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H-COAL FLUID DYNAMICS

Study of Ebullated Bed Fluid Dynamics for H-Coal

Final Technical Progress Report, August 1, 1977–December 31, 1979

By

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April 16, 1980

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U. S. DEPARTMENT OF ENERGY

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STUDY OF EBULLATED BED FLUID DYNAMICS FOR H-COAL

FINAL PROGRESS REPORT
AUGUST, 1977-DECEMBER, 1979

I. A. VASALOS, E. M. BILD, D. N. RUNDELL, D. F. TATTERSON

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FOREWORD

The H-Coal process, developed by Hydrocarbon Research, Incorporated (HRI), involves the direct catalytic hydroliquefaction of coal to low-sulfur boiler fuel or synthetic crude oil. The 200-600 ton-per-day H-Coal pilot plant is being constructed next to the Ashland Oil, Incorporated, refinery at Catlettsburg, Kentucky, under ERDA contract to Ashland Synthetic Fuels, Incorporated. The H-Coal ebullated bed reactor contains at least four discrete components: gas, liquid, catalyst, and unconverted coal and ash. Because of the complexity created by these four components, it is desirable to understand the fluid dynamics of the system. The objective of this program was to establish the dependence of the ebullated bed fluid dynamics on process parameters. This will permit improved control of the ebullated bed reactor.

The work performed was divided into three parts: review of prior work, cold flow model construction and operation, and mathematical modeling. The objective of this final report is to outline accomplishments of the project in all three areas.

OBJECTIVE AND SCOPE OF WORK

The overall objective of this work is to improve control of the H-Coal reactor through a better understanding of the hydrodynamics of ebullated beds. The project is divided into three main tasks:

- 1) Review of prior work in three-phase fluidization.
- 2) Construction of a cold flow unit and collection of data.
- 3) Development of a mathematical model to describe the behavior of gas/liquid fluidized beds. The model should be based on information available in the literature and on data generated in the cold flow unit.

Progress made in all three tasks during this project is presented in this report. A list of all publications previously issued under this project is given in Table I.

SUMMARY

This report discusses the results of work aimed at understanding the hydrodynamic behavior of the H-Coal reactor. The scope of the work extended over three areas: review of prior work, data collection, and analysis and model development. Progress in each area is reviewed below.

Review of Prior Work

From the physical description of the H-Coal reactor, it became apparent that the H-Coal reactor can be modeled as a gas/slurry/catalyst system. The review of prior work established that there are several factors which control the bed expansion achieved when gas and/or liquid or slurry flow through a bed of catalyst particles. With liquid flow alone, the bed expansion can be described by several correlations. One of these was selected for correlating the data obtained in this work. This is the Richardson-Zaki correlation describing the volume fraction occupied by the liquid phase as a function of superficial liquid velocity and particle terminal velocity. The properties of the liquid phase and catalyst particle size are implicitly taken into account. When gas is added to the system, it has been generally found that the gas bubble dynamics control the hydrodynamic properties of the three-phase system. With particles greater than 3-4 mm, the gas tends to flow into the form of small bubbles, thus resulting in the so-called ideal bubbly flow behavior. On the other hand, smaller particles tend to promote bubble coalescence, resulting in large bubbles which characterize the churn turbulent flow regime. This flow phenomenon controls not only the bed

expansion for a given flow rate, but also determines the response of the three-phase system to changes in operating variables. For example, for a three-phase system with gas flow in the churn turbulent regime, increasing the gas velocity may lead to bed contraction. Many investigators developed mathematical models to describe their data. Of all these models, two were selected for correlating the data obtained during this work: the Darton-Harrison drift flux and the Bhatia-Epstein generalized wake models. The first model attempts to utilize principles from gas/liquid flow to define whether the system studied operates in the ideal bubbly or churn turbulent regime. For this purpose, the operating conditions--gas and liquid velocity--and the gas and liquid holdups are required. The second model (Bhatia-Epstein) explicitly considers the structure of the bed. The mathematical equations are based on the assumption that the bed consists of three phases: gas bubbles, wakes, and the particulate phase. The particulate phase consists of liquid/solid, and the bubble wakes consist of liquids/solids. The model recognizes that bubble size could vary with the properties of the fluid phases and that the concentration of solids in the wake phase can vary from 0 to that in the particulate phase. The Bhatia-Epstein model was the basis for developing correlations from data obtained under the work discussed in this report.

Data Collection and Analysis

Although the review of prior work established several models which might be applicable for the H-Coal reactor, most of these are based on water systems without addition of fines. It was therefore very important to obtain data in a system which closely simulated the H-Coal reactor. Study of the fluid dynamics in the actual reactor was ruled out by H-Coal participants. The cost of the operation in the process development unit (PDU) operated by Hydrocarbon Research, Incorporated (HRI) is very high for completing a meaningful study. No experimental techniques are reliable which can be used with confidence inside the reactor.

Cold Flow Model.--For the above reason, a cold flow model was constructed which was almost identical in size with the PDU reactor. The reactor ID and length were 6" and 20', respectively. The reactor consisted of four glass sections connected with metal spool pieces. Taps inserted in the spool pieces were used for temperature and pressure measurements. Sample probes could also be inserted through special ports. Since the major goal of this contract was to simulate the PDU reactor, the distributor used in the bottom of the reactor and the cup used for the suction of the recycle pump were identical with those of the PDU reactor. Other design aspects incorporated features previously established by HRI.

The experimental techniques selected for studying the fluid dynamics of the gas/slurry/catalyst system were the gamma-ray technique and differential pressure measurements. In selected cases, direct sampling was used to establish the fines distribution along the reactor. Argon-41 radioactive tracer experiments were also conducted to establish the residence time distribution (RTD) in the reactor.

In all cases, a data acquisition system tied to a ModComp II computer was utilized. The computer operated an elevator where the gamma-ray equipment was mounted. All data inputs were used to prepare a preliminary analysis.

Data Collection.--Meaningful application of this work for the H-Coal process is possible only if liquid, gases, and solids used in the cold flow studies have properties similar to the H-Coal liquids at actual operating conditions. At the beginning of this program, some information was available from HRI which indicated that kerosene/coal char slurries closely resembled the behavior of H-Coal slurries. Since the physical properties of the H-Coal slurries were not known at the start of the program, this work followed HRI's suggestion in selecting coal char/kerosene slurries for cold flow studies. The coal char has a density and particle size distribution which is about the same as the unconverted coal and ash in the H-Coal reactor recycle stream. Kerosene with 1, 5, 10, 15, and 17.8 vol% coal char was studied. Although kerosene was the base liquid, water and mineral oil were also used to scope the effect of liquid properties.

Nitrogen, helium, and Freon-12 were the gases selected. For most of the experimental program, standard H-Coal catalyst was used with dimensions of 1/16" ID, 3/16" particle length.

As part of this program, a sampling system was designed for obtaining samples from the PDU reactor for viscosity measurements. These were carried out by Battelle Labs along with coal char/kerosene slurries with 5.1, 10.5, and 17.8 vol% coal char. An analysis of these results demonstrated that the H-Coal liquids and coal char/kerosene slurries behave as Bingham fluids. The viscosity of the H-Coal liquids at PDU conditions is about the same as the viscosity of the coal char/kerosene slurries. Further work is required in the future to establish the effect of the non-Newtonian behavior on the fluid dynamics of the H-Coal reactor.

Data Analysis.--In analyzing bed expansion with liquid or slurries, it was found that the latter behave in the same way as clear liquids of equal viscosity. Thus, bed expansion with mineral oil (175°F) is the same as with a 17.8 vol% coal char/kerosene slurry of about equal viscosity when operated at the same liquid velocity. Hence, coal char and kerosene can be treated as a single phase. This was supported by direct sampling data which showed that coal char is uniformly distributed throughout the reactor. The Richardson-Zaki correlation was validated with slurry catalyst data. With gas added to a liquid, it was generally found that the viscosity of the fluid medium has a significant effect on gas/liquid/solid holdup. For kerosene, addition of nitrogen results in an increase in bed expansion. Application of the Darton-Harrison model indicated that in this case most of the gas flow is in the ideal bubbly regime. When the viscosity of the kerosene is increased by the addition of coal fines, or when a clear liquid with high viscosity is used, addition of gas may result in bed contraction. The Darton-Harrison drift flux model indicates that in this case most of the gas flows in the form of large bubbles.

Model Development

The experimental data were used to develop parameters for a modified Bhatia-Epstein type model with the capability of predicting not only bed expansion, but also the volume fraction occupied by the various phases. For the successful development of this model, it was necessary to derive correlations which describe a) the bubble terminal velocity as a function of gas and liquid rates; b) the wake volume as a function of liquid velocity; and c) the solids holdup in the wake as a function of bubble terminal velocity. It was generally found that high viscosity due to either clear liquids or addition of fines in kerosene increased the bubble terminal velocity and the size of the wake volume.

Recommendations

Additional measurements of the viscosity of H-Coal liquids from the PDU reactor are recommended to establish whether indeed the H-Coal liquids are Bingham fluids. These measurements should be carried out concurrently with gamma-ray scans of the H-Coal reactor to establish the response of the bed height to changes in liquid and gas flow rates. A comparison of these data with cold flow studies will give information regarding the significance of the results from this work.

INTRODUCTION

The H-Coal process, developed by Hydrocarbon Research, Incorporated (HRI), involves the direct catalytic hydroliquefaction of coal to low-sulfur boiler fuel or synthetic crude oil (1,2). The process has been demonstrated by HRI in a process development unit (PDU) at Trenton, New Jersey. This unit is capable of processing up to three tons of dried coal per day. A schematic of the PDU reactor appears in Figure 1. A 200-600 ton/day pilot plant is under construction at Catlettsburg, Kentucky (2).

In the H-Coal process, coal is dried, pulverized to approximately 100-mesh size, and slurried with coal-derived oil. This slurry is charged with hydrogen to an ebullated bed reactor containing a hydrogenation catalyst (3). The reactor operates between 427-482°C and 136-218 atm.

The unique feature of HRI's H-Coal process is the use of an ebullated bed as the means of achieving good mixing between the catalyst particles, the slurried coal, and the hydrogen gas. The catalyst is in extruded form and is similar to those used in petroleum hydrotreating processes. Typical catalyst dimensions are 1.6 mm diameter and 4.8 mm length.

In the ebullating bed, the upward flow of the slurried coal and gases forces the catalyst bed to expand. The expansion of the bed allows fine particles such as coal and ash to pass through the bed and out with the liquid stream, leaving the catalyst bed within the reactor.

The variation in feed composition and slurry concentration causes significant variation in the feed viscosity. These viscosity changes affect bed expansion and lead to a range of operating conditions, as shown below (4):

% Expansion of Catalyst Bed	63-80
Reactor Slurry Composition:	
% Resid	13-18
% Solids (Coal and Ash)	9-22
Slurry Properties (at Reactor Conditions):	
Viscosity, cp	0.2-0.5
Specific Gravity, g/cc	0.6-0.7
Gas Velocity, cm/sec	1.2-4.9
Slurry Velocity, cm/sec	2.1-5.5

From the description of the H-Coal reactor, it becomes evident that bed behavior is a complex function of process variables and gas, liquid, and solid properties. Several models have been proposed in the literature for describing the behavior of three-phase fluidization as a function of these variables. However, these models have been tested only for air/water fluidized beds. Only limited data exist for liquids other than water, and no data exist for slurries which are of primary importance to the H-Coal process. In order to fill these voids, Amoco Oil has contracted with DOE to study ebullating beds. The ultimate objective is the development of an improved mathematical model which will accurately describe the behavior of the ebullating beds in the H-Coal reactor.

The work was divided into three main tasks: 1) review of literature; 2) construction of a cold flow model to simulate the PDU reactor; and 3) data collection and analysis. This report will present results from all areas. It will be shown that based on experiments conducted with slurries of coal char in kerosene, a model has been developed which describes the holdups of the various phases when the operating conditions and the physical properties of the fluids are known.

Reviewing available data, correlations, and models of three-phase systems is the first important task of the contract. For convenience and reasons which will be understood later, this task has been subdivided into the following areas:

- 1) Liquid/solid fluidization.
- 2) Vertical gas/liquid flow.
- 3) Gas/liquid/solid fluidization.

The review of these areas served as a basis for the development of a mathematical model to correlate three-phase fluidization data.

According to Darton and Harrison (5), a three-phase fluidization model should fulfill the following criteria:

- 1) As the volume fraction of gas approaches zero, the model should describe liquid/solid fluidization behavior.
- 2) As the volume fraction of solids approaches zero, the model should reduce to a suitable gas/liquid model.
- 3) The sum of the volume fraction of the various phases should add up to one.
- 4) The model should predict whether the bed will expand or contract upon the addition of gas.

In addition to the first criteria, understanding the liquid/solid fluidization is also important in predicting the H-Coal reactor bed height when there is an unexpected interruption in the hydrogen flow. Numerous correlations for predicting the bed height in liquid fluidized systems have been reported. These are summarized in Table II. The most widely used method of correlating liquid/solid fluidization data is that of Richardson and Zaki (6). Their analysis recognizes the similarities between sedimentation and fluidization. They observed that the settling velocity of a suspension relative to a fixed horizontal plane was equal to the upward liquid superficial velocity needed to maintain the suspension at the same concentration.

As shown in Table II, the correlation relates the liquid volume fraction (ϵ_l) due to bed expansion to the ratio of the superficial liquid velocity (U_l) to the terminal velocity of a single particle (U_t):

$$\epsilon_l^n = U_l/U_t \quad (1)$$

For spherical particles:

$$n = f(\text{Re}_t, d/D) \quad (2)$$

where: $\text{Re}_t = d_p \rho_l U_t / \mu_l$, Reynolds number
 d = particle diameter
 ρ_l = liquid density
 μ_l = liquid viscosity
 D = bed diameter

The functional form of Equation 2 for different ranges of Reynolds number is given in Table II. It should be noted that for values of Re_t less than 0.2 or greater than 500, the exponent n in Equation 1 is independent of liquid viscosity.

For non-spherical particles, the exponent n in Equation 2 also becomes a function of particle shape. For turbulent Reynolds numbers (greater than 500), Richardson and Zaki found:

$$n = 2.7 K^{0.18} \quad (3)$$

$$\text{where: } K = (\pi/6)d_s^3 d_p^{-3} \quad (4)$$

The constant K is the particle shape factor proposed by Heywood (7). In Equation 4, d_s is the diameter of a sphere with the same volume as the particle, and d_p is the diameter of a circle of the same area as the projected particle when lying in its most stable position.

The addition of gas to a liquid fluidized bed increases the complexity of the system considerably. Kim, Baker, and Bergougnou (11) and Darton and Harrison (5) reported that for the range of superficial gas and liquid flow rates of interest to the H-Coal process, two gas/liquid flow regimes exist in the catalyst bed: 1) the ideal bubbly or bubble disintegration regime, in which the bubbles rise as a uniform, steady cloud with little interaction; and 2) the churn turbulent or bubble coalescing regime, which is a transition between ideal bubbly flow and fully developed slug flow. The churn turbulent regime is dominated by bubble coalescence. Hence, the bubble size is larger than in the ideal bubbly regime, bubble wake effects become important, and the flow is unsteady.

The effect of catalyst particle size on the rate of bubble coalescence has been pointed out by Kim, et al. (11) and by Ostergaard (12,13). Small particles promote the churn turbulent (bubble coalescence) flow regime and reduce gas holdup. Large particles, on the other hand, tend to favor the bubble disintegrating (ideal bubbly) flow regime and increase gas holdup.

Numerous investigators (17,18,19,20,21) have found evidence to support this effect of catalyst particle size. The experimental conditions used by these investigators are listed in Table III. From these studies it follows that the critical particle size between coalescing and non-coalescing beds is between 3 and 4 mm for the systems studied.

In a related phenomenon, Turner (14) was the first to report the contraction of some liquid fluidized beds upon the addition of gas. Kim, et al. (11), proposed that this contraction will occur only in beds in the bubble coalescing flow regime. According to Stewart and Davidson (15), contraction occurs because of the formation of liquid wakes behind the bubbles which move through the bed at the bubble velocity. This action reduces the interstitial liquid velocity in the rest of the bed, thus causing its contraction. Ostergaard has also offered a similar explanation.

A summary of additional experimental studies on three-phase fluidized beds is shown in Table IV. Many of these investigators have correlated their data in various forms, as shown in Table V. It should be emphasized that these correlations should be used only for the specific system and range of variables from which they are developed.

The correlations presented in Table V were developed through empirical observation or dimensional analysis. Semi-theoretical models have also been proposed to correlate three-phase fluidization data. These models take into consideration details of the bed structure, such as bubble rise velocity or bubble wake volume. The latter are often described by empirical relations.

Ostergaard (16) has proposed a model based on the assumption that the bed consists of three phases: gas, wake, and a liquid/solid particulate phase. It is assumed that the wake phase moves through the bed at the bubble velocity. The model consists of seven empirical equations which must be solved by an iterative procedure for the volume fractions of the various phases. Ostergaard (16) reports qualitative agreement between his proposed model and experimental data available in the literature. However, the model assumes that the porosity of the wake phase is equal to that of the liquid/solid particulate phase, and thus cannot predict a bed contraction.

Darton and Harrison (5) have also proposed a semi-empirical model. The model is based on the assumption that the bubble wakes are particle-free and that the liquid flux in the bubble wakes is given by kU_g , where k is the mean value of the liquid wake volume/bubble volume ratio. Darton and Harrison develop an empirical correlation for k as a function of U_1 and U_g :

$$1 + k = 1.4(U_1/U_g)^{0.33} \quad (5)$$

The superficial liquid velocity and the liquid holdup in the particulate phase are given by $(U_1 - kU_g)/(1 - \epsilon_g - k\epsilon_g)$ and $(\epsilon_1 - k\epsilon_g)/(1 - \epsilon_g - k\epsilon_g)$, respectively. These expressions are then introduced into the Richardson-Zaki correlation to obtain:

$$\epsilon_1 = \left(\frac{U_1}{U_t} - \frac{\bar{k}U_g}{U_t} \right)^{1/n} (1 - \epsilon_g - \bar{k}\epsilon_g)^{1-1/n} + \bar{k}\epsilon_g \quad (6)$$

This equation and the requirement that the fractional holdups must sum to unity give two equations with three unknowns.

