follows:

$$Q_f = l_g \pi (D - \delta) \delta \rho_L c_p (T_w - T_c) F \quad (17)$$

The amount of heat that can be transferred through the pipe wall into the liquid slug of length l_L for the same period t_g is as follows:

$$Q_e = h_e \pi D l_L (T_w - T_c) t_g \quad (18)$$

The total equivalent heat transfer coefficient equals the total amount of heat $Q_f + Q_e$ divided by the time t_g over which this amount of heat was obtained and by the total pipe surface πD ($l_L = l_g$) through which this heat was transferred:

$$\begin{aligned} h_{tot} &= \frac{Q_f + Q_e}{t_g \pi D (l_L + l_g) (T_w - T_c)} = \frac{(Q_f + Q_e) V_g}{\pi D l_g (l_L + l_g) (T_w - T_c)} = \\ &= \frac{l_g \pi (D - \delta) \delta \rho_L c_p (T_w - T_c) F V_g + h_e \pi D l_L (T_w - T_c) \overbrace{t_g}^{= l_g/V_g} V_g}{\pi D l_g (l_L + l_g) (T_w - T_c)} = \\ &= \frac{(D - \delta) \delta \rho_L c_p V_g F}{(l_L + l_g) D} + h_e \frac{l_L}{l_L + l_g} = h_f + \frac{l_L}{l_u} h_e \quad (19) \end{aligned}$$

$$l_u = l_L + l_g = \text{length of slug unit}$$

$$F = \frac{2}{\sqrt{\pi}} \left(\frac{\alpha t_G}{\delta^2} \right)^{1/2} = \frac{2}{\sqrt{\pi}} \left(\frac{\kappa \ell_G}{c_p \rho_L V_G \delta^2} \right)^{1/2} \quad \text{see Eq (15)}$$

$$h_f = \frac{(2-\delta) \delta \rho_L c_p V_G \kappa^{1/2} \ell_G^{1/2}}{\ell_u \delta c_p^{1/2} \rho_L^{1/2} V_G^{1/2} \delta} \frac{2}{\sqrt{\pi}} = \frac{2}{\sqrt{\pi}} \frac{2-\delta}{\delta} \left(\frac{\rho_L c_p \kappa V_G \ell_G}{\ell_u} \right)^{1/2} \quad (20)$$

$$= \frac{2}{\sqrt{\pi}} (\rho_L c_p \kappa f)^{1/2} \left(\frac{\ell_G}{\ell_u} \right)^{1/2} \left(1 - \frac{\delta}{2} \right); \quad f = \frac{V_G}{\ell_u} = \text{slug frequency} \quad \frac{1}{s}$$

$$V_G = V_L = \text{translation slug velocity}$$

The combined heat transfer coefficient of the liquid slug and liquid film is as follows:

$$\frac{1}{h_{tot}} = \frac{1}{h_f} + \frac{\ell_G}{\ell_u} \frac{1}{h_b}$$

where h_f = heat transfer coefficient of the liquid slug.

For the laminar liquid slug, there are conventional equations in terms of Nusselt and Peclet numbers. We should use the one that satisfies the following conditions:

1. There is a developing velocity profile within the liquid slug.
2. Because $V_G = V_L$ the gas slugs dictate a flat velocity profile at both ends of the liquid slug. Therefore, the velocity profile is still developing from one end of the liquid slug to its midsection. This profile

develops then back to a flat profile at the other end of the slug in a symmetrical fashion. Consequently, one can treat the developing slug velocity in terms of $l_L/2$ (see Fig. 4).

3. We are interested in the averaged values of Nu or h_f over the length $l_L/2$.
4. The correlation should satisfy the condition $T_w - T_f = \text{const}$

Consider the correlation

$$Nu = 1.86 Pe^{1/3} \left(\frac{D}{x} \right)^{1/3}, \quad \text{where } Nu = \frac{h_c D}{K}$$

for $x = l_L/2$ the above become $Pe = \frac{c_p \rho_L V_{Ts} D}{K}$

$$Nu = 2.34 Pe^{1/3} \left(\frac{D}{l_L} \right)^{1/3}$$

or $h_c = 2.34 \frac{K}{D} Pe^{1/3} \left(\frac{D}{l_L} \right)^{1/3}$

Now the total heat transfer coefficient h_{tot} can be written as a sum of h_f and $\frac{l_L}{l_u} h_f$

$$h_{tot} = \frac{2}{\sqrt{\pi}} \left(\rho_L c_p K f \right)^{1/2} \left(\frac{l_c}{l_u} \right)^{1/2} \left(1 - \frac{S}{D} \right) + 2.34 \frac{K}{D} \left(\frac{c_p \rho_L V_{Ts} D}{K} \right)^{1/3} \left(\frac{D}{l_L} \right)^{1/3} \frac{l_L}{l_u}$$

$$Nu_{tot} = \frac{h_{tot} D}{K} = \frac{2}{\sqrt{\pi}} \left(\frac{P_L C_p f D^2}{K} \right)^{1/2} \left(\frac{e_g}{e_u} \right)^{1/2} \left(1 - \frac{\delta}{D} \right) + 2.34 \left(\frac{C_p P_L V_{TS} D^2}{K e_u} \right)^{1/3} \left(\frac{e_L}{e_u} \right)$$

Note that $f = \frac{V_t}{e_u} = \frac{(1 + \frac{4\delta}{D}) V_{TS}}{e_u}$; hence $V_{TS} = \frac{f e_u}{1 + \frac{4\delta}{D}}$ (21)

$$Nu_{tot} = \frac{2}{\sqrt{\pi}} P_{mod}^{1/2} \left(\frac{e_g}{e_u} \right)^{1/2} \left(1 - \frac{\delta}{D} \right) + 2.34 P_{mod}^{1/3} \left(1 + \frac{4\delta}{D} \right)^{-1/3} \left(\frac{e_L}{e_u} \right)^{2/3} \quad (22)$$

Where modified Peclet number $P_{mod} = \frac{C_p P_L f D^2}{K}$

From Fig. 5 it is seen that:

$$\frac{e_g}{e_u} = \frac{e_g}{e_g + e_L} = \frac{1}{1 + \frac{e_L}{e_g}} = \left(1 + \frac{e_L}{e_g} \right)^{-1}$$

$$\frac{e_L}{e_u} = \frac{e_L}{e_g + e_L} = \frac{1}{1 + \frac{e_g}{e_L}} = \left(1 + \frac{e_g}{e_L} \right)^{-1}$$

$$\frac{e_g}{e_L} = \frac{Q_g A_L}{A_g Q_L} = \frac{Q_g}{Q_L} \frac{D^2}{D^2 - 4D\delta} = \frac{Q_g}{Q_L} \left(1 - \frac{4\delta}{D} \right)^{-1}$$

Fig. 5.

$$\frac{e_L}{e_g} = \frac{Q_L}{Q_g} \left(1 - \frac{4\delta}{D} \right) ; \text{ here } A_L \text{ and } A_g \text{ are the pipe cross section available for liquid and gas flows, respectively}$$

$$\frac{l_c}{l_u} = \left[1 + \frac{Q_L}{Q_G} \left(1 - \frac{4\delta}{D} \right) \right]^{-1} \quad (23)$$

$$\frac{l_L}{l_u} = \left[1 + \frac{Q_G}{Q_L} \left(1 - \frac{4\delta}{D} \right)^{-1} \right]^{-1} \quad (24)$$

Substituting (23) and (24) into (22):

$$Nu_{tot} = \frac{2}{\sqrt{\pi}} Pe_{mod}^{1/2} \sqrt{\frac{\left(1 - \frac{\delta}{D} \right)^2}{1 + \frac{Q_L}{Q_G} \left(1 - \frac{4\delta}{D} \right)}} + 2.34 Pe_{mod}^{1/3} \left[\left(1 + \frac{4\delta}{D} \right) \left(1 + \frac{Q_G}{Q_L} \left(1 - \frac{4\delta}{D} \right)^{-1} \right)^2 \right]^{-1/3}$$

$$Nu_{tot} = \frac{2}{\sqrt{\pi}} Pe_{mod}^{1/2} \left[\frac{\frac{Q_G}{Q_L} \left(1 - \frac{\delta}{D} \right)^2}{\frac{Q_G}{Q_L} + 1 - \frac{4\delta}{D}} \right]^{1/2} + 2.34 Pe_{mod}^{1/3} \left[1 + \frac{4\delta}{D} + \frac{2 \frac{Q_G}{Q_L} \left(1 - \frac{4\delta}{D} \right)}{1 - \frac{4\delta}{D}} + \frac{\left(\frac{Q_G}{Q_L} \right)^2 \left(1 - \frac{4\delta}{D} \right)}{\left(1 - \frac{4\delta}{D} \right)^2} \right] \quad (25)$$

(Refer to Figure 6)

Experimental data available in the literature suggests that the film thickness depends on slug Reynolds number. If one expresses the relation between V_t and V_{ts} as

$$V_t = (1 + c) V_{ts}, \text{ then } c = 0.021 \ln \left(\frac{D V_{ts}}{\nu} \right) + 0.022,$$

which give $c = 0.16$ for $Re_{Ts} = 500$ and

$$c = 0.22 \text{ for } Re_{Ts} = 10,000$$

If one assumes $c = \frac{4\delta}{D} = 0.2$, then $\frac{\delta}{D} = 0.05$ and

$$\delta = 10^{-2} \text{ m for } D = 0.2 \text{ m}$$

Fig. 6.