In order to obtain a third relation, Darton and Harrison use the drift flux approach proposed by Wallis (31) for describing gas/liquid flow regime. Physically the drift flux represents the volumetric flux of gas relative to a surface moving at the average gas plus liquid velocity. The drift flux (V_{CD}) is defined as follows:

$$V_{CD} = U_s \epsilon_g (1 - \epsilon_g) \quad (7)$$

where U_s is the gas liquid slip velocity defined below:

$$U_s = \frac{U_g}{\epsilon_g} - \frac{U_l}{\epsilon_l} \quad (8)$$

Combining Equations 7 and 8, the following relationship is obtained:

$$V_{cd} = U_g(1 - \epsilon_g) - \frac{U_l \epsilon_g}{\epsilon_l} (1 - \epsilon_g) \quad (9)$$

Darton and Harrison analyzed the fluidization data of Michelsen and Ostergaard (23) in terms of Equation 9. A plot of V_{CD} vs. ϵ_g revealed two flow regimes: the ideal bubbly and the churn turbulent. This plot is reproduced in Figure 2. As can be seen from the figure, the data from the two flow regimes fall along different lines. Typical transition data are indicated by the dashed lines between the two regimes. The data for the ideal bubbly regime fall on the line given by:

$$V_{cd} = 180 \epsilon_g (\text{mm/sec}) \quad (10)$$

The condition for bed contraction can be found by adding ϵ_g to both sides of Equation 6 and then taking the derivative of $(\epsilon_l + \epsilon_g)$ with respect to U_g as $\epsilon_g \rightarrow 0$. The condition for contraction is:

$$(1 + \bar{k})(1 - \epsilon_l + \frac{\epsilon_l}{n}) < \epsilon_l \frac{\bar{k}}{n} \frac{U_b}{U_l} \quad (11)$$

Epstein (32) has proposed a similar criterion for bed contraction.

Bhatia and Epstein (33) have also proposed a generalized wake model. Following Ostergaard (16), they also propose three phases to the bed: gas, wake, and the liquid particulate phase. The solids concentration in the wake phase can be varied between zero and that in the particulate liquid/solid fluidized phase by means of an adjustable parameter, X_k . The value of X_k ranges between 0 and 1. The wake volume is related to that of the gas by an empirical function. The particulate liquid/solid phase is described by a Richardson-Zaki type model. The model also takes into consideration the possible existence of two gas/liquid flow regimes in the bed. Bhatia and Epstein develop eight equations for the eight unknowns in the system. These equations are listed in Table VI. The equations must be solved by an iterative procedure.

The authors test this model vs. the data of Michelsen and Ostergaard (23) and data of their own. Their data are for the fluidization of spheres ranging in size from 0.25 to 3.0 mm. The solid density ranged from 2.5 to 11.1 mg/m³. The gas used was air, and the liquids were water (1 cp), aqueous glycerol (2.1 cp), or aqueous polyethylene glycol (63 cp). The experiments were carried out in 20 and 50 mm columns. Best agreement between the model and data was obtained when $X_k = 0$ (solids-free wake).

This model, along with the contraction criteria proposed later by Epstein (32), also meets the four criteria suggested by Darton and Harrison. In many respects Bhatia and Epstein's wake model is very similar to the model proposed by Darton and Harrison.

The literature search has indicated that the gas flow behavior is very critical in determining not only bed expansion, but also conditions under which bed contraction occurs. It therefore follows that a successful model should take into account the bubble size and wake volume following the bubbles. Only semi-theoretical models such as Ostergaard's (16), Darton and Harrison's (5), and Bhatia and Epstein's (33) are considered promising in correlating experimental data.

CONSTRUCTION OF COLD FLOW UNIT AND DATA COLLECTION

From the literature search it becomes evident that although significant contributions have been made in the field of gas/liquid/solid systems, no data exist for slurry systems directly applicable to the H-Coal process. For this reason, equipment was constructed in Amoco Oil which is almost on scale with the 3 T/D coal liquefaction process development unit (PDU) operated by HRI for several years. Data were obtained at ambient conditions with gas/slurry/catalyst systems having similar physical properties to the H-Coal liquids at reactor conditions. Equipment details and properties of the liquid/slurry used will be described in this section.

Equipment

A schematic diagram of the H-Coal fluid dynamics unit is shown in Figure 3. Design and construction of this unit were coordinated by the systems design group of Amoco Oil. Systems design incorporates elements of process design, mechanical design, instrumentation, automation, and computerization. Details on the work done in each area are given in Reference 34.

The process flow of the H-Coal fluid dynamics unit resembles that of the process development unit (PDU) built by HRI. The reactor consists of a 6" ID vessel 20' in length. The slurry is prepared in a 60-gallon tank. Other major vessels include one 100-gallon feed tank, and a 60-gallon gas/liquid separator with a mist eliminator.

Three pumps are needed for the operation of the unit: 1) a transfer mixing pump for facilitating the slurry preparation and the transfer of the slurry to the feed tank; 2) a slurry feed pump for supplying the slurry to the reactor; and 3) a slurry recycle pump for internal slurry circulation in the reactor vessel. Under typical test operating conditions, the slurry feed rate amounts to 3 gallons per minute, while the slurry recycle pump supplies 15 gallons per minute. Gas and liquid overflow from a 1" ID pipe located at the same level as the top of the recycle cup. The gas is separated from the liquid in the separator, D-3. Entrained liquid droplets accumulate in a demister located on top of the separator. The gas passes through Cooler E-2, and it can either be vented to the atmosphere through Valve PV-1 or can be directed to a gas recycle compressor. This flexibility is necessary to use gases such as Freon-12 and helium. The gas combines with the total liquid stream before it returns to the reactor. Some details on selected equipment items are given in Appendix A.

The reactor is constructed from four glass sections 6" in ID and 5' in length. The glass sections are connected through flanges to five metal spool pieces. The spool pieces have entries for sample taps, pressure taps, and thermowells to monitor the system. Samples of the slurry withdrawn here are used to establish the coal fines concentration along the reactor.

An elevator is designed so that a ten-millicurie Cs-137 gamma-ray source and detector can travel vertically along the reactor.

Physical Properties of Liquids and Solids

The objective of the cold flow modeling studies was to establish the hydrodynamic properties of the H-Coal reactor. The usefulness of these studies will greatly depend on the selection of the fluids which at ambient temperatures have physical properties similar to those of the H-Coal liquids at actual operating conditions. For this reason, as part of this contract samples of H-Coal liquids from the HRI PDU were sent to Battelle for viscosity measurements. As will be discussed later, it was found that the viscosity of H-Coal fluids is in the range of 1 to 2 cp at actual operating conditions.

In this work, fluids and slurries were used which at about room temperature have properties similar to those of H-Coal slurries in the H-Coal reactor. Water, kerosene, and mineral oil were selected with properties shown in Table VII. The properties were measured as a function of temperature. A Fischer Scientific surface tensiometer (Model 20) was used. The results indicate that surface tension varies linearly with temperature over the range studied. Since during this study He, N₂, and Freon-12 gases were used, it was essential to establish the effect of these gases on interfacial tension. Results from these measurements are reported in Table VIII. It can be seen that there is a small variation of surface tension with gas type, most likely due to gas dissolving in the liquid phase.

In addition to the viscosities of the pure liquids reported in Table VII, the viscosities of kerosene/coal char slurries were measured. Initially, a capillary tube viscometer having an ID of 1.6 mm and a length of 100 cm was constructed and connected to the unit as shown in Figure 4. The viscosities of 11.9 and 15.5 vol% coal char/kerosene slurries were found to be 3.5 cp at 72°F and 3.7 cp at 77°F, respectively. Operation of this viscometer was hampered by frequent settling of the fines and subsequent plugging. Samples of kerosene slurried with 5.1, 10.4, and 17.8 vol% coal char were also sent to outside laboratories for viscosity measurements. These measurements are reported in Table IX. From the comparison of these data with the viscosity of H-Coal liquids at actual operating conditions (see Appendix D), it is concluded that the viscosity of kerosene slurries with 10 to 15 vol% coal char have about the same viscosity as the H-Coal reactor liquids at actual operating conditions.

Another important factor for the successful completion of the cold flow modeling studies is the selection of the fines used for the preparation of the slurries. Previous studies by HRI (35) have established a material (coal char) which has similar properties to those found in the recycle stream of the H-Coal reactor. Results from these measurements are reported in Table X. For a more detailed particle size analysis of the coal char, measurements were conducted by the Illinois Institute of Technology using an optical microscope interfaced with a Quantimet 720 computerized image analyzer. The distribution is reported in Table XI. The geometric mean particle diameter is 3.4 microns.

The catalyst particle size distribution has also been measured. The nominal length of catalyst chosen for this study is 4.8 mm; the diameter of the catalyst is 1.6 mm. This catalyst is similar to the one previously used by HRI in PDU studies. The catalyst has been supplied by American Cyanamid. Cyanamid has guaranteed that the variation of length does not exceed $\pm 20\%$ for each of the above lengths. The length distribution and other catalyst properties are listed in Table XII.

Unit Data

A summary of all experimental runs conducted during this program is reported in Table XIII. For each run, a given amount of catalyst was loaded to the reactor filled with liquid. The catalyst particles were allowed to soak in the test fluid for a period of over 24 hours. The experimental techniques then used to find the volume fraction occupied by the various phases for each run were pressure drop measurements, gamma-ray scans, and radioactive gas tracers.

Pressure Drop Data.--These can be used to supplement the other techniques for calculating the volume holdup of the various phases. The method of calculation is outlined in Appendix B. Because pressure taps through the spool pieces were used, only average holdups through the bed can be calculated.

Gamma-Ray Scan Data.--A zero gamma-ray scan was then carried out to find the zero bed height (H_0) at zero gas and liquid flow. For each subsequent set of operating conditions, the gamma-ray elevator was used to scan the reactor length. These data, obtained by the ModComp II computer, were stored for subsequent analysis and plotting. An example of gamma-ray scans of various gas velocities is shown in Figure 5. From plots of this type, a bed height was established for each test. Based on this bed height, the initial catalyst charge, the pressure drop, and the gamma-ray measurements along the reactor, the holdup of each individual phase was determined. When fines are present, additional information is needed. This was established by measuring the fines distribution along the reactor by direct sampling through the spool pieces. Details of the calculation procedures to establish the holdup of each phase are reported in Appendix B.

Tracer Data.--The use of a radioactive gas tracer (argon-41) has previously been considered by Michelsen (23). The gas tracer used for this work was also argon-41 (A^{41}), a gamma-ray emitter which is prepared by neutron bombardment of highly purified argon-40. Enough radioactive A^{41} was prepared so that the activity was at least 150 mc when injections into the tracer were started.

The tracer was injected through a tube inserted into the reactor above the bubble cap distributor. A previous study (23) using A^{41} as the tracer noted difficulties in analyzing tracer data due to the interphase transfer of argon. Therefore, non-radioactive A^{40} was used to presaturate the system before injection of the tracer. Experiments also showed that absorption of argon by coal fines is negligible.

Six scintillation detectors were mounted on the system as shown in Figure 6. The spacing between detectors is shown in the same figure. Detectors 1 and 2 always viewed the catalyst bed. Detectors 3 and 4 always viewed the liquid or slurry phase above the bed. Detectors 5 and 6 viewed gas lines only and were used primarily to estimate the overall gas holdup in the system.

All the detectors were highly collimated with lead shielding to view only a thin vertical segment, but to cover the entire reactor cross-section. The detectors were connected to an electronics train which included high-speed data recorders and a data acquisition system linked to a ModComp II mini computer. The data were collected at two rates--either 5 or 10 measurements per second.

Tracer tests and gamma-ray scans of the reactor were obtained at the same system conditions. Reactor holdups determined by the scans were compared with tracer results. The tracer data analysis procedure is reported in Appendix C.

Catalyst and Coal Fines Settling Data.--These were determined by shutting off all gas and liquid flows and using the gamma-ray equipment to determine gamma-ray attenuation at a specific location of the reactor as a function of time. Successive runs of this type were used to measure the settling rate of the fines and the catalyst particles from the slope of the graph of interface height vs. time.

A summary of the settling data for HDS-2A catalyst ($l/d = 3$) fluidized by nitrogen appears in Table XXIV. A plot showing the settling of the fines appears in Figure 7.

Physical Properties of H-Coal Liquids

Measurement of the viscosity of H-Coal liquids is very important to determine if the viscosity of H-Coal liquids at reactor conditions is in the same range as the viscosity of the liquids and slurries studied in the cold flow unit. As a part of the contract, this comparison was made. With the assistance of HRI, H-Coal liquid samples were obtained from the PDU at HRI. The viscosities of these samples, along with three char/kerosene samples, were determined by Battelle Labs (Columbus, Ohio) under a subcontract. Battelle was selected to make these measurements because they had previous experience in measuring the viscosity of Synthoil liquids.

PDU Sampling Technique.--The H-Coal samples were taken during PDU Run 8 in July, 1979. The sampling point on the PDU was located between the external separator and the two-stage pressure letdown valves. The sampling technique is described in Appendix D.

Four H-Coal samples were obtained. The reactor temperature and pressure during the sampling were approximately 850°F and 2600 psig, respectively. Additional data for the samples are contained in Table XIV.

Viscosity Measurements.--Battelle measured the viscosity of the H-Coal samples and the char/kerosene samples (5.1, 10.4, and 17.8 vol%) in a viscometer described in Appendix D. It was found that both the H-Coal samples and the char/kerosene samples could be characterized as Bingham fluids. The relationship between the shear stress and the shear rate for a Bingham fluid is illustrated in Figure 8. As can be seen from the figure, the relationship is linear, with slope of the line being equal to the viscosity. However, unlike a Newtonian fluid, a certain stress must be overcome before deformation of the fluid occurs. This stress is known as the yield stress and is illustrated by O in Figure 8. Thus, the relationship between shear stress and shear rate for a Bingham fluid can be summarized as:

$$\gamma = 0 \quad \sigma < \theta \quad (12)$$

$$\sigma = \theta + \eta' \dot{\gamma} \quad \sigma > \theta \quad (13)$$

where: η' = viscosity
 $\dot{\gamma}$ = shear rate

The measured values of η' and θ for the H-Coal samples and the char/kerosene samples are listed in Appendix D. The viscosity of the H-Coal samples is plotted as a function of temperature in Figure 9. It is shown that the viscosity is a strong function of temperature. It was found that the pressure had very little effect on viscosity. A comparison of the viscosities of H-Coal liquids with the viscosity and yield stress of coal char kerosene samples at ambient conditions shows that both are in the same range. The viscosity of the H-Coal samples decreased from 48 to 1.2 cp as the temperature increased from 340 to 831°F. The viscosity of the char/kerosene samples displayed the same behavior, decreasing by a factor of approximately 2 as the temperature increased from 70 to 140°F.

The overall effect of the yield stress on the fluid dynamics of the ebullated bed would be to reduce mixing in the bed. This is because before any relative motion could occur between the liquid and gas or solid phases, the yield stress would have to be overcome. For instance, for a bubble rising in the bed, the velocity would be slower because the buoyancy forces would have to overcome the yield force in addition to drag forces. The magnitude of the effect would of course depend on the relative magnitude of the buoyancy and yield forces. For instance, for a yield stress of 10 dynes/cm² (1000 milli N/m²), typical value found by Battelle, the yield force on a bubble 1 cm in diameter would be 7.85 dynes. If the bubble were rising in a fluid of density 1 gm/cc (neglecting gas density), the buoyancy force would be 512 dynes. Thus, the effect of the yield stress would be expected to be small. However, as the bubble size decreases, the effect will become more important.

The effect of non-Newtonian liquids on the behavior of bubbles and solids in them is a subject which has received relatively little attention in the literature. However, it is a subject which could be of great potential interest in further understanding the hydrodynamics of ebullating beds, and it should be studied further.

The principal conclusions which can be drawn from the viscosity measurements are:

- 1) H-Coal liquids are characterized as Bingham fluids.
- 2) The viscosity of H-Coal liquids is a strong function of temperature.

- 3) The viscosity of H-Coal liquids is a weak function of pressure.
- 4) Char/kerosene slurries used in the cold flow unit are also characterized as Bingham fluids.
- 5) The viscosity of char/kerosene liquids used in the cold flow unit is in the same range as the viscosity of H-Coal liquids at reactor conditions.

Data Analysis

In this section, cold flow model data are presented which establish the effect of coal char concentration, temperature, gas and slurry velocity, viscosity, and gas type on bed behavior. The effect of the various parameters on bed expansion, catalyst holdup, and gas holdup will be presented. Following these data, correlations will be considered developed on the basis of existing models described in the literature section. However, it is important first to establish whether or not the coal char fines are uniformly distributed throughout the reactor.

Fines Distribution.--The distribution of coal char in the fluid dynamics reactor has been measured by taking samples from the unit. The weight per cent of coal char in the slurry sample is found by millipore filtration. The results of this analysis for 5.1, 10.4, 11.9, 15.5, and 17.8 vol% coal char/kerosene slurries are given in Table XV. In general, little variation in char concentration with position in the reactor is observed. Particle size analysis was also done on reactor bottom and 451-cm-level samples from Test 211-07. These results are given in Table XVI. This analysis was performed to determine if there were particle segregation in the reactor due to flotation effects. The results for the samples taken at the top and bottom of the reactor are similar; however, the sample from the top of the reactor has a smaller average particle size, 2.8μ vs. 3.5μ .

Comparison of the coal char concentration at the reactor bottom and the 457 cm level for Tests 206-5 and 211-11 in Table XV also is a check of coal fine segregation due to flotation. If small fines are preferentially rising to the reactor top with gas bubbles, the difference between concentrations should be greatest for the high gas flow rate case. However, Table XV indicates no effect of gas flow rate. It therefore follows that assuming that the fines are well mixed throughout the reactor is a good approximation. Other investigators have established the validity of the same assumption (37).

Liquid or Slurry Fluidization.--The effect of coal fines concentration on bed expansion at zero gas flow is shown in Figure 10. Increasing the coal char concentration in kerosene from 0 to 17.8 vol% (viscosity increase

from 1.4 cp to 8 cp at 75°F) increased bed expansion by a factor of 2 in most cases over the range of slurry velocities studied. The effect of slurried temperature on bed expansion is also presented in Figure 10. Changing the temperature from 148°F to 85°F (viscosity change from 6.8 to 8.0 cp) also increases the bed expansion. However, the increase in bed expansion is lower because there is a smaller viscosity change.

The need to establish whether the increased bed expansion from coal char/kerosene slurries is due to their high viscosity necessitated experiments with mineral oil. The viscosity changed with temperature from 4.2 cp at 175°F to 22.4 cp at 70°F. Bed expansions at various liquid velocities with the temperature as a parameter are plotted in Figure 11. It is again observed that increasing the viscosity has a large effect on bed expansion. In fact, Figure 12 indicates that coal fines/kerosene slurries and mineral oil result in the same bed expansion.

The effects of the catalyst properties on bed expansion are shown in Figure 13. For 1/16" diameter catalyst particles, decreasing the length from 3/16" to 2/16" will increase the bed expansion by about 20%. No studies with fines were performed using the $l/d = 2$ catalyst; however, the conclusions drawn from Figure 13 are expected to be applied in these cases. A much smaller increase in bed expansion was obtained when for particles with $l/d = 3$, the bulk density increased from 0.7 to 0.9. The high-density catalyst was used in a PDU run, and as a result it had a low pore volume (~ 0.25 cc/gm) compared with 0.66 of the catalyst used for most of the tests in this study. Since the bed expansion depends on the density of the catalyst particles soaked with liquid, the small bed increase is due to the fact that both catalysts have about the same soaked catalyst density.

For a liquid/solid system, the liquid holdup can be calculated from the bed expansion with the procedures described in Appendix B. Correlation of the liquid holdup with operating conditions and the physical properties of the system can be done by the application of one of the several models listed in Table II. The Richardson-Zaki (6) model was used during this work because of its successful application in previous studies with systems similar to those considered in this contract (45). Following Equation 1, plotting λ vs. U_1 in log-log paper or a plot of $\ln \lambda$ vs. $\ln U_1$ in linear paper should result in a straight line with slope equal to the reciprocal of the Richardson-Zaki index n . This is illustrated in Figure 14, where the liquid volume fraction is plotted as a function of superficial slurry velocity. Utilizing standard linear regression techniques, the experimental Richardson-Zaki index n was determined for each run. These are reported in Table XVII. The results indicate that the Richardson-Zaki index changed very little with either coal char fines concentration or viscosity. This is expected because the index n is insensitive to changes in this Reynolds number range. For example, according to the correlation for n given in Table II ($1 \leq Re \leq 200$), to obtain a 10% change in n , the viscosity must change by a factor of 3.

Three-Phase Fluidization--Gamma-Ray.--Bed expansion for both kerosene and kerosene slurries is presented in Figure 15. It has been previously discussed that the addition of gas in a liquid fluidized bed will generally result in an increased bed expansion. Figure 17 indicates that although this is generally true for the nitrogen/kerosene/catalyst system with zero fines, bed expansion is in fact reduced in some cases with a 17.8 vol% coal fines/kerosene slurry. It is believed that the fines uniformly distributed throughout the reactor increase the viscosity of the kerosene, promoting the formation of large bubbles due to coalescence. Each large bubble carries with it a substantial wake of liquid/catalyst. As a result, the liquid in the wake goes through the bed much faster than if it flowed through the interstices between the particles when the gas was absent. Thus, the wake liquid may be regarded as not flowing through the bed at all and, as far as the particles are concerned, the liquid flow is reduced and therefore the bed contracts. The effect of temperature on bed expansion with a 15.5 vol% coal char slurry system is illustrated in Figure 16. It is shown that as the temperature is reduced, the bed expansion increases due to the viscosity increase of the coal char/kerosene slurry. A comparison of Figures 15 and 16 establishes again that the effect of coal fines is mainly due to the increased viscosity of the liquid kerosene. Figures 15 and 16 illustrate the conditions under which bed contraction occurs. As Ostergaard (30) has reported previously, the bed contraction is a function of the relative liquid/gas flows and the viscosity of the liquid phase.

In an effort to compare a coal char/kerosene slurry with mineral oil of about the same viscosity, bed expansions are plotted in Figure 17 for both systems at different operating conditions. Kerosene liquid data are also presented for comparison. The comparisons shown in Figure 17 indicate that coal char/kerosene slurries behave in a manner similar to a pure liquid of the same viscosity. A similar conclusion was reached in an earlier section with a two-phase slurry/catalyst system.

Although the viscosity has great influence on gas flow behavior, it has also been reported previously (20,21,18,19) that small particles promote bubble coalescence and hence favor conditions under which bed contraction takes place. Large particles, on the other hand, tend to reduce bubble coalescence. It has been found that the critical particle size is between 3 and 4 mm for conditions reported in the literature. In an effort to establish the effect of particle size on gas flow behavior, the fluidization of catalyst particles with nitrogen and kerosene was studied when the l/d ratio of the particles changed from 3 to 2. Figure 18 indicates that over the range of particle dimensions studied, the gas flow behavior remains the same. As expected, the small particle size results in greater bed expansion due to its lower terminal velocity.

Another factor which has also been considered in this study is the density of gas and the influence of gas type on the surface tension of the liquid. As reported in Table VIII, the surface tension of kerosene slightly

increases when it is in contact with He instead of N₂. As shown in Figure 19, this small difference does not appear to have a significant effect on either bed expansion or the conditions under which bed contraction occurs. A similar conclusion is shown in Figure 20, where catalyst particles are fluidized with mineral oil and helium or nitrogen.

Bed expansions can be expressed in equivalent catalyst holdup (ϵ_c) with the equations listed in Appendix B. Plots of ϵ_c as a function of operating conditions and system physical properties are also reported in Appendix B. These plots will be very useful if it is desired to explore the application of various models in predicting the catalyst holdup ϵ_c .

In reviewing the gas/liquid (or slurry)/catalyst data, it has been shown that increasing the viscosity of the liquid phase results in increased bed expansion as the liquid rate increases, and in possible bed contraction as gas is added to the system. In explaining the bed contraction, the suggestions of other investigators were accepted stating that increased viscosities result in bubble coalescence and significant wake effects. Bubble coalescence should in addition result in a decrease in gas holdup because the latter is related to bubble velocity, U_b , and the average gas superficial velocity, U_g , by the following equation (16):

$$\epsilon_g = \frac{U_g}{U_b} \quad (14)$$

The gas holdup in the catalyst bed is plotted versus the gas velocity in Figure 21. It is shown that gas holdup increases with gas superficial velocity with kerosene at zero coal char fines concentration. However, when coal char fines are added to the system, the gas holdup in some cases is in fact reduced with increasing gas velocity. It appears that this reduction in gas holdup is a function of the relative liquid/gas flows and the viscosity of the continuous fluid phase. The low-viscosity kerosene has the highest holdup. Gas bubbles remain small and do not coalesce. The average bubble rise velocity (U_b) is correspondingly slow. In contrast to this, when coal fines are added to the system, gas coalesces to larger bubbles having a higher rise velocity. As a result of this, the gas holdup is reduced. Although various reasons can be invoked explaining the effect of coal fines on bubble coalescence, Figures 21 and 22 indicate that viscosity effects are important: the increased viscosity of the mineral oil decreases the holdup in a similar manner to the coal char kerosene slurries. The need to understand the changes in gas holdup discussed above necessitated the gas tracer tests discussed in the next section.

Three-Phase Fluidization--Tracer Data.--Residence time distribution curves obtained from radioactive gas tracer experiments can in principle be used to calculate the gas holdup (ϵ_g) inside the reactor using the linear velocity (V_g) and the superficial velocity (U_g) as follows:

$$\epsilon_g = \frac{U_g}{V_g} \quad (15)$$

Typical residence time distribution data with (15.5 vol%) and without coal fines are shown in Figures 24 and 23, respectively.

The first and second moments for curves such as those shown in Figures 23 and 24 were calculated using the equations included in Appendix C. A comparison of the first and second moments calculated by procedures described in Appendix C for selected cases is shown in Table XVIII.

The gas linear velocity in the catalyst bed and above the catalyst (dilute phase) was calculated from the tracer results for several cases. The calculated velocities and test conditions are shown in Table XIX. Replicates for each case show generally good agreement with typically less than 10% error. However, experiments performed near the onset of churn turbulent behavior show wider variation.

Comparison of gas velocities with and without coal char at similar operating conditions shows that the average gas velocity in the catalyst bed is not significantly changed. However, addition of the fines causes a more rapid tracer response at the top of the bed, combined with increased dispersion and apparent backmixing. These results support earlier statements that the addition of coal fines enhances bubble coalescence, causing the formation of larger bubbles which rise quickly through the reactor.

As discussed earlier, this bubble coalescence is probably caused by the higher slurry viscosity due to coal fines. As reported by Calderbank, et al. (42), bubble coalescence is enhanced by more viscous fluids. The same author noted that larger bubbles tend to form and rise in the center of the column. This results in a downward motion of gas at the walls.

Rigby, et al. (46), also reported the tendency of gas bubbles to preferentially rise in the center of a gas/liquid fluidized bed. This results in a downward motion of gas at the walls which may cause the spreading of the tracer concentration, as indicated by the comparison of the second moments for Tests 1 and 2. This comparison indicates that coal fines cause the tracer to spread out in the reactor, possibly as a result of the downflow of gas near the walls.

A comparison of gas holdup calculated from gamma-ray scans with tracer results is shown in Table XX. In all cases, the values calculated from tracer results are significantly higher than those determined from gamma-ray scans. As discussed earlier, results from the tracer test reflect the gas holdup in the entire reactor cross-section, whereas the gas holdup calculated from gamma-ray scans reflects a value corresponding to a measurement taken across the diameter of the reactor.

To determine if a radial gas distribution could account for the difference between holdups calculated by the two methods, gamma-ray scans through three chords of the reactor were obtained. Gas holdups were determined through the reactor center and on two chords at different flow conditions. The test results are given in Table XXI.

Except for one value, the results are all within experimental error. This implies there is not a significant radial distribution of gas which could account for the large differences between holdups calculated by the two techniques.

It is believed that the discrepancy in the calculated holdups can be explained by the gas flow pattern described earlier. Flow of large gas bubbles in the center of the reactor followed by downward flow of gas near the walls will result in long residence time distributions. This in turn will give apparently low linear gas velocities and high gas holdups. On the other hand, the gas holdup calculated from other techniques is independent of the direction in the gas flow.

The qualitative picture of large bubbles traveling up in the center of the reactor causing downflow at the walls is consistent with the differences in holdups calculated by the two methods. Visual observation of the reactor also supports this model. A significant downflow of small gas bubbles can be seen at the reactor walls.

Results from the gamma-ray scans through different chords of a horizontal section are also consistent with this model for gas flow in the reactor. The gas holdup in the two different flow regimes could be the same, although the gas is traveling in different directions.

The ratio between the gas holdups calculated from gas tracer and gamma-ray tests can be related to the amount of gas in each flow regime. However, it must be assumed that the detector views the reactor impartially; tracer in the center of the reactor contributes equally to the total signal to the tracer at the walls. As the ratio of tracer to gamma-ray scan calculated gas holdups increases, the fraction of gas traveling down in the reactor or the cross-sectional area occupied by gas moving down increases. This ratio increases with the addition of coal fines to the reactor, supporting previous indications that coal fines enhance bubble coalescence to form large bubbles which would rise through the center of the reactor.

To quantify the extent of gas mixing inside the reactor, equations described in Appendix C were used to calculate the gas dispersion coefficients inside the bed (Detectors 1 to 2) and the region above the catalyst bed (Detectors 3 to 4). These results are reported in Table XXII. From the magnitude of the dispersion coefficients, it is confirmed that there is apparent gas backmixing throughout the reactor. Work describing a mixing model quantifying the observed results is discussed in a later section.

Three-Phase Fluidization--Data Analysis.--Three-phase fluidization data were analyzed using two different correlations: Darton and Harrison's drift flux approach, and Bhatia and Epstein's generalized wake model. Results of the drift flux analysis will be reviewed first. The drift flux V_{CD} is defined in Equation 9. Calculated values of drift fluxes for all runs are reported in Appendix B. As indicated in Figure 2, plotting V_{CD} vs. gas holdup ϵ_g could identify two flow regimes: churn turbulent and ideal bubbly.

This type of plot for the nitrogen/kerosene/catalyst system with 0 vol% and 17.8 vol% fines is shown in Figure 25. For kerosene with 0 vol% fines, the data fall in the ideal bubbly flow regime. On the other hand, addition of coal fines enhances the transition from ideal bubbly to churn turbulent flow by increasing the slurry viscosity. Also shown in Figure 25 is the effect of increased liquid flow rate on inhibiting the ideal bubbly to churn turbulent transition. Figure 26 confirms the fact that viscosity increase is the main reason for bubble coalescence. Mineral oil of about the same viscosity as coal char/kerosene slurry enhances the transition to churn turbulent. Figures 25 and 26 indicate that a large amount of the data at high viscosities lies in the transition region, suggesting that bubble size continually changes as the gas rate increases. For this reason, it is very difficult to use the drift flux model as a quantitative tool in correlating holdups for the experiments conducted in this study. Instead, an effort was made to test the Bhatia and Epstein model. The equations describing this model were previously presented in Table VI.

The iterative method used to solve the Bhatia-Epstein equations given in Table VI is shown in Table XXIII. Prior to correlating the data obtained in this study, the sensitivity of this model was tested for changes in X_k (ratio of solids holdup in the wake to the solids holdup in the particulate phase), the Richardson-Zaki index n , terminal catalyst particle velocity (U_t), and gas bubble velocity (U_{tB}). It was found that the model was most sensitive to terminal bubble velocity. Other variables have secondary effect. For example, a 10% change in either the Richardson-Zaki index or the catalyst terminal velocity (the error in determining these values) does not have a significant effect. Therefore, use of the Bhatia and Epstein correlation requires a good estimate of the bubble terminal velocity (i.e., bubble diameter).

A preliminary analysis of the data in this study with the Bhatia-Epstein model was performed. the variables which must be known to solve the set of equations are: U_l , U_g , X_k , U_t , U_{tB} , and n . The gas and liquid velocities are known. The Richardson-Zaki index and catalyst terminal velocity were determined experimentally. The relative wake solids content, X_k , was calculated from an empirical correlation developed by El-Temtamy and Epstein (38):

$$X_k = 1 - 0.877 \frac{U_t}{U_s} \quad (16)$$

The value of the terminal bubble velocity was then varied to give the best fit to experimental data.

Comparisons of experimental catalyst holdups with results from the Bhatia-Epstein model for tests with 0 vol% fines and 17.8 vol% showed that agreement between calculated and experimental holdups is generally good except when bed contraction occurred for the 17.8 vol% fines tests. In this case, the Bhatia and Epstein correlation did not predict bed contraction.

For the 0 vol% fines case, the calculated value for X_k was always about zero. Best agreement with the model was obtained with low terminal bubble velocities, corresponding to relatively small bubbles. With 17.8 vol% coal fines, X_k ranged from 0.1 to 0.8. For this case, best agreement with the model resulted in terminal bubble velocities significantly higher than the 0 vol% fines case. This indicates that with coal fines the gas coalesces to large bubbles which rapidly move through the reactor. Radioactive gas (argon-41) residence time distribution tests discussed earlier confirmed that these phenomena are actually occurring.

From the preceding discussion it appears that a Bhatia-Epstein type model offers the best promise in determining the phase holdups once operating conditions are specified. However, in fitting this model, it was found that in certain cases it failed to match some experiments. For this reason, a modified model describing the experimental results of this study was developed. This will be discussed in a later section.

Catalyst and Coal Fines Settling--Data Analysis

Examination of the data summarized in Table XXIV suggests that the bed settling rate after loss of flow is appreciably slower than the corresponding particle terminal velocity. The ratio of terminal velocity to settling rate appears to decrease as bed expansion increases. It is therefore likely that one factor controlling bed settling rate is the dynamics of displacing the excess liquid and gas held up in the expanded bed.

The settling rate of the coal fines in the reactor (as shown in Figure 7) was significantly slower. Particles of this low average diameter and low excess density above the fluid phase are probably forming a colloidal suspension. Although these results support the earlier claims that the fines are uniformly dispersed in the liquid phase, further settling tests should be performed using the higher-density reactor fines.

Comparison of Experimental Results with HRT PDU Data

The most important criterion for the correlations developed from these experimental data is to be able to predict the effect of operating conditions in an H-Coal reactor. Therefore, to test the validity of these correlations, data from cold flow studies should be compared with similar information from the H-Coal reactor at actual operating conditions.

A comparison of bed expansion in the HRI PDU with experimental bed expansion is shown in Figure 27. Results from PDU Run 7 (40), indicated by triangles, are compared with cold flow unit tests with 17.8 vol% coal char in kerosene at 75°F. Reactor slurry velocity was 0.09 ft/sec in all cases. PDU bed expansions are predicted well by these cold flow tests. The differences between bed expansions are within experimental error. The results indicate that a correlation developed from experimental data should closely model H-Coal PDU reactor bed expansions.

MODEL DEVELOPMENT

The objective of the model development is the description of the hydrodynamics of the H-Coal reactor. It is desired to obtain bed height and volume fraction occupied by gas, liquid slurry, and catalyst as a function of gas and liquid superficial velocity. Although it is not known with certainty that a model developed from cold flow studies will be applicable for the H-Coal reactor at actual operating conditions, the agreement of catalyst bed expansions at equivalent operating conditions suggests that the coal char/kerosene slurries are suitable model fluids. Additional recommendations for model validation will be made with the conclusions.