For $\frac{\delta}{D} = 0.005$, Equation (25) becomes

$$Nu_{TOT} = \frac{2}{\sqrt{\pi}} Pe_{Mod}^{1/2} \left[\frac{0.9025 \frac{Q_G}{Q_L}}{0.8 + \frac{Q_G}{Q_L}} \right]^{1/2} + 2.34 Pe_{Mod}^{1/3} \left[1.2 + 3 \frac{Q_G}{Q_L} + 1.875 \left(\frac{Q_G}{Q_L} \right) \right] \quad (26)$$

For $\frac{\delta}{D} = 0.005$, Equation (25) would be

$$Nu_{TOT} = \frac{2}{\sqrt{\pi}} Pe_{Mod}^{1/2} \left[\frac{0.99 \frac{Q_G}{Q_L}}{0.98 + \frac{Q_G}{Q_L}} \right]^{1/2} + 2.34 Pe_{Mod}^{1/3} \left[1.02 + 2.08 \frac{Q_G}{Q_L} + 1.06 \left(\frac{Q_G}{Q_L} \right)^2 \right] \quad (26^*)$$

And for $\frac{\delta}{D} = 0.2$, Equation (25) gives

$$Nu_{TOT} = \frac{2}{\sqrt{\pi}} Pe_{Mod}^{1/2} \left[\frac{0.64 \frac{Q_G}{Q_L}}{0.2 + \frac{Q_G}{Q_L}} \right]^{1/2} + 2.34 Pe_{Mod}^{1/3} \left[1.8 + 18 \frac{Q_G}{Q_L} + 4.5 \left(\frac{Q_G}{Q_L} \right)^2 \right] \quad (26^{**})$$

It can be shown for $\frac{Q_G}{Q_L} > 2$ that the major contribution to the overall heat transfer is achieved by the liquid film zone.

For $\frac{Q_G}{Q_L} = 2$ and $Pe_{Mod} = 200,000$, Equations (26), (26*) and (26**) give the following:

$$Nu_{TOT} = Nu_f + Nu_c = 403 + 56 = 459 ; \quad \frac{\delta}{D} = 0.05$$

$$Nu_{TOT} = Nu_f + Nu_c = 411 + 65 = 476 ; \quad \frac{\delta}{D} = 0.005$$

$$Nu_{TOT} = Nu_f + Nu_c = 385 + 27 = 412 ; \quad \frac{\delta}{D} = 0.2$$

It is also seen from the above that the heat transfer rate is relatively insensitive to the liquid film thickness δ . This fact follows from the assumption in the model, that there is thorough mixing between the film and the liquid slug (complete surface renewal). In reality, the film thickness is unlikely to vary over such a wide range (the experimental value of $C = \frac{4\delta}{D} = 0.16 - 0.22$).

D

It is seen from (26) that the larger the ratio $\frac{Q_G}{Q_L}$, the larger the contribution of the first term of Equation (26) and the less sensitive it becomes to the liquid film thickness. The same can be said about the Peclet number in (26).

All the parameters in Equation (26) are easy to determine except for the slug frequency f , which determines the Peclet number

$$Pe_{ind} = \frac{c_p \rho_L f D^2}{K}$$

The variation of the experimentally determined values of the heat transfer coefficients h_{exp} along the pipe (Ft. Lewis Pilot Plant) could possibly be explained by the variation in slug frequency as a function of viscosity. The experimental pressure drop plots show that the viscosity increases along the pipe, reaches its maximum, and then decreases again. If the increase in viscosity results in a decrease in the slug frequency, then this could explain the continuous decrease of h_{exp} until the middle of the pipe which has a temperature range of 500-550°F (coal gel region - highest viscosity). From that point h_{exp} rapidly increases until the end of the pipe. The higher rate of increase of h_{exp} in the second half of the pipe and its slower rate of decrease, in the first half could be explained by the additional influence of higher gas flow rates, which increase the slug frequency. The third possible reason for the high heat transfer rate in the second part of the pipe is the formation of gas bubbles within the liquid slugs, which tend to break up the liquid slugs, and hence increase their infrequency. The opposite phenomenon could exist at the pipe entrance zone where there is a tendency for the relatively short and unstable

transfer rate. If one accepts the hypothesis that the ratio $\frac{l_u}{D}$, $\frac{Q_G}{Q_L}$ and

the temperature profile along the pipe are maintained constant on scale-up, then for the same liquid-gas system, the modified Peclet number could be replaced with the conventional one

$$Pe_{mod} = \frac{c_u}{D} = \frac{c_p \rho_L f D^2 c_u}{K D} = \frac{c_p \rho_L V_t D^2 c_u}{K D c_u} = \frac{c_p \rho_L V_t D}{K}$$

Due to the approximate relation (Equation (26))

$$Nu_{Tot} \propto Pe_{mod}^{1/2} \quad \text{or} \quad \frac{h D}{K} \propto \left(\frac{c_p \rho_L V_t D}{K} \right)^{1/2}$$

$$h \propto \left(\frac{V_t}{D} \right)^{1/2}$$

for $\frac{l_u}{D}$ and $\frac{Q_G}{Q_L}$ constant; the second term of (26) was shown to be negligible.

Ft. Lewis Plant

$$(D = 1.689' = 0.0429m, L/D = 3310)$$

$$C_p = \begin{cases} 0.53 \text{ Btu/lb}^\circ\text{F} & 300 \leq T \leq 600^\circ\text{F} \\ 0.74 & T > 600^\circ\text{F} \end{cases}$$

Run # SL-45R

Run	T, °F	h, $\frac{\text{Btu}}{\text{ft}^2 \text{ hr}^\circ\text{F}}$	h, $\frac{\text{cal}}{\text{m}^2 \text{ s}^\circ\text{C}}$	Nu = $\frac{hD}{K}$	Pe _{mod} = $\frac{C_p \rho_L f D^2}{K}$	ρ_L , $\frac{\text{kg}}{\text{m}^3}$	f, 1/s	$\frac{L_u}{D}$	V _{GS} , m/s	V _{LS} , m/s	V _{TS} , m/s	V _{ET} , m/s	$\frac{Q_c}{Q_L}$	ΔP , PSI	$\frac{L}{D}$	C _p , $\frac{\text{cal}}{\text{g}^\circ\text{C}}$	K, $\frac{\text{cal}}{\text{m s}^\circ\text{C}}$	t _e , sec
-3	415	229	311	460	180,000	1027	5.2	35.7	5.63	0.95	6.84	7.97	5.99	10	4.2	530	0.029	0.17
3-5	480	192	260	386	130,000	1002	3.9	50.8	6.11	0.97	7.08	8.50	6.30	11	5.7	530	0.029	0.23
5-7	540	162	220	326	90,000	972	2.8	75.3	6.55	0.98	7.53	9.04	6.68	18	8.1	530	0.029	0.32
7-9	610	310	420	623	320,000	938	7.3	30.7	7.03	0.99	8.02	9.62	7.10	13	3.1	740	0.029	0.12
9-11	670	304	412	611	310,000	911	7.2	33.1	7.53	0.98	8.51	10.21	7.68	16	3.1	740	0.029	0.13
11-13	710	338	458	680	400,000	905	9.4	26.8	8.08	0.93	9.01	10.81	8.69	14	2.2	740	0.029	0.1

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Run # SL-43R

1-3	410	184	249	368	130,000	1031	3.7	27.5	2.70	0.94	3.64	4.37	2.87	5	6.0	530	0.029	0.21
3-5	475	180	244	361	130,000	1002	3.9	27.9	2.91	0.98	3.89	4.67	2.97	6	5.9	530	"	0.20
5-7	535	123	167	247	60,000	973	1.8	63.5	3.09	0.99	4.08	4.90	3.12	8	13.0	530	"	0.44
7-9	605	155	210	311	90,000	934	2.1	57.4	3.23	1.02	4.31	5.17	3.23	15.5	11.4	740	"	0.38
9-11	660	286	388	574	330,000	913	7.7	16.5	3.50	1.04	4.54	5.45	3.37	15	3.2	740	"	0.10
11-13	710	446	605	895	800,000	905	18.8	7.1	3.76	1.0	4.76	5.71	3.76	12	1.3	740	"	0.04