Since from the data analysis reported in the previous section it becomes evident that gas flow dynamics control the holdup of the various phases and the gas mixing inside the reactor, Bhatia and Epstein's approach will be used to take into account the changing bubble diameter with viscosity and operating conditions. Subsequent to the development of the model for predicting volume holdups, a gas mixing model will be presented.

Three-Phase Holdup Model

In developing this model it will be assumed that fines are distributed uniformly throughout the liquid phase. The ebullated bed is then assumed to be composed of three phases: gas, wake, and particulate. Following previous investigators (5,6,16), the assumption will be made that the particulate phase is composed of liquid and catalyst only, and that bubble wakes contain liquid and solids. Crucial parameters of the model are: bubble rise velocity (U_{tB}), relative solids holdup in wake (X_k), and the wake volume behind the bubbles.

Following Henriksen and Ostergaard (48), it is assumed that bubbles in a three-phase fluidized bed behave qualitatively in the same manner as bubbles in pure liquids. The included angle θ of spherical cap bubbles depends on the viscosity of the fluidized bed as shown in Figure 28 (same as Figure 3 in Reference 48). As the bed expands, the apparent viscosity decreases, and this will result in a decrease in the bubble-included angle. Because the wake volume is that which completes the sphere defined by the bubble cap (49), the volume of the bubble wake increases with bed expansion.

The wake volume ratio $K (= \epsilon_k / \epsilon_g)$ was incorporated into the model following El-Temtamy and Epstein's (38) correlation:

$$K = K_0 \exp(-5.08 \epsilon_g) \quad (17)$$

where $K_0 = \epsilon_k / \epsilon_g$ for a single bubble in a fluidized bed.

In deriving the parameters U_{tB} , K_0 , X_k for the model, the experimental phase holdups were matched with model predictions by standard non-linear optimization techniques. The following procedure was adopted:

- 1) Match experimental runs with U_{tB} (the bubble rise velocity) and K_0 as the independent variables with an optimization technique.
- 2) Correlate K_0 and recalculate U_{tB} values.
- 3) Correlate Π_{LB} .
- 4) Using the correlated U_{tB} , K_0 , Equations 16 and 17, and the mathematics of Table XXIII, predict phase holdups.
- 5) Compare predictions with experimental data. If substantial deviations exist, calculate X_k and make predictions using correlated U_{tB} , K_0 , and X_k .

Correlation of Wake Volume Ratio, K_0 .--As discussed previously, the wake volume increases as the bed viscosity is reduced. It therefore follows that high bed expansions will result in an increased wake volume. Because with coal char/kerosene slurries bed expansion is mainly a function of liquid velocity, the calculated ratio K_0 was correlated with liquid velocity. Figure 29 indicates that with kerosene and zero fines the wake volume shows little change with liquid velocity. In this case the gas flows in uniform, small bubbles and wake effects become unimportant. When coal fines are added to the system or when the viscosity of the liquid increases, Figures 29 and 30 indicate that bubble wake volume increases with liquid velocity. In this case bubble coalescence is predominant, the bubble size increases significantly, and wake effects are important. The correlations describing K_0 as a function of U_l are given in Appendix G.

Correlation of Bubble Terminal Velocity, U_{tB} .--It has been reported previously (50) that liquid rate has little effect on the rise velocity of bubbles in the moving liquid phase. For a given fluid system, it has been found that gas velocity is the major operating variable in determining bubble size and velocity. However, from the drift flux analysis shown in Figures 25 and 26, it was found that the transition from ideal bubbly to churn turbulent is delayed as the liquid velocity increases. For this reason, the bubble terminal velocity was correlated with $U_g - U_l$, as shown in Figures 31 and 32. Liquid kerosene alone results in low bubble terminal velocities. This is consistent with the drift flux analysis illustrated in Figure 25, which indicates that nitrogen/kerosene

fluidization of HDS-2A catalyst involves gas flow in the ideal bubbly regime. On the other hand, Figure 31 indicates that with coal fines the bubble terminal velocity increases very rapidly as $U_g - U_l$ increases, again consistent with the drift flux analysis. A comparison of calculated bubble terminal velocities with coal char kerosene slurries and mineral oil is shown in Figure 32. The calculated bubble terminal velocities for mineral oil are lower than for the coal char/kerosene slurry. It is uncertain what causes this difference. As shown in Table VII, the viscosity of mineral oil at 175°F is 4.2 cp. The viscosity of coal char/kerosene slurries at 80°F is about 8 cp according to data reported in Table IX. The authors of these reports do not have great confidence in these viscosity measurements because of possible settling of coal fines. The following explanations are therefore likely:

- 1) If it is assumed that the viscosity of the coal char/kerosene slurries is indeed twice as high as that of mineral oil, the higher bubble terminal velocity is consistent with previous findings (50), indicating that higher-viscosity liquids lead to increased bubble coalescence.
- 2) If, however, the viscosity of the coal char/kerosene slurry is assumed to be equal to that of mineral oil at 175°F (based on capillary viscometer measurements and matching bed expansions), the increased bubble coalescence with coal fines may be due to promotion of rapid draining of the liquid film between two adjacent bubbles.

The work in determining the viscosity of coal char/kerosene slurries will continue in the future. In the meantime, correlations have been derived to describe the functional relationship of U_{tB} with $U_g - U_l$. These are also reported in Appendix G.

Correlation of Solids Holdup, X_k .--The application of Equation 16 along with the previously described correlations for K_o and U_{tB} could be used to predict the holdup of all phases in three-phase systems discussed in this work. It was found, however, that in certain cases the solution for ϵ_g , ϵ_l , ϵ_c did not converge to realistic values. This failure resulted from the functional dependence of X_k on the slip velocity, which in turn depends on the gas and liquid holdups. It was therefore necessary to correlate the experimental values of X_k with bubble terminal velocity as shown in Figure 33. No correlation is shown for liquid kerosene because in this case the solids holdup ratio was found to be equal to zero. For all other cases it is seen that the solids holdup ratio X_k is described by a correlation of the following form:

$$X_k = \frac{a}{U_{tB} + b} \quad (18)$$

Coefficient values are given in Appendix G.

The structure of Equation 18 was chosen to be consistent with that reported by others (38). As the viscosity of the liquid or slurry phase increases,

X_k increases as the bubble terminal velocity becomes higher. With liquid kerosene alone, X_k is negligible because the terminal velocity of small bubbles is very low.

Model Predictions.--Figures 34 and 35 show a comparison of experimental data with model predictions. In general, the agreement is good and it demonstrates that the application of this model for predicting holdups of the phases gives realistic values. The flow diagrams and Fortran coding of this predictive program appear in Appendix E. Further work is planned in the future to establish the application of this model for predicting the gas and catalyst holdup in the PDU H-Coal reactor.

Gas Mixing Model

Gas flow in gas/liquid systems has been shown (42,52) to be generally characterized by an upflow of gas bubbles in the center region of the bed and a downflow of gas bubbles near the wall region. An extension of these findings to the systems studied under this work could be used to describe the various degrees of gas mixing which are implied by the tracer data presented earlier. For this reason, a circulation model consisting of n and m completely mixed tanks is proposed. This is shown schematically in Figure 36, where n and m tanks (CSTR's) represent the gas mixing characteristics in the upflow and downflow regions, respectively. This model is in agreement with visual observation on the cold flow unit (small gas bubbles may be seen flowing down along the outer walls) and the literature (42). A general method for calculation of residence time distribution in such recycle systems has been published recently (53). A similar model was independently developed during the summer of 1979 as part of this project. This model is documented in Appendix F.

Figure 37 shows the application of this model to the residence time data for one tracer test with kerosene/no coal char. The definitions of $E(\theta)$ (the exit age distribution at the top of the catalyst bed) and θ (dimensionless time) may be found in Levenspiel (54).

The model parameters, found by trial and error, illustrate that the gas flow up through the catalyst bed is intermediate between plug flow and backmix. The 10% downflow through one CSTR illustrates the magnitude of the recycle flow. The volume ratio of 0.875 indicates that all tanks are of equal volume.

Model parameters for the 15.5 vol% coal char/kerosene slurries appear in Appendix F. This tracer work indicates the flow regimes in the ebullated bed are complex, and that a proper description of this flow is necessary to predict gas holdup and residence times.

CONCLUSIONS AND RECOMMENDATIONS

This report presents the results of work aimed at understanding the hydrodynamic behavior of the H-Coal reactor. A summary of the literature search related to the fluid dynamic behavior of gas/liquid/solid systems was presented. Design details of a cold flow unit were discussed. The process design of this cold flow model followed practices established by HRI in their process development unit.

The cold flow unit has been used to conduct experiments with nitrogen, kerosene, or kerosene/coal char slurries, and HDS catalyst, which at room temperature have properties similar to those existing in the H-Coal reactor. Mineral oil, a high-viscosity liquid, was also used. The volume fractions occupied by gas/liquid slurries and catalyst particles were determined by several experimental techniques. The use of a mini computer for data collection and calculation has greatly accelerated the analysis and reporting of data. Data on nitrogen/kerosene/HDS catalyst and coal char fines were presented in this paper.

Correlations identified in the literature search were utilized to analyze the data. From this analysis it became evident that the Richardson-Zaki correlation describes the effect of slurry flow rate on catalyst expansion.

Three-phase fluidization data were analyzed with two models. Through the use of a correlation developed by Darton and Harrison, the two bubble flow regimes were identified: ideal bubbly and churn turbulent. The effect of viscosity on the flow transition between regimes was described. Increased viscosity due to coal char addition in kerosene or use of a viscous fluid enhances bubble coalescence and the transition to churn turbulent flow. Although the Darton-Harrison model was useful in identifying flow regimes, it could not provide quantitative correlations (drift flux vs. gas holdup) because most of the data with viscous fluids fall in the churn turbulent regime. For this reason, the Bhatia and Epstein model was used to derive parameters describing the experimental data from this work. It was shown that the bed could be described by the gas bubbles, wakes following the bubbles, and the liquid/solid particulate phase. Correlations were derived describing bubble rise velocity, wake volume, and solids holdup in the wake as a function of operating parameters. Predictions with this model appear realistic.

One of the significant findings in this work is the observation by Battelle that H-Coal fluids are non-Newtonian. This should be further explored with additional measurements of H-Coal liquids.

It is recommended that an experimental program be carried out in the PDU reactor in which the ebullated bed height will be externally monitored with changes in operating conditions at low, medium, and high viscosities. These viscosities will be achieved via different slurry recycle rates. Additional studies at ambient temperatures are also needed to establish the bubble size along with direct measurements of the bubble rise velocity.

Use of optically clear solid/liquid systems will permit the use of laser techniques for this purpose. Finally, since this study established that coal char is uniformly distributed in the liquid phase, additional liquid tracer tests will be useful to derive mixing parameters for the liquid phase.

ACKNOWLEDGMENTS

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NOMENCLATURE

Typical
Units

A	Cross-sectional area of column	cm ²
C ₀	Drag coefficient for single particle or bubble	--
C _{0M}	Drag coefficient for multi-particle system	--
d	Particle diameter	cm
$\bar{d}$	Average particle diameter	cm
d _e	Equivalent spherical diameter of bubble	cm
d _p	Diameter of a circle if the same area as the projected particle when lying in its most stable position	cm
d _s	Diameter of a sphere with the same volume as the particle	cm
d _{sm}	Sauter-mean bubble diameter	cm
D	Bed or tube diameter	cm
D _c	Diameter of capillary	cm
E _g	Dispersion coefficient	sec/cm ²
f	Frequency of formation of bubble cluster	sec ⁻¹
Fr	Froude number	--
g	Acceleration of gravity	cm/sec ²
G	Dimensionless group	--
H _{GL}	Three-phase bed height	cm
H _L	Liquid/solid bed height	cm
j _l	Liquid flux	gm/cm ² sec
j _l *	Dimensionless liquid flux	--
$\bar{k}$	Average of wake volume to bubble volume ratio	--
K	Shape factor	--
K'	Effective hydrodynamic volume of particle	cm ³
K ₀	Single bubble wake volume to bubble volume ratio	--
l	Particle length	cm
M	Morton number	--
M'	Rheological parameter	gm/cm sec

		Typical Units
n	Richardson-Zaki index or exponent in ideal bubbly flow regime models	--
n'	Rheological parameter	--
Pe	Peclet number	--
Qg	Gas volumetric flow rate	cm ³ /sec
r	Particle or bubble radius	cm
r*	Dimensionless particle radius	--
re	Equivalent spherical bubble radius	cm
ro	Orifice radius	cm
R	Radius of curvature	cm
Re	Particle Reynolds number	--
Reb	Bubble Reynolds number	--
Ref	Minimum fluidization particle Reynolds number	--
Rem	Reynolds number for multi-particle system	--
Ret	Particle Reynolds number based on U _t	--
Ub	Bubble rise velocity	cm/sec
Ug	Superficial gas velocity	cm/sec
U _l	Superficial liquid velocity	cm/sec
Um	Mean velocity	cm/sec
Ur	Relative velocity between particles and liquid	cm/sec
Us	Gas/liquid slip velocity	cm/sec
Ut	Terminal velocity of an isolated particle or bubble	cm/sec
Utb	Bubble terminal velocity	cm/sec
U ₁₀	Superficial liquid velocity at incipient fluidization	cm/sec
U _{s1}	Velocity of gas slug	cm/sec
U _l '	Superficial liquid velocity in the particulate fluidized phase in a three-phase system	cm/sec
v _b	Bubble volume	cm ³

V_g	Gas linear velocity	
V_{cd}	Gas drift flux	cm/sec
We	Weber number	--
X_k	Ratio of solids holdup in wake to solids holdup in particulate phase	--

Greek

β	Number of small bubbles forming a cluster	--
ϵ	Bed voidage	--
ϵ_g	Volume fraction of gas	--
ϵ_l	Volume fraction of liquid	--
ϵ_s, ϵ_c	Volume fraction of solids (catalyst particles)	--
ϵ_w	Volume fraction of wake phase	--
δ	Pore diameter of gas distributor	cm
ρ_g	Gas density	gm/cm ³
ρ_s	Density of particles	gm/cm ³
ρ_l	Density of liquid	gm/cm ³
ρ_b	Density of fluidized bed	gm/cm ³
τ_w	Wall shear stress	dynes/cm ²
μ	First moment	sec
μ_e	Effective viscosity	poise
μ_l	Liquid viscosity	poise
σ	Surface tension	dynes/cm
σ^2	Second moment	sec ²
λ	Wavelength of disturbance	cm
ϕ	Solids volume fraction	--

Subscripts

k	Wake phase
f	Liquid fluidized phase (particulate phase)
l	Liquid

Nomenclature

Subscripts (Continued)

g Gas

c Catalyst

Superscripts

11 Two-phase region

111 Three-phase region

TABLE I

LIST OF PUBLICATIONS DURING THE PROGRAM

- Vasalos, I. A., et al., Monthly Progress Report No. 1, FE-2588-1, October, 1977.
- Vasalos, I. A., et al., Monthly Progress Report No. 2, FE-2588-2, November, 1977.
- Vasalos, I. A., et al., Quarterly Progress Report No. 1, FE-2588-3, December, 1977.
- Vasalos, I. A., et al., Monthly Progress Report No. 4, FE-2588-4, January, 1978.
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- Vasalos, I. A., et al., Quarterly Progress Report No. 3, FE-2588-10, June, 1978.
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- Vasalos, I. A., et al., Monthly Progress Report No. 13, FE-2588-13, October, 1978.

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Vasalos, I. A., et al., Monthly Progress Report No. 14, FE-2588-14, November, 1978.

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Vasalos, I. A., et al., Monthly Progress Report No. 21, FE-2588-21, May, 1979.

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Vasalos, I. A., et al., Monthly Progress Report No. 22, FE-2588-23, July, 1979.

Vasalos, I. A., et al., Monthly Progress Report No. 23, FE-2588-24, August, 1979.

Vasalos, I. A., et al., Monthly Progress Report No. 25, October, 1979.

Vasalos, I. A., et al., Monthly Progress Report No. 26, November, 1979.

Vasalos, I. A., et al., Monthly Progress Report No. 27, December, 1979.

Vasalos, I. A., E. M. Bild, and D. N. Rundell, "Experimental Techniques for Studying the Fluid Dynamics of the H-Coal Reactor," 72nd Annual AIChE Meeting, San Francisco, November 25-29, 1979.