Ft. Lewis Plant

Run # SL-42 R

Run	T, °F	h_i , $\frac{Btu}{ft^2 \cdot hr \cdot F}$	h_o , $\frac{cal}{m^2 \cdot s \cdot ^\circ C}$	$Nu = \frac{hD}{k}$	$Pe = \frac{\rho u c_p D}{k}$	ρ_L , $\frac{kg}{m^3}$	f , $\frac{1}{s}$	$\frac{C_u}{D}$	V_{Gs} , $\frac{m}{s}$	V_{Ls} , $\frac{m}{s}$	V_{Ts} , $\frac{m}{s}$	V_t , $\frac{m}{s}$	$\frac{Q_G}{Q_L}$	ΔP , PST	$\frac{e_L}{D}$	C_p , $\frac{cal}{kg \cdot ^\circ C}$	K , $\frac{cal}{m \cdot s \cdot ^\circ C}$	t_G , sec
1-3	410	116	157	232	54,000	1032	1.6	39.9	1.34	0.94	2.28	2.74	1.43	6.5	14.3	530	0.029	0.4
3-5	470	130	176	260	70,000	1004	2.1	31.7	1.42	0.96	2.38	2.86	1.48	2	11.1	530	"	0.3
5-7	540	103	140	207	42,000	974	1.3	53.4	1.49	0.99	2.48	2.98	1.51	3.5	18.5	530	"	0.5
7-9	600	74	100	148	20,000	944	0.5	145	1.57	1.02	2.59	3.11	1.54	6	49.6	635	"	1.3
9-11	655	153	207	306	100,000	915	2.3	32.7	1.65	1.04	2.69	3.23	1.59	14	10.9	740	"	0.3
11-13	705	324	439	649	460,000	905	10.8	20.2	1.75	1.04	2.79	3.35	1.68	13.5	6.5	740	"	0.2

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Run # SL-41 R

1-3	410	110	149	220	48,000	1032	1.4	31.1	0.62	0.94	1.56	1.87	0.66	3.5	17.0	530	0.029	0.3
3-5	470	74	100	148	20,000	1004	0.6	75.4	0.64	0.98	1.62	1.94	0.65	3	41.6	530	"	0.7
5-7	530	51	69	102	8,500	976	0.3	155.4	0.67	1.0	1.67	2.00	0.67	6	84.6	530	"	1.5
7-9	600	79	107	158	24,000	945	0.6	80.8	0.69	1.04	1.73	2.08	0.66	9	44.3	635	"	0.8
9-11	650	166	225	333	130,000	915	3.0	16.5	0.70	1.07	1.77	2.12	0.65	21	9.1	740	"	0.15
11-13	705	264	358	530	320,000	905	7.5	6.7	0.73	1.07	1.80	2.16	0.68	15	3.6	740	"	0.06

Ft. Lewis Plant

Run # SL-73 (The slurry only)

turn	T, °F	ΔP, PSI	ℓ, m	$\frac{D}{ℓ}$	h , $\frac{\text{cal}}{\text{m}^2 \text{s}^\circ \text{C}}$	h_{exp} $\frac{\text{cal}}{\text{m}^2 \text{s}^\circ \text{C}}$	h_{exp} $\frac{\text{Btu}}{\text{ft}^2 \text{hr}^\circ \text{F}}$	Nu _{exp}	$Pe_{\text{mod}} =$ $\frac{c_p \rho_L f D^2}{K}$	V _{LS} , m/s	ρ _L , $\frac{\text{kg}}{\text{m}^3}$	f, 1/s	$\frac{ℓ_u}{D}$	Pe = $\frac{c_p \rho_L V_{LS} D}{K}$
1-3	410	3	32	$1.34 \cdot 10^{-3}$	15.9	43	32	64	3200	0.94	1028	0.09	236	757,628
3-5	465	1	54	$7.94 \cdot 10^{-4}$	13.4	54	40	80	5400	0.98	1004	0.16	143	771,427
162 5-7	525	8	76	$5.64 \cdot 10^{-4}$	11.9	44	33	65	3300	0.99	980	0.10	231	760,670
7-9	575	19	98	$4.38 \cdot 10^{-4}$	10.9	52	38	77	4800	1.0	959	0.15	155	751,889
9-11	620	22	120	$3.58 \cdot 10^{-4}$	11.5	89	66	132	13000	1.05	932	0.30	82	1,071,263
11-13	675	13	142	$3.02 \cdot 10^{-4}$	10.8	318	235	470	250000	1.07	909	5.86	4.3	1,064,728

Laminar regime

$$Nu = 2.34 Pe^{1/3} \left(\frac{D}{ℓ} \right)^{1/3}$$

$$h = 2.34 \frac{K}{D} Pe^{1/3} \left(\frac{D}{ℓ} \right)^{1/3}$$

$$Pe = \frac{c_p \rho_L V_{LS} D}{K}$$

APPENDIX V

Radioactive Tracer Tests

The tracer studies carried out at Fort Lewis during the heater tests were previously analyzed and reported. It was concluded, that owing to the high solubility of the gas tracer in the slurry phase, results obtained are very much questionable. The attached outlines the analysis procedure.

Date 7 July 1981

INTEROFFICE
MEMORANDUM

Subject Fort Lewis Tracer Tests

To R. M. Thorogood

From B. O. Brown

ICRC

(ORGANIZATION OR DEPARTMENT)

(ORGANIZATION OR DEPARTMENT)

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Introduction

In conjunction with the ICRC Fired Heater Program in June-August 1980, at Fort Lewis, Washington, twelve radioactive tracer tests were run. These tests served the purpose of measuring the individual phase residence times and thus average phase holdups within the slurry preheater. Tests conditions were varied so as to duplicate those within the High Temperature, SL-40 and SL-50 series of the heater program.

Radiation detectors placed near the inlet and outlet of the furnace monitored the passage of tracer material. The time elapsed between the measured signals gave the respective phase residence time.

The measurement technique involved the rapid injection of a radioactive material into the process stream. Liquid phase measurements were carried out using radioactive bromine in bromophenanthrene. Argon-41 was used for the gas phase.

Equipment and Test Description

Two sodium iodide scintillation detectors were located on the process lines to monitor the flow of the tracer material. One was placed 2 feet upstream of the heater inlet, the other 32 feet from the outlet. The signal from each was electronically processed and finally fed to a dual pen chart recorder and multichannel pulse height analyzer for analog and digital recording. Fig. 1 outlines the instrumentation system.

Injection of the radioactive tracers into the process stream was accomplished by means of pressurized nitrogen. Both tracers (Bromophenanthrene and Argon-41) were separately irradiated in advance in a nuclear reactor at the University of Washington. The injection system for each tracer is shown in Fig. 2.

Gas and liquid rates were varied so as to encompass those of the heater series. With the plant lined out at each test condition, separate tracer injection and measurements were carried out for each phase.

Results

Residence Time - A typical response plot is shown in Fig. 3. Phase residence times were previously calculated and reported⁽¹⁾, the results of which are listed in Table 1. These may be summarized as follows:

- 1) Gas phase residence time varied from 0.57-0.88 mins.;
liquid phase from 1.19-2.18 mins.
- 2) Increasing gas rate at a constant liquid rate decreases both gas and liquid residence times.
- 3) Increasing liquid rate at a constant gas rate tends to decrease liquid residence time, however, negligible effect is observed on gas residence time.

Phase Holdup - The respective phase residence time can be defined by the following

$$\text{For Gas} \quad T_G = \frac{L \times A \times \bar{R}_G}{\bar{Q}_G}$$

$$\text{For Liquid} \quad T_L = \frac{L \times A \times \bar{R}_L}{\bar{Q}_L}$$

Where L = pipe length
A = flow cross section area

$\bar{R}_L, \bar{R}_G$ - Average liquid and gas holdup

$\bar{Q}_L, \bar{Q}_G$ - " " " " flowrates

Rearranging and with $\bar{R}_G = 1 - \bar{R}_L$, we obtain

$$R_L = \frac{T_L \bar{Q}_L}{T_L \bar{Q}_L + T_G \bar{Q}_G}$$

With the use of Cycsyn integrated average phase flow rates were calculated for the heater. The respective run holdups are listed in Table 2. By using data from the most similar run of the pre-heater test sequence, holdups predicted from Hughmarks correlations (2, 3) were obtained. These are also included in Table 2. Generally the predicted holdups from Hughmarks slug flow correlation are higher than those of the other two methods.

Slug Frequency - The outlet response curves for the liquid phase injections, show the occurrence of small (somewhat regularly spaced) peaks superimposed upon the surface of the larger response curve (e.g., Fig. 1). Due to the smooth injection indicated by the inlet curve, it may be speculated that these are due to the presence of a slug flow regime within the heater. If this is true then the slug frequency ν may be determined as

$$\nu = 1/t$$

where t = average time between adjacent small peaks. As per the theory of Hubbard and Dukler (1975), the length of a slug unit (liquid and gas bubble) can be found from

$$L = V_s (1 + C)/\nu$$

Where V_s = no slip homogeneous combined phase velocity

C = ratio of rate of mass pickup to rate of mass flow in the slug = 0.0625

ν = frequency

Values for each run obtained by the two above equations are listed in Table 3. Runs where peaks are not easily distinguished are omitted. From the results shown, a reduction in frequency is observed for either an increase in gas rate or a decrease in liquid rate. Values (at inlet conditions) as predicted by the methods of Taitel and Dukler (1977) and Greskovich and Shrier (1973) are compared in Table 4 with the average of the values listed in Table 3. Although both methods have not been tested near plant conditions (high gas density, high liquid viscosity), that of Taitel and Dukler shows the better agreement.

Discussion

Phase Residence Time - In the determination of residence time distribution within a single phase system it is important that the tracer obey the following requirements:

- 1) Is inert to the system being studied and is affected by only the fluid hydrodynamics present,
- 2) Be distinct (dye, radioactive, etc.) from the system so that it can be monitored separately,

with the presence of a second phase a third requirement is

- 3) That the tracer remains within the phase being studied.

In order to verify the accuracy of the latter assumption for the Fort Lewis Tracer Tests, "Cycsyn" was used to determine equilibrium "K" values for the tracer materials. Values for Argon-41 are shown in Fig. 4. These indicate the reverse solubility property of Argon and the occurrence of "K" values less than one over a major portion of the temperature range in the heater. K is defined as

$$K = \frac{\text{lb moles of Ar in Vapor}}{\text{lb moles of Ar in Slurry}} \times S$$

$$\text{Where } S = \frac{\text{Total lb moles of Slurry}}{\text{Total lb moles of Vapor}}$$

Where "S" having values of 0.2-1.0 (staying approximately constant within a run) and K varying as shown, a major portion of the tracer should be within the liquid phase for most runs. The system could thus show considerable retention of the tracer in the liquid phase thus resulting in the measurement of a gas residence time larger than the correct value.

Due to a deficiency in the thermodynamic system used in Cycsyn Bromophenanthrene could not be modelled directly. However, due to its M.W. (256.9) and B.P. (680°F) it is expected to behave similarly to the 650-700°F cut of the SBC-I solvent. The latter exhibits "K" values varying from 7.4×10^{-5} - 3.3×10^{-2} over the temperature ranges studied. Thus the liquid residence time measurements are considered reliable.

Presence of Slug Flow - The present understanding of slug flow propagation is that as the slug progresses down a tube it picks up liquid at its front and sheds it at an equal rate at its rear. An idealized picture is shown in Fig. 5. It is thus conceivable that the small peaks as observed in Fig. 1 are due to the passage of alternate gas and liquid slugs. However, it must be realized that these peaks might also be controlled by the coil geometry outside of the heater. This becomes apparent when one reads the following excerpt taken from a Fort Lewis Report(7).

"The coil outlet flange is about 10 ins. from the outside furnace wall. The tubing size then increases to 3 ins. Sch 160. Note that there is an 8 ft. vertical drop in this tubing about 10 ft. downstream of the coil outlet. The rest of the 32 ft. of outlet tubing is horizontal. It was necessary to place this detector some distance from the furnace and the inlet injection point to prevent large background effects."

(Note: the outlet detector is placed at the end of the 32 ft.)

Conclusion

The above analysis has shown cause for concern about the accuracy of the measured gas phase residence times. Predicted "K" values would indicate that the actual gas residence times could be much smaller than measured. Calculated liquid holdups from the measured residence times showed good agreement with those predicted from Hughmarks correlation (2). If the actual residence time of the gas is small in comparison to that for appreciable transfer of Argon to the liquid phase, then the two could be reconciled. However, since Cycsyn (which assumes equilibrium) was used in generating flow rates for both methods, it is recommended that the results presented herein be treated with caution until the accuracy of "Cycsyn" can be verified.

B. C. Brown

Attachments

- (1) Thew, T.W., "ICRC Fired Heater Test Program", memo to distribution, Dec. 1, 1980.
- (2) Hughmark, G. A., "Holdup in Gas-Liquid Flow", Chemical Engineering Progress, 58, 1962.
- (3) Hughmark, G. C., "Holdup and Heat Transfer in Horizontal Slug Gas-Liquid Flow", Chemical Eng. Science, Vol. 20, 1965.
- (4) Taitel, Y., and Dukler, A.E., "A Model for Slug Frequency during Gas-Liquid Flow in Horizontal and Near Horizontal Pipes", Int. J. Multiphase Flow, Vol. 3, 1977.
- (5) Greskovich, E. J., and Shrier, A. L., "Slug Frequency in Horizontal Gas-Liquid Slug Flow", Ind. Eng. Chem. Process Des. Develop., Vol. 11, 1972.
- (6) Dukler, A. E., and Hubbard, M. G., "A Model for Gas-Liquid Slug Flow in Horizontal and Near Horizontal Tubes", Ind. Eng. Chem. Fundamentals, 14, 1975.
- (7) Ackerman, C.D., "Radiotracer Studies of SRC-I Coal Slurries through Preheater B at the Fort Lewis SRC Pilot Plant", Fort Lewis Report, March 1981.

FIGURE 1
ELECTRONIC EQUIPMENT
AND SCHEMATIC OF TRACER FLOW

LIQUID-SLURRY APPARATUS

GAS INJECTION APPARATUS

Figure 2. Injection Apparatus used in Radiotracer Tests.

Figure 3. Typical Response Curve

Figure 4. Argon Equilibrium Distribution Constant
vs. Temperature

40 0493

10-2 EQUILIBRIUM DISTRIBUTION CONSTANT

$$K = \frac{y}{x}$$

Figure 5. Slug Flow Model

Table 1
CALCULATED PHASE RESIDENCE TIMES

<u>Tracer Test No.</u>	<u>V*</u> <u>lb/hr</u>	<u>L*</u>	<u>Residence Times</u> <u>Min.</u>
RT-1	227	9080	V = 0.88
RT-2	224	9070	L = 1.83
RT-3	574	9050	V = 0.57 L = 1.23
RT-4	569	9070	L = 1.24
RT-5	537	10960	V = 0.57 L = 1.20
RT-6	539	10960	L = 1.19
RT-7	269	10970	V = 0.77 L = 1.49
RT-8	270	10960	L = 1.50
RT-9	264	5600	L = 2.18
RT-10	266	5550	V = 0.79
RT-11	532	5490	V = 0.51
RT-12	534	5530	L = 1.62

*V = Vapor
*L = Slurry

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TABLE 2
LIQUID PHASE HOLDUPS

<u>Test No.</u>	<u>Hughmark^{(1)*}</u>	<u>Hughmark^{(2)*}</u>	<u>Tracer Residence Time Date</u>
RT 1 & 2	0.62	0.72	0.61
RT-3 & 4	--	--	--
RT-5 & 6	0.42	0.60	0.40
RT-7 & 8	0.57	0.70	0.56
RT-9 & 10	0.45	--	0.46
RT-11 & 12	0.33	0.54	0.32

*Hughmark (1) - generalized correlation (Ref. 2)

(2) - slug flow correlation (Ref. 3)

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TABLE 3
SLUG FREQUENCY AND LENGTH

<u>Run</u>	<u>ν (Sec⁻¹)</u>	<u>Slug Length (ft.)</u>	<u>lb/hr</u>	
			<u>V</u>	<u>L</u>
RT-2	0.44	16.7	224	9070
	0.44	16.7		
	0.67	11.1		
	0.51	14.4		
	0.56	13.3		
RT-4	0.30	44.2	569	9070
	0.30	44.2		
	0.83	16.0		
	0.33	40.2		
RT-9	0.28	25.3	264	5600
	0.48	14.3		
	0.21	33.7		
	0.33	21.1		
RT-12	0.33	38.2	534	5530
	0.20	63.0		

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TABLE 4
AVERAGE SLUG FREQUENCIES (SEC⁻¹)

<u>Run No.</u>	<u>Tracer Tests</u>	<u>Dukler</u>	<u>Greskovich</u>
RT-2	0.52	0.45	1.2
RT-4	0.44	--	--
RT-9	0.33	0.22	0.6
RT-12	0.27	0.27	0.3

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APPENDIX VI

Computer Program Listing

The following is a complete listing of the computer program used in the Fort Lewis Fired-Heater Tests Analysis. Input data and calculation procedures are described by comment cards. The sections on Heat Transfer and Vertical Two-Phase Flow were not used in the analysis.