TABLE II

CORRELATIONS FOR THE EXPANSION OF LIQUID FLUIDIZED BEDS

Author	Solids	Equation	Range of Applicability
Richardson and Zaki (6)	Spherical particles	$\epsilon_1^n = U_1/U_t$ where: $n = 4.65 + 20 d/D$ $n = (4.4 + 18 d/D) Re_t^{-0.03}$ $n = (4.4 + 18 d/D) Re_t^{-0.1}$ $n = 4.4 Re_t^{-0.1}$ $n = 2.4$	$Re_t < 0.2$ $0.2 < Re_t < 1.0$ $1 < Re_t < 200$ $200 < Re_t < 500$ $Re_t > 500$
Richardson and Zaki (6)	Glass or steel cylinders and steel plates	$\epsilon_1^n = U_1/U_t$ $n = 2.7 K^{0.18}$ where: $K = (\pi/6) d_s^3/d_p^3 = (\pi/4)^{5/2} (d/l)^{1/2}$ (for a cylinder)	$Re_t > 500$
Fouda and Capes (8)	Irregularly shaped particles	$(1 - K' \epsilon_s)^n = U_1/U_t$ where: $n = \text{same as in Richardson-Zaki correlation}$ $K' = 1$ for spheres $K' = 1.17$ to 1.43 for equi-dimensional but irregularly shaped particles	Same as in Richardson-Zaki correlation
Barnea and Mizrahi (9)	Spherical particles	$C_{DM} = \left(0.63 + \frac{4.8}{\sqrt{Re_m}}\right)$ where: $Re_m = Re \left(\frac{U_r/U_t}{\exp(5\epsilon_s/3(1-\epsilon_s))} \right)$ $C_{DM} = C_0 \left(\frac{U_r}{U_t} \right)^2 \left(\frac{1 - \epsilon_s}{1 + \epsilon_s^{1/3}} \right)$ $\frac{U_r}{U_t} = \frac{1}{1 + \epsilon_s^{1/3}}$	$10^{-3} < Re_m < 3 \times 10^4$

(Table Continued)

TABLE LI
CORRELATIONS FOR THE EXPANSION OF LIQUID FLUIDIZED BEDS
-2-

Author	Solids	Equatlor.	Range of Applicability
Wallis (10)	Spherical particles	$j_1^* = 2/9 \ r^{*2} \epsilon_1^{4.7}$	$Re_m < 1.5$
		$j_1^* = 0.307 \ r^{*1.21} \epsilon_1^{3.47}$	$1.5 < Re_m < 100$
		$j_1^* = 0.693 \ r^{*0.858} \epsilon_1^{2.92}$	$100 < Re_m < 1080$
		$j_1^* = 2.5 \ r^{*1.2} \epsilon_1^{2.35}$	$1080 < Re_m$
where:			
$j_1^* = j_1 \left(\frac{\rho_1}{\mu_1 g (\rho_s - \rho_1)} \right)^{1/3}$			
$r^* = \left(\frac{\rho_1 g (\rho_s - \rho_1)}{\mu_1} \right)^2$			

TABLE III
EXPERIMENTAL CONDITIONS FOR BUBBLE COALESCENCE

<u>Investigator</u>	<u>Liquid</u>	<u>Gas</u>	<u>Solid</u>	<u>Solid Dimension (mm)</u>	<u>Bed Diameter (mm)</u>	<u>Character of Bed</u>
Ostergaard (17)	Water	Air	Sand	0.64	250 x 250	Coalescing
Massimilla, et al. (18)	Water	Air	Sand	0.22, 0.26	86 x 61	Coalescing
	Water	Air	Glass Beads	0.79, 1.09	86 x 61	Coalescing
Viswanatan, et al. (19)	Water	Air	Quartz Particles	0.649, 0.928	50.8	Coalescing
	Water	Air	Glass Beads	4.0	50.8	Non-Coalescing
Adlington and Thompson (2)	Water	-	Sand	3.0	76.0	Coalescing
	White Spirit	-	Alumina	0.3-3.0	76.0	Coalescing
Lee (21)	Water	Air	Glass Beads	6		Non-Coalescing

TABLE IV

SUMMARY OF DATA FOR GAS/LIQUID/SOLID FLUIDIZATION

<u>Investigators</u>	<u>System</u>	<u>Particle Size</u>	<u>Experimental Unit</u>	<u>Parameters Studied</u>	<u>Measured Quantities</u>
1) Razumov, Manshilin, Nemets(22)	Air/Water/Sand Air/Water/Glass Beads	0.493 to 1.27 mm	300 mm Diameter Column	U_g, U_l	ϵ_g
2) Michelsen, Ostergaard(23)	Air/Water/Glass Beads	1, 3, 6 mm	152 mm Diameter Column	U_g, U_l , Particle Size	ϵ_g, ϵ_l
3) Dakshinamurty, Subramanyam, Rao(24,25)	Air/Water/Various Spherical Particles Air/Kerosene/Various Spherical Particles	1.05 to 6.8 mm	56 mm Diameter Column	Particle Size and Density σ, U_g, U_l	ϵ_g
4) Viswanathan, Kakai, Murti(19)	Air/Water/Quartz Particles Air/Water/Glass Beads	0.928, 0.649 mm 4 mm	50.8 mm Diameter Column	U_g, U_l , Particle Size	ϵ_l, ϵ_g
5) Kim, Baker, Bergougnou(11)	Air/Water/Glass Beads Air/Water/Irregular Gravel	6 mm 2.6 mm	660x25 mm Rectangular Channel	U_g, U_l	ϵ_l, ϵ_g , Bubble Size and Velocity
6) Kim, Baker, Bergougnou(26)	Air-Water/Acetone-Glass Beads Air-Water/Acetone-Irregular Gravel Air-Sugar/Water-Glass Beads Air-Carboxymethyl Cellulose/Water-Glass Beads Air-Sugar/Water-Irregular Gravel Air-Carboxymethyl Cellulose/Water-Irregular Gravel	1.6 mm 2.6 mm 1.6 mm	660x25 mm Rectangular Channel	U_l, σ, U_l, U_g	ϵ_g, ϵ_l
7) Blum and Toman(27)	Nitrogen/Light Mineral Oil	3.2x3.2mm	102 mm Diameter Column	U_l, U_g, U_l	ϵ_g
8) Nemets, Razumov, Manshilin(28)	Air/Water/Sand Air/Heptane/Sand Air-Water/Glycerol-Sand	0.920 mm	90 mm Diameter Column	U_l, U_g, U_l	ϵ_l, ϵ_g
9) Bruce, Revel-Chlon(29)	Air/Water/Glass Spheres	2, 4, 6, 8 mm	46.3 mm Diameter Column	U_l, U_g	ϵ_g , Bubble Size
10) Ostergaard, Thiesen (30)	Air/Water/Glass Balloons	0.28 to 2.2 mm	51 and 102 mm Diameter Column	U_l, U_g	ϵ_g

TABLE V

EMPIRICAL CORRELATIONS FOR THREE-PHASE BEDS

Author	Correlation	Gas/Liquid	Solids (Dimension)	Column Diameter or Dimensions (mm)	Comments										
Kim, Baker, Bergougnou(11)	$(e_1)_{U_g=0} - e_1 = 0.0025 \left(Fr_1 \frac{f_2}{f_1} \right)^{0.140} \left(Fr_2 \frac{f_2}{f_1} \right)^{0.161} \left(\frac{Re_1}{Re_2} \right)^{0.280}$ $(e_1)_{U_g=0} = 0.409 (Fr_1 f_2 / f_1)^{3.183} (Re_1)^{0.074}$	Air/Water	Glass Beads (6 mm)	660 x 25	Empirical, no consideration for bed contraction										
Kim, Baker, Bergougnou(26)	$(e_1 + e_g) = 1.40 (Fr_1)^{0.17} (We)^{0.070} \text{ (Expanding Beds)}$ $(e_1 + e_g)_{U_g=0} = 1.3 (Fr_1)^{0.120} (We)^{0.073} \exp[0.31 U_1 / U_g (e_1)_{U_g=0}] \text{ (Contracting Beds)}$ $(e_1)_{U_g=0} = 1.353 (Fr_1)^{0.200} (Re_1)^{-0.082} (We)^{0.082}$	Air/Sugar Solutions Air/Carboxymethyl Cellulose Solution Air/Water Acetone	Glass Beads (6,1 mm) Irregular Gravel (2.6 mm)	660 x 25	Empirical for expanding and contracting beds										
Dakshinamurty, Subrahmanyam, Rao (24,25)	$(e_g + e_1) = \left(\frac{U_1}{K U_c} \right)^m \left(\frac{U_1 U_g}{\sigma} \right)^n$ <p>$n = 0.08$ $K = 2.12, m = 0.41, Re_c < 500$ $K = 2.65, m = 0.6, Re_c > 500$</p>	Air/Water Air/Kerosene	Numerous Diameters 1.06 to 6.8 mm	56 mm Diameter	Empirical, no consideration for bed contraction										
Blum, Toman(27)	$\frac{(e_g + e_1) - (e_1)_{U_g=0}}{1 - (e_1)_{U_g=0}} = f(U_g)$	Nitrogen/Light Mineral Oil	<table><tr><th colspan="2">Cylinders</th></tr><tr><th>Dia, mm</th><th>Length, mm</th></tr><tr><td>3.2</td><td>3.2</td></tr><tr><td>4.8</td><td>2.4</td></tr><tr><td>4.8</td><td>4.8</td></tr></table>	Cylinders		Dia, mm	Length, mm	3.2	3.2	4.8	2.4	4.8	4.8	102	Empirical, no consideration for bed contraction
Cylinders															
Dia, mm	Length, mm														
3.2	3.2														
4.8	2.4														
4.8	4.8														
Razumov,	$e_g = 0.578 - 3.198 U_1 - 0.538 U_g$ $e_1 = 0.422 + 0.135 U_1 / d^{0.660} - 1.82 U_g$ $e_g = K(1 - e_s)^{0.200} (U_g / U_1)^{0.700}$	Air/Water	Sand, Slag Beads 0.49 to 1.27 mm	300 mm	Empirical, no consideration for bed contractions										

BHATIA-EPSTEIN MODEL

Equations

1) Tabulation of $\epsilon_k''/\epsilon_g''$ vs. ϵ_g (Letan and Kehat*).

$$2) \epsilon_1 = \epsilon_k(1 - X_k) + \epsilon_1''_f(1 - \epsilon_g - \epsilon_k + X_k\epsilon_k)$$

$$3) \epsilon_c + \epsilon_g + \epsilon_1 = 1$$

$$4) U_g = v_g \epsilon_g$$

$$5) \epsilon_1''_f = \left[\frac{U_1 - v_g \epsilon_k(1 - X_k)}{U_t(1 - \epsilon_g - \epsilon_k)} \right]^{1/n}$$

$$6) v_g = \frac{u_1 + U_g + \epsilon_1''_f(1 - \epsilon_g - \epsilon_k)v_{g1}'''}{1 - \epsilon_c}$$

v_{g1} = relative velocity between bubble phase and liquid
in particulate phase

$$7) \epsilon_k = \frac{\epsilon_g \epsilon_k''}{\epsilon_g''} (1 - \epsilon_c)^3$$

$$8) v_{g1}''' = U_{t_0} + 2 U_g$$

*Letan, R., and E. Kehat, AIChE J, 14, 398, 1969.

TABLE VII

PHYSICAL PROPERTIES OF LIQUIDS USED IN COLD FLOW STUDIES

<u>Viscosity (cp)</u>	<u>50°F</u>	<u>70°F</u>	<u>74°F</u>	<u>100°F</u>	<u>150°F</u>	<u>175°F</u>
Water		1.0		1.0	1.0	
Kerosene		1.39		1.15	0.8	
Mineral Oil		22.4		14.6	6.08	4.2
<u>Surface Tension (Dynes/cm)</u>						
Water	77.6	--	75.5	--	68.3	
Kerosene	30.0	--	28.6	--	24.6	
Mineral Oil	34.0	--	32.6	--	29.1	
<u>Density, g/cc</u>						
Water		0.99		0.99	0.98	
Kerosene		0.79		0.78	0.77	
Mineral Oil		0.85		0.84	0.82	

TABLE VIII

SUMMARY OF GAS/KEROSENE DATA

Temperature = 76°F

<u>Gas</u>	<u>Pressure, psi</u>	<u>cDensity, g/ml</u>	<u>Surface Tension, Dynes/cm.</u>	<u>Remarks</u>
Air	0	0.81	--	
Nitrogen	2	0.82	35.1	Surface tension average of 24.1/24.0/24.2
	15	0.82	26.8	Surface tension average of 27.2/26.5/26.7
Helium	2	0.82	26.8	Surface tension average of 26.9/26.8/26.9
	15	0.82	29.1	Surface tension average of 28.9/29.2/29.3
Freon-12	2	0.84	23.4	Surface tension average of 23.3/22.9/23.8
	15	0.85	20.7	Surface tension average of 20.8/20.5/20.8

TABLE IX

VISCOSITIES OF COAL CHAR/KEROSENE SLURRIES

<u>Temperature, °F</u>	<u>70</u>	<u>72</u>	<u>73</u>	<u>78</u>	<u>86</u>	<u>100</u>	<u>112</u>	<u>143</u>	<u>150</u>	<u>212</u>	<u>Measured By</u>
Kerosene	1.39					1.15			0.80		Rotary Drilling Battelle
Kerosene					1.38		1.05		0.88		
Kerosene/5.1 Vol% Coal Char			3.25			3.0			2.5		Rotary Drilling Battelle
Kerosene/5.1 Vol% Coal Char			3.80			3.1		2.2			
Kerosene/10.5 Vol% Coal Char			6.5			4.5			3.5		Rotary Drilling Battelle
Kerosene/10.5 Vol% Coal Char			4.2			3.2			2.2		
Kerosene/11.9 Vol% Coal Char		3.5									Amoco
Kerosene/15.5 Vol% Coal Char				3.7							Amoco
Kerosene/17.8 Vol% Coal Char			9.25			8.25			6.75		Rotary Drilling Battelle
Kerosene/17.8 Vol% Coal Char			7.20				3.6	3.3			

Note: Viscosities in centipoises.

TABLE X
COMPARISON OF COAL CHAR WITH H-COAL REACTOR FINES

	<u>Coal Char</u>	<u>Recycle Oil Solids (Pyridene Insolubles)</u>
Particle Density, g/cc	1.7	2.3
Density Distribution	90% 0.8-2.2 g/cc. Even distribution.	85% 1.2-2.8 g/cc. Even distribution.
Particle Size	70%--325 Mesh.	90%--325 Mesh.

Source of Information: Reference 35.

CUMULATIVE SIZE DISTRIBUTION OF COAL CHAR

CUMULATIVE NUMBER AND NUMBER %
GREATER THAN STATED SIZE

<u>Size, μm</u>	<u>Cumulative Number</u>	<u>Cumulative Number %</u>
0	2916	100
1.1	2635	90.4
2.7	1921	65.9
3.8	1396	47.9
5.4	922	31.6
8.1	529	18.1
13.5	212	7.3
18.7	115	3.9
29.7	64	2.2
51.3	24	0.8
72.8	13	0.4
94.4	8	0.3

TABLE XII

PROPERTIES OF HDS-2A CATALYST

Nominal Diameter: 1.6 mm
Nominal Length: 4.8 mm

Particle Density, g/cc	1.0794
Pore Volume, cm/g	0.66
Surface Area, m ² /g	296
<u>Length Distribution (mm)</u>	
Below 3.8	1
3.8-4.0	0
4.0-4.2	2
4.2-4.4	6
4.4-4.6	9
4.6-4.8	5
4.8-5.0	4
5.0-5.1	7
5.1-5.3	9
5.3-5.5	4
5.5-5.7	0
5.7-5.9	1
5.9-6.1	3
6.1-6.3	4
Over 6.3	5
Average	5.1

TABLE XIII

SUMMARY OF EXPERIMENTAL RUNS

<u>Date</u>	<u>Run No.</u>	<u>Catalyst</u>	<u>Cat. L/D</u>	<u>Gas</u>	<u>Liquid</u>	<u>Fines, Vol%</u>	<u>Temp, °F</u>
8/25/78	200	None	-	Nitrogen	Kerosene	0	72
8/30	201	HDS-2A	3	"	"	0	72
10/12	203	"	"	"	"	1.0	68
10/20	204	"	"	"	"	5.1	70
11/1	205	"	"	"	"	10.4	67
12/13	206	"	"	"	"	15.5	85
1/11/79	207	"	"	Freon-12	"	15.5	81
1/23	208	"	"	Nitrogen	"	15.2	148
2/2	209	"	"	"	"	15.5	97
2/22	210	"	"	"	"	11.9	71
2/16	211	"	"	"	"	17.8	80
4/12	212	"	"	"	"	0	71
4/17	213	"	"	Helium	"	0	70
5/17	214	"	"	Nitrogen	"	15.5	150
5/30	215	"	"	"	"	15.5	71
6/15	216	"	"	Helium	"	15.5	73
6/14	217	"	"	Nitrogen	"	15.5	73
3/13/79	300	None	-	Nitrogen	Kerosene	0	84
3/29	301	HDS-2A	6	"	"	0	74
6/21	310	"	3	"	"	0	72
6/26	311	"	"	Helium	"	0	77
7/2	320	"	2	"	"	0	75
7/11	321	"	"	Nitrogen	"	0	77
7/24/79	400	None	-	Nitrogen	Mineral Oil	0	100
7/25	401	"	-	"	"	0	125
7/27	410	"	-	Helium	"	0	125
8/1	420	HDS-2A	3	Nitrogen	"	0	100
8/2	421	"	"	"	"	0	125
8/3	422	"	"	"	"	0	150
8/9	423	"	"	"	"	0	175
8/13	424	"	"	Helium	"	0	177
8/15	425	"	"	"	"	0	150
8/22	426	"	"	Nitrogen	"	0	154
8/27	427	"	"	"	"	0	99
8/28	428	"	"	"	"	0	127
10/4	429	"	"	"	"	0	176
10/9	430	None	-	"	"	0	170
10/11	431	"	-	"	"	0	149

TABLE XIV
DATA FOR PDU LIQUID SAMPLES

DATE		7/16/79	7/18/79	7/18/79	7/18/79
TIME		1530	1030	1100	1630
SAMPLE NO.		1	2	3	4
PDU RUN NO.		130-88	130-88	130-88	130-88
PDU PERIOD NO.		11A	13A	13A	13A
<u>DATA</u>					
Catalyst Bed Expansion		150%	150%	150%	150%
<u>FEED RATE</u>					
Coal	Avg.	(321)	(323)	(323)	(323)
	Instant	342 #/Hr.	341 #/Hr.	341 #/Hr.	341 #/Hr.
Liquid	Avg.	(637)	(655)	(655)	(655)
	Instant	675 #/Hr.			
Gas	Avg.	(5820)	Avg.	Avg.	Avg.
	Instant	5600 SCFH	6152 SCFH	6152 SCFH	6152 SCFH
Coal Type		Illinois #6	Illinois #6	Illinois #6	Illinois #6
Coal Concentration					
Toluene Extraction		92.8	92.5	92.5	92.5
Conversion(DMF Filtration)		(89.7%)	(88.9)	(88.9)	(88.9)
Catalyst Type		HDS-1442A			
Reactor Temperature	Avg.	840	848	848	848
	Instant	848°F	847°F	847°F	848°F
Reactor Pressure	Avg.	2600	2623	2623	2623
	Instant	2600 PSIG	2610 PSIG	2610 PSIG	2623