COMPILER OPTIONS - NAME= MAIN,OPT=02,LINECHT=60,SIZE=0000K,
SOURCE,EBCDIC,NOLIST,MODECK,LOAD,MAP,NOEDIT,NOID,XREF

```

ISN 0002      DIMENSION TITLE(10)
ISN 0003      REAL M,L,K,KV,JG,JL,NLB,NLS,NLM,NGV
ISN 0004      NAMELIST /DATA/AK,CP,E,SUFTEEN,V,L,DENVAP,DENLIQ,VISGAS,
              1DIAM,AREA,ALPHA,P,TITLE,DPDLME
C*****
C***** E=PIPE ROUGHNESS,FT
C***** AK=THERMAL CONDUCTIVITY, BTU/HR-FT-F
C***** CP=HEAT CAPACITY, BTU/LB-F
C***** V,L=VAPOR AND LIQUID RATES, LB/HR
C***** DENVAP,DENLIQ=VAPOR AND LIQUID DENSITIES, LB/CU.FT
C***** DPDLME=MEASURED PRESSURE GRADIENT, PSI/1000 FT
C***** DIAM,AREA=PIPE DIAMETER AND FLOW AREA, FT; SQ.FT
C***** SUFTEEN=SURFACE TENSION, DYNES/CM
C***** VISGAS= GAS VISCOSITY, LB/FT-HR
ISN 0005      20 READ(5,DATA,END=580)
ISN 0006      WRITE(6,10) TITLE
ISN 0007      10 FORMAT(1H120X,10A4)
ISN 0008      WRITE(6,1001)
ISN 0009      WRITE(6,DATA)
ISN 0010      WRITE(6,1001)
C***** CALCULATION OF PRESSURE GRADIENTS
ISN 0011      II=0
C*****
C***** INITIAL GUESS FOR VISCOSITY, 500 LB/FT HR
ISN 0012      VISLIQ=500.0
ISN 0013      1 II=II+1
ISN 0014      IF(VISLIQ.LT.0.1) VISLIQ=0.1
C***** DETERMINE REYNOLDS NUMBERS
ISN 0016      REV=(V/AREA)*DIAM/VISGAS
ISN 0017      REL=(L/AREA)*DIAM/VISLIQ
C***** CALL SUBROUTINE TO DETERMINE FRICTION FACTORS
ISN 0018      CALL FRFA(FL,REL,DIAM,E,M)
ISN 0019      CALL FRFA(FV,REV,DIAM,E,M)
C*****
C***** CALCULATION OF MARTINELLI PARAMETER X
ISN 0020      X=(L/V)*(((FL*DENVAP)/(FV*DENLIQ))*0.5)
C*****
C***** CALCULATION OF PRESSURE GRADIENTS FOR GAS AND LIQUID
ISN 0021      DPDXV=(144*FV*((V/AREA)**2.))/(3.0E+10*DENVAP*DIAM)
ISN 0022      DPDXL=(144*FL*((L/AREA)**2.))/(3.0E+10*DENLIQ*DIAM)
ISN 0023      GT1=(V+L)/AREA
ISN 0024      QGAS=V/DENVAP
ISN 0025      QLIQ=L/DENLIQ
ISN 0026      VGAS=QGAS/(AREA*3600.)
ISN 0027      VLIQ=QLIQ/(AREA*3600.)
C*****
C***** CALCULATION OF PRESSURE GRADIENT VIA DUKLER CONSTANT-SLIP MODEL
C***** DETERMINING CONSTANT CONSTANT USED IN HUGHMARK HOLDUP CORRELATION
ISN 0028      VELVL=VGAS+VLIQ
ISN 0029      C2=0.642*SQRT(VELVL)*(GT1**0.1667)*(DIAM**0.0417)/(VLIQ**0.25)
ISN 0030      ETA=VLIQ/(VLIQ+VGAS)
ISN 0031      DENNS= ETA*DENLIQ+DENVAP*(1.0-ETA)
ISN 0032      VISNS= ETA*VISLIQ+VISGAS*(1.0-ETA)
ISN 0033      RELNS=((V+L)/AREA)*DIAM/VISNS
ISN 0034      CALL HOLDUP(HHU,VISGAS,VISLIQ,ETA,C2)
C*****

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      C***** CALCULATE ACCELERATION TERM BETA
      BETA=((DENLIQ)*(ETA**2.0)/HHU+((1.0-ETA)**2.0*DENVAP/(1.0-HHU)))/D
      LENNS
      C*****
      C***** DETERMINE REYNOLDS NUMBER
      RETP=BETA*RELS
      CALL FRFA(FO,RETP,DIAM,E,M)
      C*****
      C***** EVALUATE FRICTION FACTOR NORMALISER FROM EXPRESSION WHICH IS A
      C***** FUNCTION OF ETA, THE LIQUID VOLUME FRACTION.
      Z=ALOG(ETA)
      Z2=Z*Z
      Z3=Z2*Z
      Z4=Z2*Z2
      ZETA=1.281+0.478*Z+0.444*Z2+0.094*Z3+0.00843*Z4
      C*****
      C***** FRICTION FACTOR NORMALISER ALPH
      ALPH=1.0-(Z/ZETA)
      C*****
      C***** TWO PHASE FRICTION FACTOR
      FTP=FO*ALPH
      C*****
      C***** TWO PHASE PRESSURE GRADIENT
      DUKCS1=FTP*4.0*BETA*((V+L)**2.0)/(DENNS*DIAM*(AREA**2.0)*144.0*(36
      100**2.0)*64.34)
      DENCSP=BETA*DENNS
      DUKCS=DUKCS1*1000.0-DENCSP*(SIN(ALPHA*3.141593/180.0))/0.144
      DIF=DPDLME-DUKCS
      IF(11.GE.2) GO TO 2
      DIFO=DIF
      VISLQ=VISLIQ
      VISLIQ=VISLQ-200.0
      GO TO 1
      2 IF(11.GT.50) GO TO 3
      C***** IS VISCOSITY ITERATION CONVERGED
      VDIF=ABS(VISLIQ-VISLQ)
      IF(VDIF-5.0) 3,3,4
      C***** DETERMINE NEW ESTIMATE FOR VISCOSITY
      4 SL=(DIFO-DIF)/(VISLQ-VISLIQ)
      VISLQ=VISLIQ
      DIFO=DIF
      VISLIQ=VISLIQ-DIF/SL
      GO TO 1
      3 IF(11.GT.50) WRITE(6,7)
      7 FORMAT(1X,'+++++ DUKLER LIQUID VISCOSITY NOT CONVERGED +++++
      1++++')
      WRITE(6,480)
      480 FORMAT(///,15X,'TWO-PHASE PRESSURE GRADIENT PREDICTIONS')
      WRITE(6,1001)
      WRITE(6,1001)
      WRITE(6,2000) VISLIQ
      2000 FORMAT(///,' LIQUID VISCOSITY PREDICTED BY DUKLER CONSTANT SLIP CO
      RRRELATION = ',E10.4,' LB/FT-HR')
      WRITE(6,690)RETP,DENCSP,HHU,DUKCS,DPDLME
      690 FORMAT(///,' BY DUKLER CONSTANT SLIP CORRELATION',/,10X,'REL=',E10.
      14/,10X,'DENSITY=',E10.4,' LB/CU.FT.',/,10X,'LIQUID HOLDUP=',E10.4/,
      210X,'PRESSURE GRADIENT=',E10.4,' PSI/1000 FT',/,10X,'MEASURED PRESS
      SURE GRADIENT=',E10.4,' PSI/1000 FT')

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C*****
C***** CALCULATION VIA LOCKHART MARTINELLI
ISN 0075 IF(REL.GT.2000.0) GO TO 450
ISN 0077 IF(REV.LE.2000.0) C1=5.0
ISN 0079 IF(REV.GT.2000.0) C1=12.0
ISN 0081 GO TO 470
ISN 0082 450 IF(REV.LE.2000.0) C1=10.0
ISN 0084 IF(REV.GT.2000.0) C1=20.0
ISN 0086 470 CONTINUE
ISN 0087 Q2V=1.0+(C1*X)+(X**2.0)
ISN 0088 DPTP1=((Q2V*DPDXV)/0.144)
ISN 0089 XLN=ALOG(X)
ISN 0090 XLN2=XLN*XLN
ISN 0091 XLN3=XLN2*XLN
ISN 0092 RL= 1.00000000-EXP((-0.883826E-02)*(XLN3)-(0.289274E-01)*(XLN2)-(0
1.105733)*(XLN)-0.255216)
RL1=RL
ISN 0093 ETA=VLIQ/(VLIQ+VGAS)
ISN 0094 IF(RL.LT.ETA) RL1=ETA
ISN 0095 DENCSP=DENLIQ*(ETA**2.0)/RL1+DENVAP*(1.0-ETA)**2.0/(1.0-RL1)
ISN 0097 DPTP=DPTP1-(DENCSP*SIN(ALPHA*3.141593/180.0))/0.144
ISN 0098 WRITE(6,490) DPDLME,DPTP,VISLIQ,REL,REV,ETA,RL,X
ISN 0100 490 FORMAT(//////,' BY LOCKHART-MARTINELLI CORRELATION',10X,'MEASURED
1 PRESSURE GRADIENT=',E10.4,' PSI/1000 FT',10X,'CALCULATED PRESSURE
2 GRADIENT=',E10.4,' PSI/1000 FT',10X,'CALCULATED LIQUID VISCOSITY
3 = ',E10.4,' LB/FT HR',10X,'RE=',E10.4/10X,'REV=',E10.4/10X,'L
4 LIQUID VOL. FRACT.=',E10.4/10X,'LIQUID HOLDUP=',E10.4/10X,'X=',E1
50.4)

C*****
C***** CALCULATION OF PRESSURE GRADIENT VIA DUKLER NO-SLIP MODEL
ISN 0101 DENNS=ETA*DENLIQ + DENVAP*(1.0-ETA)
ISN 0102 VISNS=ETA*VISLIQ + VISGAS*(1.0-ETA)
ISN 0103 RELNS=((V+L)/AREA)*DIAM/VISNS
ISN 0104 CALL FRFA(F,RELNS,DIAM,E,M)
ISN 0105 DUKNS1=(4.0*F*(V+L)**2.0)*1000.0/(DENNS*DIAM*AREA*AREA*3600*3600*1
144*64.34)
ISN 0106 DUKNS=DUKNS1-(SIN(ALPHA*3.141593/180.0))*DENNS/0.144
ISN 0107 WRITE(6,590)RELNS,DENNS,DUKNS
ISN 0108 590 FORMAT(//////,' BY DUKLER NO SLIP CORRELATION',10X,'RELNS=',E10.4/1
10X,'DENSITY=',E10.4,' LB/CU.FT.',10X,'PRESSURE GRADIENT=',E10.4,'
2 PSI/1000 FT')

C*****
C***** OPTIONAL VERTICAL PRESSURE DROP CALCULATION BY ORZIEWSKI
ISN 0109 IF(ALPHA+90.0) 890,890,440
ISN 0110 890 IF(Z.EQ.1.0) GO TO 894
ISN 0112 IF(Z-3.0) 990,990,991
ISN 0113 894 CONTINUE

C*****
C***** CALCULATING LIQUID HOLDUP ROR
ISN 0114 ROR=0.5-0.625*VELVL+((0.5+0.625*VELVL)**2.0-(1.25*VGAS))**0.5

C*****
C***** REYNOLDS NUMBER
ISN 0115 DENTPB=ROR*DENLIQ+(1.0-ROR)*DENVAP
ISN 0116 REOB=DENLIQ*DIAM*VLIQ*3600/(VISLIQ*ROR)
ISN 0117 CALL FRFA(FIP,REOB,DIAM,E,M)
ISN 0118 DPORKB=(4*FIP*DENLIQ*((VLIQ/ROR)**2.0)/(2.0*32.2*DIAM)+DENTPB)*100
10.0/144
ISN 0119 WRITE(6,895)ROR,DENTPB,DPORKB