←----- Amoco Bombs -----→

7/18/79

TABLE XV

COAL CHAR DISTRIBUTION ALONG THE REACTOR:
HDS-2A CATALYST, L = 4.8 MM, D = 1.6 MM

Test No.	Fines* Vol-%	Slurry Flow Cm/Sec	Gas Flow Cm/Sec	Concentration in Wt%				Recycle Line	Separ- ator Feed	Average
				Reactor Bottom	152 Cm	305 Cm	457 Cm			
204-23	5.1	5.5	1.1	7.4	7.5	6.7	11.9	--	--	8.4
205-26	10.4	3.0	7.6	14.5	14.7	16.4	16.7	--	--	15.6
206-05	15.5	4.3	0	21.8	22.0	21.8	20.5	20.3	16.3	21.5
211-07	17.8	3.0	4.6	21.8	24.6	25.2	25.2	23.3	22.2	24.2
211-11	17.8	3.0	6.1	23.9	24.4	25.9	24.8	23.9	22.6	24.7
210-09	11.9	3.0	3.0	17.2	17.2	17.2	17.2	--	--	17.2

TABLE XVI

PARTICLE SIZE DISTRIBUTION OF REACTOR COAL CHAR SAMPLES

Slurry Concentration = 17.8 vol%
Liquid Velocity, Cm/Sec = 3.0
Gas Velocity, Cm/Sec = 4.6

CUMULATIVE NUMBER AND CUMULATIVE NUMBER PERCENT
GREATER THAN STATED SIZE

Size, μm	Reactor Bottom Sample AU77-13		457 cm Level Sample AU77-16	
	Cumulative Number	Cumulative Number %	Cumulative Number	Cumulative Number %
0	4409	100	4142	100
1.1	3588	81.4	3220	77.7
2.7	2485	56.4	2109	50.9
3.8	1971	44.7	1632	39.4
5.4	1472	33.4	1148	27.7
8.1	1017	23.1	714	17.2
13.5	542	12.3	316	7.6
18.9	291	6.6	164	4.0
29.7	113	2.6	61	1.5
51.3	32	0.7	8	0.2
70.2	5	0.1	4	0.1
91.8	2	0.05	2	0.05
Max. Size, μm	-----	108.0 -----	-----	118.0 -----
Avg Size, μm	-----	3.5 -----	-----	2.8 -----

TABLE XVII

VARIATION IN RICHARDSON-ZAKI INDEX

<u>Run No.</u>	<u>Test Nos.</u>	<u>Liquid</u>	<u>Fines, Vol%</u>	<u>Temp, °F</u>	<u>Index n</u>
201	2-11	Kerosene	0	72	2.68
203	1-4	"	1	68	3.07
204	1-6	"	5.1	70	3.44
205	1-8, 35	"	10.4	67	3.57
206	1-9	"	15.5	85	3.27
208	1-6, 26-28	"	15.5	148	3.42
209	1-4, 24-25	"	15.5	97	3.22
210	1, 7, 13, 19	"	11.9	71	3.58
211	1, 7, 13	"	17.8	80	3.71
212	1, 3, 7, 10	"	0	71	3.58
214	1-11	"	15.5	150	4.98
310	1-5, 6, 13	Kerosene	0	72	2.46
320	1-5	"	0	75	3.84
427	3-6	Mineral Oil	0	100	3.8
428	3-5	"	0	125	3.5
426	4-7	"	0	150	3.4
423	2-4	"	0	175	3.5

IAV/ml
2/20/80

TABLE XVIII

FIRST AND SECOND MOMENTS CALCULATED BY TWO METHODS

<u>Test</u>	<u>Detector</u>	<u>m (1)</u> <u>Sec</u>	<u>mm (1)</u> <u>Sec</u>	<u>m (2)</u> <u>Sec²</u>	<u>mm (2)</u> <u>Sec²</u>
1a	1	--	--		
	2	22.9	22.9	117	120
	3	47.6	47.5	785	786
	4	75.7	74.4	1170	1300
	5	90.5	90.0	1502	1600
2a	1	19.6	19.5	--	850
	2	25.2	25.1	1088	1100
	3	48.6	48.4	2184	2160
	4	78.4	77.4	3040	3048
	5	95.2	95.0	3860	3900
3a	1	14.8	14.8	94.8	95.0
	2	24.2	24.2	394	380
	3	53.8	53.4	2560	2520
	4	69.6	69.0	2604	2608
	5	73.2	72.6	2632	2636

m (1), m (2): Moments calculated using Equations C-2 and C-3.

mm (1), mm (2): Moments calculated using the modified method.

TABLE XIX

CALCULATED GAS LINEAR VELOCITIES

Kerosene, Nitrogen, American Cyanamid
HDS-2A Catalyst, $l = 4.8$ mm, $d = 1.6$ mm

<u>Test</u>	<u>U_1, cm/sec</u>	<u>U_g, cm/sec</u>	<u>Vol% Fines</u>	<u>Tracer</u>		<u>Gamma-Ray</u>	
				<u>V_g in Bed cm/sec</u>	<u>V_g Above Bed cm/sec</u>	<u>V_g in Bed cm/sec</u>	<u>V_g Above Bed cm/sec</u>
1a	3.0	3.0	0	13.6	5.6	15.8	12.5
1b	3.0	3.0	0	11.9	5.6	15.8	12.5
2a	3.0	3.0	15.5	14.1	6.2	21.4	25.0
2b	3.0	3.0	15.5	9.8	5.5	21.4	25.0
3a	3.0	4.6	15.5	14.8	10.0	35.3	30.7
3b	3.0	4.6	15.5	14.8	9.5	35.3	30.7

TABLE XX
GAS HOLDUPS CALCULATED FROM GAS TRACER
AND GAMMA-RAY TESTS

<u>Test</u>	<u>U_l, cm/sec</u>	<u>U_g, cm/sec</u>	<u>Vol% Fines</u>	<u>Tracer</u>		<u>Gamma-Ray</u>	
				<u>In Bed</u>	<u>Above Bed</u>	<u>In Bed</u>	<u>Above Bed</u>
1a	3.0	3.0	0	0.22	0.54	0.19	0.24
2a	3.0	3.0	15.5	0.22	0.49	0.14	0.12
3a	3.0	4.6	15.5	0.31	0.46	0.13	0.15

TABLE XXI
RESULTS OF GAMMA-RAY SCANS THROUGH CHORDS
OF THE CROSS-SECTION

<u>U_l,</u> <u>cm/sec</u>	<u>U_g,</u> <u>cm/sec</u>	<u>Position</u>			<u>Delta</u>
		<u>1</u>	<u>2</u> <u>ε_g</u>	<u>3</u>	
4.6	6.1	0.19	0.18	--	±0.02
5.8	3.0	0.11	0.11	0.18	±0.03

TABLE XXII
CALCULATION OF DISPERSION COEFFICIENT

<u>Test</u>	<u>Detectors</u>	<u>Gas Velocity, cm/sec</u>	<u>P_e</u>	<u>E_g*</u>
1a	1-2	13.6	5.1	492
	3-4	5.6	3.4	260
2a	1-2	14.1	0.78	3354
	3-4	6.2	2.39	408
3b	1-2	14.8	1.40	1960
	3-4	9.5	1.65	910

*E_g in sec/cm².

TABLE XXIII

SOLUTION OF THE BHATIA-EPSTEIN MODEL

- 1) Select initial values of ϵ_c , ϵ_1 , ϵ_g .
- 2) Compute: $V_{g1}''' = U_{t_0} + 2 U_g$.
- 3) Compute: $\epsilon_k''/\epsilon_g''$ - values given by Letan and Kehat.
- 4) Compute: $\epsilon_k = \epsilon_g \frac{\epsilon_k''}{\epsilon_g''} (1 - \epsilon_c)^3$
- 5) Compute: $V_g = U_g/\epsilon_g$
- 6) Compute: ϵ_{1f}'' from Richardson-Zaki relationship
- 7) Compute: $V_g = \frac{U_1 + U_g + \epsilon_{1f}'' (1 - \epsilon_g - \epsilon_k) V_{g1}'''}{1 - \epsilon_c}$
- 8) Compute: $\epsilon_{12} = \epsilon_k(1 - X_k) + \epsilon_{1f}''(1 - \epsilon_g - \epsilon_k + X_k \epsilon_k)$
- 9) Compute new ϵ_g : $\epsilon_{g2} = U_g/V_g$
- 10) Average: $\epsilon_{g1} + \epsilon_{g2} = \epsilon_{gavg}$
- 11) Compute new ϵ_c from ϵ_{gavg} and ϵ_{12}

Iterate until $\Delta\epsilon_g$ $\Delta\epsilon_1$ are small.

TABLE XXIV

BED SETTLING RATECatalyst $l/d = 3$

Gas: Nitrogen

Test No.	U_1 , Ft/Sec	U_g , Ft/Sec	Liquid	Fines, Vol%	Temp, °F	Viscosity, °F	Bed Settling Rate, Ft/Sec	% Bed Expansion	Particle Terminal Velocity, Ft/Sec	
									(1)	(2)
212-05	0.10	0.10	Kerosene	0	71	1.39	0.17	54	0.45	--
217-01	0.10	0.10	"	15.5	150	6.75	0.12	72	0.45	--
426-13	0.10	0.10	Mineral Oil	0	150	6.08	0.16	107	0.36	0.30
423-12	0.10	0.10	"	15.5	175	4.2	0.17	75	0.35	0.42
217-05	0.15	0.10	Kerosene	15.5	150	6.75	0.23	156	0.45	--
423-17	0.15	0.10	Mineral Oil	0	175	4.20	0.22	153	0.35	0.30
426-17	0.15	0.10	"	0	150	6.08	0.22	203	0.36	0.42

(1) Extrapolated terminal velocity from Richardson-Zaki correlation.

(2) Experimentally measured value.

Figure 1
H-Coal PDU reactor

Figure 2

Drift flux vs. gas holdup: Darton and Harrison (5)
(Gas/liquid/solid fluidization)

Figure 3
Schematic diagram of the fluid dynamics unit

Figure 4
Schematic diagram of experimental
for viscosity measurement

Figure 5

Gamma ray scans

liquid velocity 0.07 ft/sec

Figure 6

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Radiotracer detector location

Figure 7

Coal fines settling rate

Figure 8

Definition of Bingham fluid

$$\sigma = \theta + \eta' \dot{\gamma}$$

IF $\sigma < \theta$, no flow ($\dot{\gamma} = 0$)

Figure 9

Effect of temperature on viscosity of
H-coal sample, 2500 psi

Figure 10

Variation in bed expansion with temperature and coal fines concentration

Figure 11

Effect of viscosity on bed expansion

— — Mineral oil — —

Figure 12

Bed expansion with kerosene slurries and mineral oil

Figure 13

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Effect of catalyst particle properties on bed expansion

Figure 14

Correlation of liquid — solid data

Figure 15

Effect of operating conditions on bed expansion

Figure 16

Effect of temperature on bed expansion

Figure 17

Bed expansion - comparison of various liquids

Figure 18

Effect of particle size on bed expansion

Figure 19

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Bed expansion — effect of gas type

Figure 20

Bed expansion — — Effect of gas type — — Mineral oil — —

Figure 21

Gas holdup — effect of coal fines

Figure 22

Gas holdup — comparison of various liquids

Figure 23

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Tracer results with 0 vol % coal char

Figure 24

Tracer results with 15.5 vol % coal char

Figure 25

Drift flux — effect of operating conditions

Figure 26
Drift flux — effect of liquid viscosity

Figure 27

Bed expansion vs. gas velocity

Comparison of HRI PDU run 7 results with 17.8 vol % coal char in kerosene

Figure 28

Variation of bubble included angle with liquid viscosity

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Figure 29

Wake volume — effect of coal fines

Figure 30

Wake volume — effect of liquid type

Figure 31

Bubble terminal velocity — effect of coal fines

Figure 32

Bubble terminal velocity — effect of liquid type

Figure 33

Relative solids holdup — effect of liquid type

Figure 34

Catalyst holdup — predicted vs actual — kerosene and 17.8 vol % coal char

Figure 35

Gas holdup — predicted vs actual — kerosene and 17.8 vol % coal char

Figure 36
Gas mixing model

Figure 37

Gas mixing model Kerosene/zero fines

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APPENDIX A
EQUIPMENT DETAILS

APPENDIX A

EQUIPMENT DETAILS

The objective of this appendix is to give more details on various system components.

Reactor Components

The reactor is constructed from four glass sections 5' in length and 6" in diameter. The glass sections are connected through flanges to five metal spool pieces. A schematic diagram of the reactor is shown in Figure A-1.

The spool pieces have entries for sample taps, pressure taps, and thermowells to monitor the system. A drawing of a spool piece is shown in Figure A-2. As shown in this figure, the system is designed to insert two pressure taps through each spool piece. Thus, a total of eight DP measurements is feasible with this design. The lines are maintained clear by bleeding a low gas rate through each tap. Use of this arrangement resulted in significant vibrations in the lines and the pressure drop measurements. For this reason, four new pressure taps were rigidly connected to the spool pieces. Although this resulted in reliable measurements, pressure drop measurements are possible only every five feet.

A sample probe can also be inserted into the reactor through each spool piece. A drawing of the sample probe assembly is shown in Figure A-3. The probe is inserted into the reactor through a full-bore ball valve. Samples of the slurry are used to monitor the coal fines in slurry concentration along the reactor.

The reactor inlet distributor is based on a proprietary design supplied by Hydrocarbon Research, Incorporated (HRI). The distributor was modeled after the one used by HRI on the PDU. Establishing the proper distance between the bubble cup and the reactor bottom is the key for eliminating plugging problems.

The recycle cup design is also a proprietary design supplied by HRI. It is also similar to the one used by HRI in the PDU. The recycle cup is connected to the downcomer with a section of flexible pipe. The recycle cup is spaced inside the glass with Teflon tabs to keep the cup from scratching the reactor. The cup is supported from the top spool piece with four thin steel pipes.

Gamma-Ray Equipment

✓ An elevator is designed so that a gamma-ray source and detector can travel vertically along the reactor to monitor the fluid dynamics of the system. The source and detector are supported on a Unistrut frame. The frame runs in Unistrut guides with Unistrut trolley wheels. The framework is adjustable so two detector/source combinations can be mounted up to 36" apart.

The gamma-ray system attached to the elevator used for scanning the reactor was obtained from Harshaw and K-Ray. The scintillator detection system was manufactured by Harshaw. A NaI scintillation crystal closely coupled to the photomultiplier tube (PMT) and preamplifier is mounted on the elevator. Other components used are: a stabilized amplifier/ single-channel analyzer, a high-voltage source for the PMT, and a rate meter. The scan output is monitored using a Texas Instruments recorder on the panel board. Ten millicuries of Cs-137 from K-Ray is used as the gamma-ray source. Using the elevator system, the reactor can be scanned continuously from the bottom spool piece to the recycle cup.

An Ortec scintillation detection system and a Cs-137 source are used to scan the internal liquid recycle line at the bottom of the reactor. The components used in this system are the same as in the Harshaw system, except that the amplifier/single-channel analyzer does not have the stabilizer option. A vertical length of pipe is scanned. The system is used to monitor the quantity of entrained gas in the recycle liquid. The output is recorded on a Leeds and Northrup recorder.

Compressor and Pumps

A Corken reciprocating compressor, Model D-290K9, with a 5-HP motor was chosen for recycling gas. This compressor has a water-cooled aftercooler and a continuous pressure unloader. It is designed to pump 102 liters per minute at 1 atm inlet and 4 atm outlet pressures. The compressor discharges into a capacity tank to damp out flow rate fluctuations.

March magnetic drive pumps, Model TE-73-MD, were originally chosen for the feed, slurry recycle, and transfer pumps. These centrifugal pumps are magnetically coupled to the motor so there are no problems with liquids or gas leaking through a dynamic seal. Because of the magnetic coupling design of this pump, it will not uncouple from the motor, so pump stoppage is not a problem. At a head of 1.4 atm, the pump will deliver about two gallons per minute. The centrifugal pump performed well provided that the amount of gas entrained in the recycle line was not excessive. It was found that under certain conditions the centrifugal pump became vapor-locked. For this reason, a Tuthill rotary positive displacement pump was ordered to replace the recycle pump. The pump was piped in parallel to both the existing feed and recycle pumps so that it could be used to replace either pump. The pumping capacity was controlled with a Reeves motor drive. Control valves were initially used to vary the liquid flow rates, but they were prone to plugging when the fines were slurried with kerosene.

Other mechanical design details are reported in Reference 34.

Control Components

A Honeywell TDC-2000 microprocessor-based system is used for controlling slurry feed and slurry recycle flows, gas flows, gas/liquid separator level, and reactor pressure. Dual-pen analog recording is provided for hard copy of these process variables. Digital displays of variables, set points, and valve positions are also provided.

Flow Measurement

A Nusonics Model 8000 meter was chosen to monitor slurry feed flow rate. The difference in frequency of sonic beams propagated axially to the flow is correlated with flow rate. A MicroMotion flow meter is used to measure slurry recycle flow rate. The deflection of this U-shaped flow tube due to the Coriolis-type acceleration of the flowing liquid is correlated with the mass flow rate. This flow meter was selected because tests show that entrained gases which are well dispersed do not affect flow measurement and entrained gases are expected in the recycle line. A Nusonics Model 6180 concentration analyzer was chosen for continuous monitoring of the coal concentration in the slurry feed. The gas flow is monitored with a Honeywell Model 41105 differential pressure transmitter with an integral orifice assembly. The pressure drop across the orifice is dependent upon the gas velocity.

Pressure

Pressure drop measurements along reactor sections are made to determine the average ebullated bed three-phase volume fraction. Bourne Model 5020 differential pressure transmitters with high speed of response were chosen for this function. Analog recorders used are: a three-pen Gould/Brush Model 2600 with 100 mm scales for the critical high-speed measurements; and two three-pen Texas Instruments Servo/Recorders with overlapping pens on a 25 cm grid. All three recorders feature multiple chart speeds for maximum flexibility.

A small, continuous bleed of feed gas through each 3.2 mm ID pressure tap impulse line is used to keep the lines free of catalyst or slurry. Should the lines become plugged, however, a high-pressure gas surge can be initiated manually via three-way solenoid valves in the purge lines.