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ISN 0120      895 FORMAT(///,' BY ORKISZEWSKI CORRELATION FOR VERTICAL FLOW',45X,'L
              1 LIQUID HOLDUP=',E10.4/,45X,' TWO PHASE DENSITY=',E10.4,' LB/CU.FT.'/
              2,45X,' PRESSURE GRADIENT FOR BUBBLE FLOW REGIME=',E10.4,' PSI/1000
              3 FT')
ISN 0121      GO TO 440
              C*****
              C***** CALCULATION OF SLUG FLOW PARAMETERS
ISN 0122      990 EN1=(0.5721E+05)*(-0.35+(0.1225+0.04931*VELVL/(DIAM**0.5)))
ISN 0123      EN2=(0.5721E+05)*(-0.546+(0.2981+0.01849*VELVL/(DIAM**0.5)))
ISN 0124      REOS=DENLIQ*DIAM*VELVL*3600.0/VISLIQ
ISN 0125      IF(REOS.GT.EN1) VR=(1.985+4.958E-05*REOS)*(DIAM**0.5)
ISN 0127      IF(REOS.LE.EN2) VR=(3.097+4.958E-05*REOS)*(DIAM**0.5)
ISN 0129      IF(REOS.LE.EN1.AND.REOS.GT.EN2) GO TO 25
ISN 0131      GO TO 29
ISN 0132      25 GAMMA=(1.423+(4.958E-05*REOS))*(DIAM**0.5)
ISN 0133      VR=0.5*(GAMMA+(GAMMA**2.0+(13.59/2.42)*VISLIQ/(DENLIQ*(DIAM**0.5))
              1)**0.5)
ISN 0134      29 GT=GT1/3600.
ISN 0135      IF(VELVL.LE.10.0) CGAMMA=(0.0127*ALOG10(VISLIQ/2.42+1))/(DIAM**1.4
              115)-0.284+0.167*ALOG10(VELVL)+0.113*ALOG10(DIAM)
ISN 0137      IF(VELVL.GT.10.0) CGAMMA=(0.0274*ALOG10(VISLIQ/2.42+1))/(DIAM**1.3
              171)+0.161+0.569*ALOG10(DIAM)-((0.01*ALOG10(VISLIQ/2.42+1))/(DIAM**
              21.571)+0.397-0.63*ALOG10(DIAM))*ALOG10(VELVL)
ISN 0139      IF(VELVL.LE.10.0.AND.CGAMMA.LT.(-0.065*VELVL)) CGAMMA=-0.065*VELVL
ISN 0141      CGAM1=VR*(GT/DENLIQ-VELVL)/((VR+VELVL)*(VR+VELVL+1.0))
ISN 0142      IF(VELVL.GT.10.0.AND.CGAMMA.LT.CGAM1) CGAMMA=CGAM1
ISN 0144      DENTPS=((GT+DENLIQ*VR)/(VELVL+VR)+CGAMMA*DENLIQ)
ISN 0145      CALL FRFA(FTP,REOS,DIAM,E,M)
ISN 0146      DPORKS=((4*FTP*DENTPS*VELVL*VELVL/(2*32.2*DIAM))+DENTPS)/0.144
ISN 0147      IF(Z.EQ.3.0) GO TO 991
ISN 0149      WRITE(6,50)DENTPS,DPORKS
ISN 0150      50 FORMAT(///,' BY ORKISZEWSKI CORRELATION FOR VERTICAL FLOW',45X,'T
              TWO PHASE DENSITY=',E10.4/,45X,' PRESSURE GRADIENT FOR SLUG FLOW RE
              2 GIME= ',E10.4,' PSI/1000.0 FT')
              GO TO 440
ISN 0151      C*****
              C***** CALCULATION OF MIST FLOW PARAMETERS
ISN 0152      991 ENW=(4.52E-07)*((453.59*VISLIQ/2.42*VGAS/SUFTEN)**2.0)*(DENVAP/DEN
              1 LIQ)
ISN 0153      IF(ENW.GT.0.005) E1=(174.8*SUFTEN*ENW**0.302)/(DENVAP*(VGAS**2.0))
ISN 0155      IF(ENW.LE.0.005) E1=0.075*SUFTEN/(DENVAP*(VGAS**2.0))
ISN 0157      IF(ENW.GT.0.5) E1=0.5*DIAM
ISN 0159      CALL FRFA(FTP,REV,DIAM,E1,M)
ISN 0160      DPORKM=4.0*FTP*DENVAP*(VGAS**2.0)/(2.0*32.2*DIAM)+DENNS
ISN 0161      AC=GT*VGAS/32.17*1.0/(P*144.0)
ISN 0162      DPORKM=DPORKM/((1.-AC)*0.144)
ISN 0163      IF(Z.EQ.3.0) GO TO 58
ISN 0165      WRITE(6,54)DENNS,DPORKM
ISN 0166      54 FORMAT(///,' BY ORKISZEWSKI CORRELATION FOR VERTICAL FLOW',45X,'T
              TWO-PHASE DENSITY= ',E10.4,' LB/CU.FT.'/,45X,' PRESSURE GRADIENT FOR
              2 MIST FLOW REGIME= ',E10.4,' PSI/1000 FT')
              GO TO 440
ISN 0167      C*****
              C***** CALCULATION OF TRANSITION FLOW PARAMETERS
ISN 0168      58 T1=(NLM-NGV)/(NLM-NLS)
ISN 0169      DENTPT=T1*DENTPS+(1-T1)*DENNS
ISN 0170      DPORKT=T1*DPORKS+(1-T1)*DPORKM
ISN 0171      WRITE(6,62)DENTPT,DPORKT

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        62 FORMAT(///, ' BY ORKISZEWSKI CORRELATION FOR VERTICAL FLOW', 45X, 'T
        TWO-PHASE DENSITY =', E10.4, ' LB/CU.FT.', 45X, 'PRESSURE GRADIENT FOR
        2 TRANSITION FLOW REGIME =', E10.4, ' PSI/1000 FT')
        440 IF(ALPHA.EQ.-90.0) GO TO 21
        C*****
        C***** CALCULATION OF FLOW REGIME IN TWO-PHASE HORIZONTAL FLOW
        C***** METHOD VIA HTRI MANUAL
        C***** CALCULATING PARAMETER RLH
        Y=V/(L+V)
        Y1= (DENLIQ/DENVAP)*(Y/(1-Y))
        RLH=1/(1+Y1)
        C*****
        C***** CALCULATING PARAMETER CGT
        Y2=(1-Y)/Y
        GT1=(V+L)/AREA
        CGT=((1/GT1)*((DIAM*4.17E+08*DENVAP*(DENLIQ-DENVAP)*Y2)**0.5)
        WRITE(6,1000)
        1000 FORMAT(1H1,10X, 'FLOW REGIME MAPPING FOR HORIZONTAL TWO-PHASE FLOW'
        1)
        WRITE(6,1001)
        1001 FORMAT(1X, '+++++',
        1+++++',
        2+++++')
        WRITE(6,1001)
        WRITE(6,1002)CGT,RLH
        1002 FORMAT(1X,////, ' BY METHOD OF HTRI', 10X, ' PARAMETERS ARE:', 2X, 'C
        1GT=', E10.4, 27X, 'RLH=', E10.4)
        C*****
        C***** METHOD VIA BAKER
        BY=((2.16*V)/(AREA*((DENVAP*DENLIQ)**0.5))
        BX1=((DENVAP*DENLIQ)**0.5)*(VISLIQ**(1/3.))
        BX2=(DENLIQ**(2/3.))*SUFTEN
        BX=((396.*L)/V)*BX1/BX2
        WRITE(6,1003)BY,BX
        1003 FORMAT(1X,////, ' BY BAKERS METHOD', 10X, 'PARAMETERS ARE:', 2X, 'BY=',
        1E10.4, 27X, 'BX=', E10.4)
        C*****
        C***** METHOD VIA DUKLER
        C*****
        C***** CALCULATION OF DUKLER PARAMETERS
        DY=((DENLIQ-DENVAP)*SIN(3.141593*ALPHA/180.0))/DPDXV
        FA=(DENVAP/(DENLIQ-DENVAP))**0.5
        QGAS=V/DENVAP
        QLIQ=L/DENLIQ
        FB=(QGAS/(AREA*3600))/((DIAM*32.2*COS(ALPHA*3.141593/180.))**0.5)
        F=FA*FB
        K=F*(REL**0.5)
        T=(DPDXL/((DENLIQ-DENVAP)*COS(ALPHA*3.141593/180.0))**0.5
        CALL HLDPAR(HLD,X,DY,REL,REV)
        WRITE(6,1004)REV,REL,X,DY,F,K,T,HLD
        1004 FORMAT(///, ' BY DUKLERS METHOD', 10X, 'PARAMETERS ARE:', 2X, 'REV=',
        1E10.4, 27X, 'REL=', E10.4, 27X, 'X=', E10.4, 27X, 'Y=', E10.4, 27X, 'F=',
        2E10.4, 27X, 'K=', E10.4, 27X, 'T=', E10.4, 27X, 'HL/D=', E10.4)
        VGAS=QGAS/(AREA*3600.0)
        VLIQ=QLIQ/(AREA*3600.0)
        GO TO 790
        C*****
        C***** CALCULATION OF FLOW REGIME FOR VERTICAL TWO-PHASE FLOW

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C***** METHOD VIA GOVIER-AZIZ
21 QGAS=V/DENVAP
   QLIQ=L/DENLIQ
   VGAS=QGAS/(AREA*3600)
   VLIQ=QLIQ/(AREA*3600)
   YL=((DENLIQ*73.)/(62.3*SUFTEEN))*0.25)
   YVL=YL*VLIQ
   XG=(DENVAP/0.075)*0.1/3)
   XVG=XG*YL*VGAS
   WRITE(6,1005)
1005 FORMAT(1H1,40X,' FLOW REGIME MAPPING FOR VERTICAL TWO-PHASE FLOW')
   WRITE(6,1001)
   WRITE(6,1001)
   WRITE(6,1007)YVL,XVG
1007 FORMAT(////,' BY METHOD OF GOVIER-AZIZ',14X,'PARAMETERS ARE:',2X
1,'YVL=',E10.4/,31X,'XVG=',E10.4)
C*****
C***** METHOD VIA JGM,JLM
   KV=VGAS*((DENVAP/(DENLIQ-DENVAP))*0.5)
   JG=KV/((32.2*DIAM)*0.5)
   JL=JGM(L/V)*((DENVAP/DENLIQ)*0.5)
   WRITE(6,1008)JG,JL
1008 FORMAT(///,' BY METHOD OF JGM-JLM',10X,'PARAMETERS ARE:',2X,'JGM=
1',E10.4/,27X,'JLM=',E10.4)
C*****
C***** METHOD VIA ORZEWSKI
   VELVL=VGAS+VLIQ
   NLB=1.071-0.2218*((VGAS+VLIQ)**2)/DIAM
   NLS=50.+70*VLIQ*((DENLIQ/SUFTEEN)*0.25)
   NLM=75.+138.*((NLS-50.)/70.)*0.75)
   NGV=1.938*VGAS*((DENLIQ/SUFTEEN)*0.25)
   IF((VGAS/VELVL).LE.NLB) Z=1.0
   IF((VGAS/VELVL).GT.NLB.AND.(NGV.LE.NLS)) Z=2.0
   IF(NLM.GT.NGV.AND.NGV.GT.NLS) Z=3.0
   IF(NGV.GE.NLM) Z=4.0
   IF(Z.EQ.1.0) GO TO 30
   IF(Z=3.0) 31,32,33
30 WRITE(6,1009)
1009 FORMAT(///,' BY METHOD OF ORZEWSKI, FLOW REGIME PRESENT IS BUBBLE
1 FLOW.')
   GO TO 790
31 WRITE(6,1010)
1010 FORMAT(///,' BY METHOD OF ORZEWSKI, FLOW REGIME PRESENT IS SLUG F
1LOW.')
   GO TO 790
32 WRITE(6,1011)
1011 FORMAT(///,' BY METHOD OF ORZEWSKI, FLOW REGIME PRESENT IS TRANSIT
1ION FLOW.')
   GO TO 790
33 WRITE(6,1012)
1012 FORMAT(///,' BY METHOD OF ORZEWSKI, FLOW REGIME PRESENT IS MIST FL
1OW.')
C*****
C***** HEAT TRANSFER CALCULATIONS
C***** BY COLLIER,LACEY,PULLING CORRELATION
790 IF(N.EQ.1) GO TO 580
   PRL=CP*VISLIQ/AK
   REPRIM=REV*VISGAS/VISLIQ

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ISN 0258      DENRAT=DENLIQ/DENVAP
ISN 0259      HCLP=(0.175*AK/DIAM)*(DENRAT**0.28)*(REPRIM**0.7)*(PRL**0.4)

C*****
C***** BY BENNETT-CHEN CORRELATION
C***** EXCLUDING PRANDTL NUMBER EFFECT
ISN 0260      HLOHLY=0.023*(REL**0.8)*(PRL**0.4)*AK/DIAM
ISN 0261      M=1.0
ISN 0262      CALL FRFA(FLIQ,REL,DIAM,E,M)
ISN 0263      DPDLQ=(4.*FLIQ*DENLIQ*(VLIQ**2.)/(2.*32.2*DIAM)-DENLIQ*(SIN(0.000
1*3.141593/180.)))/0.144
ISN 0264      QF2=DPDLME/DPDLQ
ISN 0265      HTP=HLOHLY*(QF2**0.444)

C*****
C***** INCLUDING PRANDTL NUMBER EFFECT (VIA TWO FORMULAS)
ISN 0266      PRCOR1=((PRL+1.0)/2.0)**0.444
ISN 0267      PRCOR2=(PRL**0.296)
ISN 0268      HTPCO1=HTP*PRCOR1
ISN 0269      HTPCO2=HTP*PRCOR2
ISN 0270      WRITE(6,550)
ISN 0271      550 FORMAT(///,5X,'HEAT TRANSFER COEFFICIENT CALCULATIONS',
1'- BTU/SQ.FT.HR.DEG.')
ISN 0272      WRITE(6,1001)
ISN 0273      WRITE(6,1001)
ISN 0274      WRITE(6,560)HLOHLY,HCLP,HTP,HTPCO1,HTPCO2
ISN 0275      560 FORMAT(//////,1X,'FOR THE LIQUID FLOWING ALONE,H=',1X,E10.4//,1X,'B
1Y COLLIER LACEY PULLING CORRELATION. H=',1X,E10.4//,1X,'USING CHEN
25 CORRELATION WITHOUT PRANDTL EFFECT, H=',1X,E10.4//,1X,'USING CH
3ENS CORRELATION (WITH A CORR. FACT. OF (PRLIQ+1)/2)**0.444)'//,
4H='X,E10.4//,1X,'USING CHENS CORRELATION (WITH A CORRECTION FACTO
5R OF (PRLIQ**0.296)'//, ' H= ',1X,E10.4)
ISN 0276      WRITE(6,570)
ISN 0277      570 FORMAT(///,1X,'NOTE: IN ALL THE ABOVE CALCULATIONS USING CHENS COR
1RELATION ONLY'//, ' THE MACROSCOPIC BOILING MECHANISM WAS CONSIDERED
2')
ISN 0278      GO TO 20
ISN 0279      580 CONTINUE
ISN 0280      STOP
ISN 0281      END

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COMPILER OPTIONS - NAME= MAIN,OPT=02,LINECNT=60,SIZE=0000K,
SOURCE,EBCDIC,NOLIST,NODECK,LOAD,MAP,NOEDIT,NOID,XREF

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ISN 0002      SUBROUTINE FRFA(FR,REY,DIAM,E,M)
              C*****
              C***** THIS SUBROUTINE CALCULATES FANNING FRICTION FACTOR AS A FUNCTION
              C***** OF REYNOLDS NUMBER, PIPE ROUGHNESS AND DIAMETER.
ISN 0003      J=0
              C*****
              C***** RR=RELATIVE ROUGHNESS
ISN 0004      RR=E/DIAM
              C*****
              C***** INITIAL GUESS OF FRICTION FACTOR COMES FROM A FUNCTION INVOLVING
              C***** ONLY REYNOLDS NUMBER.
ISN 0005      FR=0.0014+0.10/(REY**0.28)
              C*****
              C***** FOR LAMINAR FLOW USE LAMINAR FLOW CORRELATION
ISN 0006      IF(M.EQ.1.0) GO TO 2
ISN 0008      IF(REY.LT.2000.0) FR=16.0/REY
ISN 0010      IF(REY.LT.2000.0) GO TO 4
              C*****
              C***** FOR SMOOTH PIPE USE CORRELATION FOR A SMOOTH PIPE
ISN 0012      IF(E.LT.0.00001) GO TO 5
ISN 0014      2 J=J+1
ISN 0015      SR=FR**0.5
ISN 0016      ERR=0.5/SR+0.86*ALOG((RR/3.7)+1.255/(REY*SR))
ISN 0017      IF(J.GE.2) GO TO 1
              C*****
              C***** FOR SECOND GUESS OF FRICTION FACTOR MULTIPLY FIRST GUESS BY 0.9.
ISN 0019      ERRO=ERR
ISN 0020      FFO=FR
ISN 0021      FR=FFO*0.9
ISN 0022      GO TO 2
              C*****
              C***** SMOOTH PIPE CORRELATION
ISN 0023      5 FR=0.0014+0.10/(REY**0.28)
ISN 0024      GO TO 4
              C*****
              C***** IF ITERATION LOOP HAS NOT CONVERGED IN TEN PASSES THEN APPROX.
              C***** CORRELATION IS USED TO DETERMINE FRICTION FACTOR AND WARNING IS
              C***** GIVEN.
ISN 0025      1 IF(J.GT.10) GO TO 3
ISN 0027      FFD=ABS(FR-FFO)/FFO
ISN 0028      IF(FFD.LT.0.02) GO TO 4
              C*****
              C***** CALCULATE NEXT ESTIMATE OF FANNING FRICTION FACTOR
ISN 0030      GR=(ERRO-ERR)/(FFO-FR)
ISN 0031      FFO=FR
ISN 0032      ERRO=ERR
ISN 0033      FR=FR-ERR/GR
ISN 0034      GO TO 2
ISN 0035      3 FR=0.0014+0.10/(REY**0.28)
ISN 0036      WRITE(6,2080)
ISN 0037      2080 FORMAT(' WARNING***FRICTION FACTOR NOT CONVERGED FOR E/D,-APPROX.
              1USED')
ISN 0038      4 RETURN
ISN 0039      END

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COMPILER OPTIONS - NAME= MAIN,OPT=02,LINECT=60,SIZE=0000K,
SOURCE,EBCDIC,NOLIST,NODECK,LOAD,MAP,NOEDIT,NOID,XREF

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      ISN 0002      SUBROUTINE HGLDUP(HHU,VISGAS,VISLIQ,ETA,C1)
                   C*****
                   C***** DETERMINATION OF IN-SITU LIQUID HOLDUP USING HUGHMARK CORRELATION
                   C***** FIRST ESTIMATE
      ISN 0003      HHU=0.9999
      ISN 0004      I=0
      ISN 0005      4 I=I+1
                   C*****
                   C***** ENSURE THAT ESTIMATE OF HOLDUP IS IN RANGE
      ISN 0006      IF(HHU.GT.1.0) HHU=1.0
      ISN 0008      IF(HHU.LT.0.0001) HHU=0.0001
                   C*****
                   C***** DETERMINE HUSHMARK PARAMETER K FROM FUNCTION DEL
      ISN 0010      DEL=C1/((HHU*(VISLIQ-VISGAS)+VISGAS)**0.1667)
      ISN 0011      DEL2=DEL*DEL
      ISN 0012      DEL3=DEL2*DEL
      ISN 0013      IF(DEL.LE.10.0) RK=0.1637-0.31037*DEL+0.03525*DEL2-0.001566*DEL3
      ISN 0015      IF(DEL.GT.10.0) RK=0.1435E-04*DEL2-0.003585*DEL-0.75545
      ISN 0017      W=HHU-1.0-RK*(1.0-ETA)
      ISN 0018      IF(DEL.LE.10.0) DKDEL=0.31037-0.0705*DEL+0.0041*DEL2
      ISN 0020      IF(DEL.GT.10.0) DKDEL=+0.003585-(0.2872E-04)*DEL
      ISN 0022      DDELDR=DEL*(VISLIQ-VISGAS)/(HHU*(VISLIQ-VISGAS)+VISGAS)
      ISN 0023      DDELDR=DDELDR*(-0.1667)
                   C*****
                   C***** DETERMINE NEXT ESTIMATE FOR HOLDUP
      ISN 0024      DWDHHU=1.+(1.-ETA)*DKDEL*DDELDR
      ISN 0025      HO=HHU
      ISN 0026      HHU=HO-(W/DWDHHU)
      ISN 0027      WRITE(6,10)HO,HHU
      ISN 0028      10 FORMAT(1X,E10.4,2X,E10.4)
                   C*****
                   C***** IS HOLDUP ITERATION CONVERGED
      ISN 0029      3 HDIF=ABS(HO-HHU)/HO
      ISN 0030      IF(HDIF.LT.0.01) GO TO 5
      ISN 0032      IF(I.GT.50) WRITE(6,2050)
      ISN 0034      2050 FORMAT(' ++++++++ HUGHMARK HOLD UP NOT CONVERGED ++++++++')
      ISN 0035      IF(I.GT.50) GO TO 5
      ISN 0037      GO TO 4
      ISN 0038      5 RETURN
      ISN 0039      END