As discussed earlier, eight pressure drop measurements were made with the taps spaced about 61 cm apart. However, the first series of pilot plant tests exhibited excessive data fluctuations due to the vibration of the pressure measurement impulse lines within the reactor. A liquid-filled impulse line without purge flow was tried, but was unsuccessful because of catalyst and slurry plugging.

Thermowell taps in each spool piece are now utilized for the differential pressure measurements. This eliminated the impulse line vibration problem, but reduced the number of differential pressure measurements to four. The pressure drop across each five-foot reactor section is now being monitored.

Temperature

Process and coolant temperatures are displayed digitally via a 10⁰ resolution Doric indicator and recorded on a Honeywell Model 112 24-point recorder.

Computer Hardware Configuration

The general hardware configuration for the computerization of the unit is shown in Figure A-4. Selected equipment items are described below.

ModComp II Computer.--The computer is a general-purpose 16-bit processor with 128K bytes of memory. It employs a dual moving head disc with 5196K bytes of storage capacity. This computer is equipped with asynchronous ports for serial devices such as CRT's and teletypes and with digital I/O for interfacing with the parallel data bus.

Tektronix 4662 Plotter.--This device is used to generate user plots offline for numerous applications. It is driven by the Tektronix Plot 10 software package, which contains the following features: automatic coordinate transformations (linear, logarithmic, and polar); virtual plotting; relative plotting, segmented line drawing; character generation allowing unlimited control of character size and slant; scaling and rotating of relative virtual vectors. Besides being used for graphic representation of test results, the plotter is also used for generating graphic displays similar to those produced by the Intelligence Systems Corporation (ISC) color graphic CRT, thus allowing hard copy of CRT displays. Examples of this feature are reported in Reference 34.

ISC 8001G Color Graphic CRT.--This peripheral is employed for real-time display of unit data. The CRT has eight colors, which may be used as foreground or background colors, with which graphic displays are generated of the pilot plant. When these displays are output to the CRT, real-time data--which are scanned every minute--are inserted at appropriate locations on the screen. An additional useful feature is a software refresh of data in real time. In this way the data are kept current and trends are easy to spot. Also, this refresh feature can be used effectively when lining out the unit.

Computer Software Configuration

The ModComp II computer used for monitoring this unit is running under MAX II/III, which is a vendor-supplied executive system. In addition to this ModComp-supplied operating system, several tasks have been added to the system environment which implement some very useful features.

All of the "system" programs provide a base for application programs specific to this unit. A generalized flow chart of the applications software configuration is shown in Figure A-5. All of the programs are written primarily in Fortran with subroutine calls to Assembly language subroutines. An exception to this is Program TSC, which operates the traveling scintillation counter. This program is written with a mixture of Fortran and Assembly language. A brief description of each major applications program is given in Reference 34.

Reactor design

M80-21

Figure A-4

Computer configuration

Figure A-5

Applications software configuration

APPENDIX B

EXPERIMENTAL DATA AND METHOD OF ANALYSIS

APPENDIX B

EXPERIMENTAL DATA AND METHOD OF ANALYSIS

The experimental tests completed during this work are listed in this appendix. For each run (e.g., 200), there are three sets of tables: the first lists the operating conditions, the bed height, and bed expansion; the second provides the holdup of the various phases in the bed; and the third shows the holdups in the dilute phase.

The volume fraction of catalyst is calculated from measurement of the bed height and mass of catalyst in the reactor:

$$\epsilon_c = \frac{M}{\rho_c A H}$$

where: M = mass of dry catalyst added to the reactor.
 ρ_c = density of a dry catalyst particle.
 A = cross-sectional area of the reactor.
 H = catalyst bed height

The volume fraction of liquid can then be calculated using either gamma-ray scan or pressure drop measurements.

For liquid/catalyst tests, the following equations should be used:

For gamma-ray scan data:

$$\epsilon_l = \frac{\frac{\ln \frac{I_l}{I_m}}{d} - \mu_c \rho_c + \mu_l \rho_l}{\mu_l \rho_l - \mu_c \rho_c + \frac{\omega_f \rho_l (\mu_f \rho_f - \mu_c \rho_c)}{\rho_f (100 - \omega_f)}}$$

For pressure drop measurements:

$$\epsilon_l = \frac{\frac{\Delta P}{H D} - \rho_c}{\rho_l - \rho_c + \frac{\rho_l \omega_f}{100 - \omega_f} \left(1 - \frac{\rho_c}{\rho_f}\right)}$$

When three phases are in the reactor, gas/liquid/catalyst, these equations should be used:

Gamma-ray:

$$\frac{\frac{\ln \frac{I_l}{I_m}}{d} - \mu_c \rho_c \epsilon_c + \mu_l \rho_l}{\mu_l \rho_l + \frac{\mu_f \omega_f \rho_l}{(100 - \omega_f)}}$$

Pressure drop:

$$\epsilon_1 = \frac{\frac{\Delta P}{HD} - \rho_c \epsilon_c}{\rho_1 + \frac{\omega_f \rho_1}{100 - \omega_f}}$$

In either case:

$$\epsilon_f = \frac{\epsilon_1 \rho_1 \omega_f}{\rho_f (100 - \omega_f)}$$

$$\therefore \epsilon_g = 1 - \epsilon_1 - \epsilon_c - \epsilon_f$$

In the dilute phase above the catalyst bed, ϵ_c is set to 0.0 and the same equations are solved.

HD = height between pressure taps
 I_1 = gamma-ray intensity through liquid
 I_m = gamma-ray intensity at test conditions
 ΔP = pressure drop

ϵ_c = volume fraction catalyst
 ϵ_f = volume fraction fines
 ϵ_g = volume fraction gas
 ϵ_l = volume fraction liquid
 ρ_c = catalyst density
 ρ_f = fines density
 ρ_l = liquid density
 μ_c = catalyst mass absorption coefficient
 μ_f = fines mass absorption coefficient
 μ_l = liquid mass absorption coefficient
 ω_f = wt% fines

Based on the tables presented in this appendix, the figures in the main text of this report were prepared. Additional plots are included in this appendix for reference.

Z BED EXPANSION FOR RUN 200

M80-21
-118

CATALYST : NONE *LM=0*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, GPM/Ft2	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
200-00	0.0	0.0	0.	0.
-01	31.3	0.05	0.	0.
-02	31.3	0.10	0.	0.
-03	31.3	0.15	0.	0.
-04	31.3	0.20	0.	0.
-05	31.3	0.25	0.	0.

CALCULATED HOLDUPS, RUN 200: DENSE PHASE

CATALYST : NONE
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (Mn/Sec)
200-00	0.0	0.0	0.0	1.00	0.0	0.0	0.0
-01	31.3	0.05	0.0	0.91	0.97	0.09	11.9
-02	31.3	0.10	0.0	0.84	0.81	0.16	21.9
-03	31.3	0.15	0.0	0.77	0.79	0.23	30.1
-04	31.3	0.20	0.0	0.80	0.81	0.20	44.8
-05	31.3	0.25	0.0	0.78	0.81	0.22	54.9

CALCULATED HOLDUPS, RUN 200--DILUTE PHASE

CATALYST : NONE
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELBP	EOG
200-00	0.0	0.0	1.00	0.0	0.0
-01	31.3	0.05	0.90	1.01	0.10
-02	31.3	0.10	0.79	0.81	0.21
-03	31.3	0.15	0.70	0.81	0.30
-04	31.3	0.20	0.73	0.79	0.27
-05	31.3	0.25	0.71	0.75	0.29

Z BED EXPANSION FOR RUN 201

M80-21
-119

CATALYST : HDS-2A *2/d-3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
201-00	0.0	0.0	64.	0.
-01	0.0	0.0	66.	0.
-02	31.4	0.0	68.	3.
-03	38.0	0.0	69.	5.
-04	44.9	0.0	73.	11.
-05	51.5	0.0	78.	18.
-06	60.6	0.0	87.	32.
-07	67.3	0.0	98.	48.
-08	74.9	0.0	103.	56.
-09	81.7	0.0	108.	64.
-10	89.6	0.0	120.	82.
-11	100.1	0.0	133.	102.
-12	22.4	0.05	73.	11.
-13	22.4	0.03	72.	9.
-14	22.4	0.04	77.	17.
-15	22.4	0.04	74.	16.
-16	22.4	0.08	78.	22.
-17	22.4	0.10	79.	23.
-18	22.4	0.15	79.	23.
-19	22.4	0.20	78.	22.
-20	22.4	0.25	77.	20.
-21	31.4	0.03	73.	14.
-22	31.4	0.04	74.	16.
-23	31.4	0.04	77.	20.
-24	31.4	0.05	79.	23.
-25	31.4	0.08	79.	23.
-26	31.4	0.10	82.	28.
-27	31.4	0.15	87.	36.
-28	31.4	0.20	88.	38.
-29	31.4	0.25	89.	39.
-30	44.8	0.03	82.	28.
-31	44.8	0.04	83.	30.
-32	44.8	0.04	86.	34.
-33	44.8	0.05	87.	36.
-34	44.8	0.08	88.	38.
-35	43.8	0.10	93.	45.
-36	43.8	0.15	104.	63.
-37	43.0	0.20	107.	67.
-38	78.4	0.03	117.	83.
-39	79.3	0.04	118.	69.
-40	78.3	0.04	119.	70.
-41	78.3	0.05	122.	74.
-42	76.7	0.08	127.	81.
-43	71.3	0.10	128.	83.
-44	70.5	0.15	128.	83.
-45	44.9	0.25	97.	39.

CALCULATED HOLDUPS, RUN 201: DENSE PHASE

M80-21
-120

CATALYST : MDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, Bpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (Mm/Sec)
201-00	0.0	0.0	0.49	0.62	1.84	0.0	0.0
-01	0.0	0.0	0.49	0.54	0.58	0.0	0.0
-02	31.4	0.0	0.48	0.54	0.56	0.0	0.0
-03	38.0	0.0	0.47	0.55	0.56	0.0	0.0
-04	44.9	0.0	0.44	0.55	0.65	0.0	0.0
-05	51.5	0.0	0.41	0.60	0.68	0.0	0.0
-06	60.6	0.0	0.37	0.62	0.70	0.0	0.0
-07	67.3	0.0	0.33	0.65	0.70	0.0	0.0
-08	74.9	0.0	0.31	0.67	0.73	0.0	0.0
-09	81.7	0.0	0.30	0.73	0.64	0.0	0.0
-10	89.6	0.0	0.27	0.69	0.77	0.0	0.0
-11	100.1	0.0	0.24	0.73	0.80	0.0	0.0
-12	22.4	0.05	0.44	0.45	0.37	0.11	10.4
-13	22.4	0.03	0.45	0.47	0.41	0.08	4.9
-14	22.4	0.04	0.42	0.48	0.44	0.10	6.9
-15	22.4	0.04	0.42	0.48	0.41	0.10	9.4
-16	22.4	0.08	0.40	0.47	0.56	0.13	17.0
-17	22.4	0.10	0.40	0.46	0.33	0.15	21.9
-18	22.4	0.15	0.40	0.45	0.31	0.16	34.2
-19	22.4	0.20	0.40	0.42	0.32	0.18	44.5
-20	22.4	0.25	0.41	0.37	0.30	0.22	52.6
-21	31.4	0.03	0.43	0.55	0.42	0.02	6.9
-22	31.4	0.04	0.42	0.55	0.41	0.03	9.2
-23	31.4	0.04	0.41	0.54	-0.89	0.05	10.9
-24	31.4	0.05	0.40	0.55	0.42	0.05	12.7
-25	31.4	0.08	0.40	0.49	0.35	0.12	16.1
-26	31.4	0.10	0.38	0.46	0.40	0.16	19.7
-27	31.4	0.15	0.36	0.46	0.39	0.18	30.6
-28	31.4	0.20	0.36	0.44	0.37	0.21	40.4
-29	31.4	0.25	0.35	0.43	0.36	0.22	51.5
-30	44.8	0.03	0.38	0.60	0.50	0.02	7.1
-31	44.8	0.04	0.38	0.59	0.49	0.03	8.6
-32	44.8	0.04	0.36	0.60	0.50	0.04	11.2
-33	44.8	0.05	0.36	0.59	0.50	0.05	12.0
-34	44.8	0.08	0.36	0.56	0.46	0.09	17.0
-35	43.8	0.10	0.34	0.53	0.46	0.13	20.3
-36	43.8	0.15	0.30	0.53	0.43	0.17	29.8
-37	43.0	0.20	0.29	0.48	0.39	0.23	36.7
-38	78.4	0.03	0.27	0.74	0.68	-0.00	8.3
-39	79.3	0.04	0.29	0.66	0.60	0.05	6.7
-40	78.3	0.04	0.29	0.65	0.58	0.07	7.5
-41	78.3	0.05	0.28	0.65	0.58	0.07	9.1
-42	76.7	0.08	0.27	0.61	0.50	0.12	11.4
-43	71.3	0.10	0.27	0.56	0.50	0.17	12.7
-44	70.5	0.15	0.27	0.50	0.45	0.23	18.6
-45	44.9	0.25	0.35	0.38	0.36	0.27	40.3

CALCULATED HOLDUPS, RUN 201--DILUTE PHASE

M80-21
-121

CATALYST : HDS-2A
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EEG
201-00	0.0	0.0	0.99	1.84	0.0
-01	0.0	0.0	0.98	0.96	0.0
-02	31.4	0.0	0.98	0.96	0.0
-03	38.0	0.0	0.97	0.96	0.0
-04	44.9	0.0	0.99	0.99	0.0
-05	51.5	0.0	0.97	0.98	0.0
-06	60.6	0.0	0.08	0.98	0.0
-07	67.3	0.0	0.99	0.98	0.0
-08	74.9	0.0	0.98	0.97	0.0
-09	81.7	0.0	0.94	1.14	0.0
-10	89.6	0.0	0.98	0.89	0.0
-11	100.1	0.0	0.97	0.85	0.0
-12	22.4	0.05	0.96	0.97	0.04
-13	22.4	0.03	0.97	1.00	0.03
-14	22.4	0.04	0.97	0.98	0.03
-15	22.4	0.04	0.95	0.99	0.05
-16	22.4	0.08	0.90	0.92	0.10
-17	22.4	0.10	0.85	0.0	0.15
-18	22.4	0.15	0.82	0.0	0.18
-19	22.4	0.20	0.77	0.0	0.23
-20	22.4	0.25	0.76	0.0	0.24
-21	31.4	0.03	0.98	0.98	0.02
-22	31.4	0.04	0.96	0.91	0.04
-23	31.4	0.04	0.95	0.0	0.05
-24	31.4	0.05	0.94	0.93	0.06
-25	31.4	0.08	0.89	0.89	0.11
-26	31.4	0.10	0.84	0.85	0.16
-27	31.4	0.15	0.77	0.79	0.23
-28	31.4	0.20	0.75	0.78	0.25
-29	31.4	0.25	0.73	0.77	0.27
-30	44.8	0.03	0.97	0.98	0.03
-31	44.8	0.04	0.95	0.97	0.05
-32	44.8	0.04	0.93	0.95	0.07
-33	44.8	0.05	0.91	0.94	0.09
-34	44.8	0.08	0.86	0.88	0.14
-35	43.8	0.10	0.81	0.83	0.19
-36	43.8	0.15	0.74	0.75	0.26
-37	43.0	0.20	0.69	0.71	0.31
-38	78.4	0.03	0.96	0.98	0.04
-39	79.3	0.04	0.94	0.96	0.06
-40	78.3	0.04	0.92	0.94	0.08
-41	78.3	0.05	0.89	0.92	0.11
-42	76.7	0.08	0.83	0.85	0.17
-43	71.3	0.10	0.74	0.79	0.26
-44	70.5	0.15	0.66	0.69	0.34
-45	44.9	0.25	0.72	0.0	0.28

Z BED EXPANSION FOR RUN 203

CATALYST : HDS-2A *l/d = 3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 1.0 VOL %
TEMPERATURE : 68. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
203-01	22.3	0.0	69.	0.
-02	44.7	0.0	79.	14.
-03	67.0	0.0	101.	46.
-04	89.4	0.0	118.	71.
-05	26.1	0.05	77.	12.
-06	26.1	0.20	82.	19.
-07	44.7	0.05	84.	28.
-08	44.7	0.15	97.	45.
-09	67.0	0.05	105.	57.
-10	67.0	0.10	119.	78.
-11	89.4	0.05	124.	85.

CALCULATED HOLDUPS, RUN 203: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 1.0 VOL %
TEMPERATURE : 68. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
203-01	22.3	0.0	0.49	0.50	1.83	0.0	0.0
-02	44.7	0.0	0.43	0.56	0.62	0.0	0.0
-03	67.0	0.0	0.33	0.65	0.69	0.0	0.0
-04	89.4	0.0	0.29	0.70	0.77	0.0	0.0
-05	26.1	0.05	0.44	0.48	0.39	0.08	11.5
-06	26.1	0.20	0.41	0.39	0.34	0.19	42.2
-07	44.7	0.05	0.38	0.55	0.49	0.06	11.3
-08	44.7	0.15	0.34	0.46	0.40	0.20	26.1
-09	67.0	0.05	0.31	0.61	0.55	0.07	9.5
-10	67.0	0.10	0.28	0.57	0.50	0.15	15.6
-11	89.4	0.05	0.26	0.68	0.61	0.05	10.1

CALCULATED HOLDUPS, RUN 203--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 1.0 VOL %
TEMPERATURE : 68. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	E6G
203-01	22.3	0.0	0.93	0.87	0.0
-02	44.7	0.0	0.95	0.86	0.0
-03	67.0	0.0	0.95	0.86	0.0
-04	89.4	0.0	0.95	0.86	0.0
-05	26.1	0.05	0.93	1.07	0.06
-06	26.1	0.20	0.77	0.92	0.22
-07	44.7	0.05	0.92	1.05	0.07
-08	44.7	0.15	0.74	0.87	0.25
-09	67.0	0.05	0.90	1.03	0.09
-10	67.0	0.10	0.76	0.91	0.24
-11	89.4	0.05	0.90	1.04	0.09

X BED EXPANSION FOR RUN 204

CATALYST : HDS-2A *L/d = 3*
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 5.1 VOL %
 TEMPERATURE : 70. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
204-01	30.8	0.0	74.	7.
-02	44.0	0.0	87.	26.
-03	59.3	0.0	101.	46.
-04	65.9	0.0	106.	54.
-05	74.7	0.0	120.	74.
-06	87.8	0.0	135.	96.
-07	30.5	0.05	79.	14.
-08	30.5	0.03	75.	9.
-09	30.5	0.04	76.	10.
-10	30.5	0.04	77.	12.
-11	31.6	0.10	90.	30.
-12	31.6	0.15	89.	29.
-14	44.2	0.04	96.	39.
-15	44.2	0.04	97.	41.
-16	44.2	0.05	107.	55.
-17	44.2	0.10	115.	67.
-18	65.9	0.03	125.	81.
-19	65.9	0.04	127.	84.
-20	65.9	0.04	128.	86.
-21	65.9	0.05	130.	88.
-22	79.9	0.03	140.	103.
-23	79.9	0.04	142.	106.
-24	79.9	0.04	138.	100.
-25	30.5	0.10	86.	24.
-26	30.5	0.03	77.	11.
-27	43.6	0.10	96.	39.