```


COMPILER OPTIONS - NAME= MAIN,OPT=02,LINECHT=60,SIZE=0000K,
SOURCE,EBCDIC,NOLIST,NODECK,LOAD,MAP,NOEDIT,NOID,XREF

ISN 0002
ISN 0003

SUBROUTINE HLDPAR(HLD,X,Y,REL,REV)
M=0

C*****
C***** THIS SUBROUTINE CALCULATES DUKLER'S HL/D PARAMETER
C***** FOR USE IN HIS FLOW-REGIME MAP.

ISN 0004
ISN 0005
ISN 0006
ISN 0007

VM=-1.0
FM=-1.0
HLD=0.9
IF(REV.GT.2000.0) VM=-0.2

C*****
C***** INITIAL GUESS FOR HL/D
IF(REL.GT.2000.0) FM=-0.2

ISN 0009
ISN 0011

1 M=M+1

ISN 0012
ISN 0014
ISN 0016

IF(HLD.GT.1.0) HLD=0.9999
IF(HLD.LT.0.1) HLD=0.1
A=(2*HLD)-1.0

ISN 0017
ISN 0018

S=ABS(A)
SI=(1.0-(S**2.0))*0.5

ISN 0019
ISN 0020

SG=ARCCOS(S)
IF(A.LE.0.0) SG=3.141593-SG

ISN 0022
ISN 0023

SL=3.141593-SG
AL=0.25*(3.141593-SG+(A*SI))

ISN 0024
ISN 0025

AG=0.25*(SG-(A*SI))
UL=(AL+AG)/AL

189 ISN 0026

UG=(AL+AG)/AG

ISN 0027
ISN 0028

DL=4*AL/SL
DG=4*AG/(SG+SI)

ISN 0029
ISN 0030

B1=(UL*DL)**FM
B2=(UL**2.0)*SL/AL

ISN 0031
ISN 0032

B=B1*B2
C1=(UG*DG)**VM

ISN 0033
ISN 0034

C2=(UG**2.0)
C3=(SG/AG)+(SI/AL)+(SI/AG)

ISN 0035
ISN 0036

C=C1*C2*C3
F=(X**2.0)*B-C-(4.0*Y)

ISN 0037
ISN 0039

IF(M.GE.2.0) GO TO 3
FO=F

ISN 0040

HLDO=HLD

C*****
C***** SECOND GUESS FOR HL/D
HLD=0.7

ISN 0041
ISN 0042

GO TO 1

ISN 0043
ISN 0044

3 HLDIF=ABS(HLD-HLDO)/HLDO
WRITE(6,7) HLDO,HLD

ISN 0045

7 FORMAT(2(3X,E10.4))

C*****
C***** IS HL/D ITERATION CONVERGED
IF(HLDIF.LT.0.0001) GO TO 5

ISN 0046
ISN 0048

IF(M.GT.50) WRITE(6,1000)

ISN 0050
ISN 0051

1000 FORMAT(' ++++++++ HLD CALCULATION NOT CONVERGED ++++++++')
IF(M.GT.50) GO TO 5

C*****
C***** CALCULATE NEXT ESTIMATE OF HL/D

ISN 0053
ISN 0054

GR=(FO-F)/(HLDO-HLD)
HLDO=HLD

ISN 0055

FO=F

```
ISN 0056      HLD=HLD-F/GR
ISN 0057      IF(HLD.LE.0.4) GO TO 5
ISN 0059      GO TO 1
ISN 0060      5 RETURN
ISN 0061      END
```

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