CALCULATED HOLDUPS, RUN 204: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 5.1 VOL %
TEMPERATURE : 70. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (MM/Sec)
204-01	30.8	0.0	0.46	0.51	0.42	0.0	0.0
-02	44.0	0.0	0.39	0.58	0.47	0.0	0.0
-03	59.3	0.0	0.33	0.62	0.51	0.0	0.0
-04	65.9	0.0	0.32	0.63	0.54	0.0	0.0
-05	74.7	0.0	0.28	0.67	0.57	0.0	0.0
-06	87.8	0.0	0.25	0.69	0.60	0.0	0.0
-07	30.5	0.05	0.43	0.49	0.57	0.05	12.4
-08	30.5	0.03	0.45	0.50	0.57	0.02	7.0
-09	30.5	0.04	0.44	0.49	0.57	0.04	8.6
-10	30.5	0.04	0.44	0.47	0.56	0.07	9.9
-11	31.6	0.10	0.38	0.52	0.64	0.08	25.1
-12	31.6	0.15	0.38	0.44	0.36	0.15	32.8
-14	44.2	0.04	0.35	0.57	0.43	0.05	7.8
-15	44.2	0.04	0.35	0.57	0.48	0.05	10.1
-16	44.2	0.05	0.32	0.60	0.51	0.05	12.3
-17	44.2	0.10	0.29	0.52	0.44	0.16	18.3
-18	65.9	0.03	0.27	0.69	0.60	0.00	7.6
-19	65.9	0.04	0.27	0.67	0.58	0.02	8.9
-20	65.9	0.04	0.26	0.66	0.56	0.04	10.3
-21	65.9	0.05	0.26	0.64	0.55	0.06	10.5
-22	79.9	0.03	0.24	0.71	0.63	0.01	7.4
-23	79.9	0.04	0.24	0.69	0.60	0.03	8.1
-24	79.9	0.04	0.24	0.68	0.57	0.04	9.7
-25	30.5	0.10	0.40	0.47	0.36	0.11	22.8
-26	30.5	0.03	0.44	0.51	0.40	0.02	6.9
-27	43.6	0.10	0.35	0.48	0.38	0.14	18.9

CALCULATED HOLDUPS, RUN 204--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 5.1 VOL %
TEMPERATURE : 70. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EBO
204-01	30.8	0.0	0.96	0.82	0.0
-02	44.0	0.0	0.97	0.81	0.0
-03	59.3	0.0	0.96	0.82	0.0
-04	65.9	0.0	0.96	0.82	0.0
-05	74.7	0.0	0.97	0.81	0.0
-06	87.8	0.0	0.96	0.79	0.0
-07	30.5	0.05	0.90	1.05	0.05
-08	30.5	0.03	0.93	1.09	0.02
-09	30.5	0.04	0.93	1.08	0.02
-10	30.5	0.04	0.91	1.07	0.04
-11	31.6	0.10	0.84	1.00	0.12
-12	31.6	0.15	0.78	0.94	0.18
-14	44.2	0.04	0.90	0.97	0.06
-15	44.2	0.04	0.88	1.04	0.07
-16	44.2	0.05	0.87	1.02	0.09
-17	44.2	0.10	0.74	0.89	0.23
-18	65.9	0.03	0.94	1.08	0.01
-19	65.9	0.04	0.89	1.06	0.06
-20	65.9	0.04	0.88	1.04	0.08
-21	65.9	0.05	0.85	1.03	0.10
-22	79.9	0.03	0.68	1.13	0.28
-23	79.9	0.04	0.67	1.10	0.29
-24	79.9	0.04	0.86	1.07	0.09
-25	30.5	0.10	0.82	1.00	0.14
-26	30.5	0.03	0.93	1.10	0.02
-27	43.6	0.10	0.77	0.92	0.19

Z BED EXPANSION FOR RUN 205

M80-21
-127

CATALYST : HDS-2A *ℓ/d=3*
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 10.4 VOL %
 TEMPERATURE : 67. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
205-01	36.3	0.0	81.	17.
-02	42.7	0.0	86.	25.
-03	49.1	0.0	91.	32.
-04	57.6	0.0	107.	55.
-05	72.7	0.0	132.	91.
-06	64.4	0.0	108.	57.
-07	79.1	0.0	140.	103.
-08	85.4	0.0	149.	116.
-09	36.3	0.03	77.	12.
-10	36.3	0.04	82.	19.
-11	36.3	0.04	86.	25.
-12	36.3	0.05	86.	25.
-13	36.3	0.08	85.	22.
-14	36.3	0.10	91.	32.
-15	36.3	0.15	90.	30.
-16	36.3	0.20	86.	25.
-17	36.3	0.25	85.	23.
-18	42.8	0.03	91.	32.
-19	42.8	0.04	91.	32.
-20	42.8	0.04	89.	29.
-21	42.8	0.05	92.	33.
-22	42.8	0.08	92.	33.
-23	42.8	0.10	98.	42.
-24	42.8	0.15	93.	35.
-25	42.8	0.20	92.	33.
-26	42.8	0.25	87.	26.
-27	64.0	0.03	116.	68.
-28	64.0	0.04	120.	74.
-29	64.0	0.04	121.	75.
-30	64.0	0.05	121.	75.
-31	64.0	0.08	126.	83.
-32	85.9	0.04	134.	123.
-33	85.9	0.04	135.	125.
-34	85.9	0.05	136.	126.
-35	94.4	0.0	136.	126.
-36	94.9	0.05	136.	126.

CALCULATED HOLDUPS, RUN 205: DENSE PHASE

M80-21
-128

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 10.4 VOL %
TEMPERATURE : 67. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (MM/Sec)
205-01	36.3	0.0	0.42	0.51	0.57	0.0	0.0
-02	42.7	0.0	0.39	0.55	0.58	0.0	0.0
-03	49.1	0.0	0.37	0.56	0.60	0.0	0.0
-04	57.6	0.0	0.32	0.59	0.64	0.0	0.0
-05	72.7	0.0	0.26	0.62	0.68	0.0	0.0
-06	64.4	0.0	0.31	0.60	0.65	0.0	0.0
-07	79.1	0.0	0.24	0.66	0.71	0.0	0.0
-08	85.4	0.0	0.23	0.67	0.72	0.0	0.0
-09	36.3	0.03	0.44	0.45	0.36	0.06	4.7
-10	36.3	0.04	0.41	0.49	0.40	0.04	8.7
-11	36.3	0.04	0.39	0.51	0.42	0.04	11.2
-12	36.3	0.05	0.39	0.48	0.40	0.07	11.3
-13	36.3	0.08	0.40	0.44	0.35	0.11	16.1
-14	36.3	0.10	0.37	0.45	0.42	0.13	21.2
-15	36.3	0.15	0.38	0.43	0.41	0.14	32.8
-16	36.3	0.20	0.39	0.42	0.39	0.14	45.7
-17	36.3	0.25	0.40	0.41	0.37	0.14	58.7
-18	42.8	0.03	0.37	0.56	0.45	-0.00	8.0
-19	42.8	0.04	0.37	0.53	0.47	0.04	8.6
-20	42.8	0.04	0.38	0.50	0.43	0.06	9.8
-21	42.8	0.05	0.37	0.51	0.44	0.06	11.5
-22	42.8	0.08	0.37	0.48	0.40	0.10	16.4
-23	42.8	0.10	0.34	0.49	0.40	0.11	21.6
-24	42.8	0.15	0.36	0.43	0.37	0.16	30.5
-25	42.8	0.20	0.37	0.46	0.39	0.11	48.3
-26	42.8	0.25	0.39	0.41	0.35	0.15	56.7
-27	64.0	0.03	0.29	0.64	0.57	-0.00	8.2
-28	64.0	0.04	0.28	0.64	0.59	0.00	10.5
-29	64.0	0.04	0.28	0.61	0.56	0.03	10.8
-30	64.0	0.05	0.28	0.61	0.54	0.04	11.8
-31	64.0	0.08	0.27	0.56	0.41	0.11	14.0
-32	85.9	0.04	0.22	0.69	0.64	0.01	9.8
-33	85.9	0.04	0.22	0.68	0.62	0.02	11.6
-34	85.9	0.05	0.22	0.66	0.60	0.05	10.9
-35	94.4	0.0	0.22	0.65	0.70	0.0	0.0
-36	94.9	0.05	0.22	0.66	0.64	0.04	10.9

CALCULATED HOLDUPS, RUN 205--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 10.4 VOL %
TEMPERATURE : 67. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	EL6	ELDP	EGG
205-01	36.3	0.0	0.91	0.72	0.0
-02	42.7	0.0	0.91	0.72	0.0
-03	49.1	0.0	0.91	0.73	0.0
-04	57.6	0.0	0.91	0.72	0.0
-05	72.7	0.0	0.92	0.72	0.0
-06	64.4	0.0	0.91	0.73	0.0
-07	79.1	0.0	0.64	0.68	0.0
-08	85.4	0.0	0.66	0.66	0.0
-09	36.3	0.03	0.88	1.03	0.02
-10	36.3	0.04	0.86	1.02	0.04
-11	36.3	0.04	0.85	1.01	0.05
-12	36.3	0.05	0.83	0.99	0.07
-13	36.3	0.08	0.79	0.96	0.12
-14	36.3	0.10	0.77	0.94	0.14
-15	36.3	0.15	0.75	0.92	0.16
-16	36.3	0.20	0.74	0.91	0.18
-17	36.3	0.25	0.72	0.87	0.20
-18	42.8	0.03	0.89	0.94	0.00
-19	42.8	0.04	0.86	1.02	0.04
-20	42.8	0.04	0.86	1.00	0.04
-21	42.8	0.05	0.85	0.99	0.06
-22	42.8	0.08	0.81	0.95	0.10
-23	42.8	0.10	0.77	0.91	0.14
-24	42.8	0.15	0.76	0.92	0.15
-25	42.8	0.20	0.76	0.92	0.16
-26	42.8	0.25	0.72	0.90	0.20
-27	64.0	0.03	0.89	1.07	0.00
-28	64.0	0.04	0.87	1.02	0.03
-29	64.0	0.04	0.85	1.00	0.05
-30	64.0	0.05	0.83	0.98	0.08
-31	64.0	0.08	0.76	0.0	0.16
-32	85.9	0.04	0.68	1.10	0.24
-33	85.9	0.04	0.65	1.08	0.27
-34	85.9	0.05	0.64	1.06	0.29
-35	94.4	0.0	0.66	0.64	0.0
-36	94.9	0.05	0.64	1.10	0.29

Z BED EXPANSION FOR RUN 206

CATALYST : HDS-2A $L/d = 3$
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 15.5 VOL %
 TEMPERATURE : 85. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
206- 1	30.0	0.0	44.	10.
- 2	36.2	0.0	48.	20.
- 3	42.3	0.0	50.	25.
- 4	48.6	0.0	61.	53.
- 5	56.4	0.0	73.	83.
- 6	63.2	0.0	78.	95.
- 7	71.5	0.0	87.	118.
- 8	77.9	0.0	92.	130.
- 9	87.1	0.0	107.	168.
-10	29.8	0.07	47.	18.
-11	30.0	0.08	52.	30.
-12	30.0	0.08	54.	35.
-13	30.1	0.09	54.	35.
-14	30.0	0.12	54.	35.
-15	30.0	0.14	58.	45.
-16	30.0	0.19	60.	50.
-17	30.0	0.24	62.	55.
-18	30.0	0.29	65.	63.
-19	36.7	0.08	53.	33.

Z BED EXPANSION FOR RUN 206

CATALYST : HDS-2A *L/d = 3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 85. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
206-20	36.4	0.09	55.	38.
-21	36.7	0.12	55.	38.
-22	36.3	0.14	54.	35.
-23	36.2	0.19	55.	38.
-24	36.3	0.24	60.	50.
-25	36.7	0.29	51.	28.
-26	42.9	0.08	65.	63.
-27	43.0	0.03	62.	55.
-28	42.9	0.05	63.	58.
-29	30.0	0.14	48.	20.
-30	36.5	0.14	54.	35.
-31	43.0	0.08	62.	55.
-32	42.9	0.12	63.	58.
-33	43.3	0.15	63.	58.
-34	42.9	0.20	63.	58.
-35	42.9	0.25	64.	60.
-36	63.5	0.03	77.	93.
-37	63.7	0.05	79.	98.
-38	64.0	0.09	82.	105.
-39	30.0	0.07	53.	33.
-40	30.0	0.14	49.	23.
-41	36.7	0.14	55.	38.
-42	36.7	0.08	52.	30.
-43	63.7	0.15	78.	95.
-44	63.7	0.20	74.	85.
-45	63.7	0.34	69.	73.
-46	84.5	0.04	106.	165.
-47	84.3	0.04	112.	180.
-48	84.4	0.05	110.	175.
-49	36.1	0.04	53.	33.
-50	36.8	0.05	54.	35.
-51	36.7	0.04	54.	35.
-52	30.5	0.04	48.	20.
-53	30.5	0.15	50.	25.
-54	30.5	0.17	49.	23.
-55	30.5	0.20	47.	18.
-56	42.9	0.0	60.	50.

CALCULATED HOLDUPS, RUN 206: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 85. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
206- 1	30.0	0.0	0.44	0.43	1.73	0.0	0.0
- 2	36.2	0.0	0.41	0.47	1.73	0.0	0.0
- 3	42.3	0.0	0.39	0.49	1.73	0.0	0.0
- 4	48.6	0.0	0.32	0.56	1.73	0.0	0.0
- 5	56.4	0.0	0.27	0.60	1.73	0.0	0.0
- 6	63.2	0.0	0.25	0.61	1.73	0.0	0.0
- 7	71.5	0.0	0.23	0.64	1.73	0.0	0.0
- 8	77.9	0.0	0.21	0.64	1.73	0.0	0.0
- 9	87.1	0.0	0.18	0.66	1.73	0.0	0.0
-10	29.8	0.07	0.42	0.48	-0.77	0.01	19.6
-11	30.0	0.08	0.38	0.51	-0.71	0.02	21.9
-12	30.0	0.08	0.36	0.51	-0.68	0.03	24.0
-13	30.1	0.09	0.36	0.51	-0.68	0.04	25.6
-14	30.0	0.12	0.36	0.48	-0.68	0.06	31.5
-15	30.0	0.14	0.34	0.52	-0.64	0.05	39.3
-16	30.0	0.19	0.33	0.52	-0.62	0.06	52.9
-17	30.0	0.24	0.32	0.53	-0.60	0.06	67.7
-18	30.0	0.29	0.30	0.53	-0.57	0.07	80.4
-19	36.7	0.08	0.37	0.51	-0.69	0.03	21.3
-20	36.4	0.09	0.36	0.50	-0.67	0.05	24.5
-21	36.7	0.12	0.36	0.49	-0.67	0.07	31.0
-22	36.3	0.14	0.36	0.47	-0.68	0.08	36.0
-23	36.2	0.19	0.36	0.47	-0.67	0.09	49.6
-24	36.3	0.24	0.33	0.50	-0.62	0.08	64.4
-25	36.7	0.29	0.38	0.42	-0.72	0.12	73.2
-26	42.9	0.08	0.30	0.58	-0.58	0.01	22.5
-27	43.0	0.03	0.32	0.57	0.49	0.00	9.8
-28	42.9	0.05	0.31	0.56	0.51	0.02	14.0
-29	30.0	0.14	0.41	0.43	0.27	0.09	35.8
-30	36.5	0.14	0.36	0.47	0.37	0.08	36.3

CALCULATED HOLDUPS, RUN 206: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 85. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (MM/Sec)
206 -31	43.0	0.08	0.32	0.52	0.44	0.07	18.8
-32	42.9	0.12	0.31	0.53	0.44	0.07	32.3
-33	43.3	0.15	0.31	0.52	1.98	0.07	39.0
-34	42.9	0.20	0.31	0.51	0.44	0.09	51.7
-35	42.9	0.25	0.31	0.51	1.84	0.09	64.7
-36	63.5	0.03	0.25	0.63	2.11	0.01	10.1
-37	63.7	0.05	0.25	0.60	2.07	0.04	12.3
-38	64.0	0.09	0.24	0.58	0.50	0.07	20.8
-39	30.0	0.07	0.37	0.49	0.15	0.06	18.8
-40	30.0	0.14	0.40	0.43	0.31	0.09	35.4
-41	36.7	0.14	0.36	0.48	0.37	0.07	36.6
-42	36.7	0.08	0.38	0.46	0.37	0.08	19.5
-43	63.7	0.15	0.25	0.53	0.44	0.12	32.7
-44	63.7	0.20	0.26	0.50	0.42	0.14	43.9
-45	63.7	0.34	0.28	0.47	0.74	0.16	75.4
-46	84.5	0.04	0.18	0.69	0.85	0.00	10.6
-47	84.3	0.04	0.17	0.69	0.82	0.01	12.3
-48	84.4	0.05	0.18	0.66	0.80	0.04	11.8
-49	36.1	0.04	0.37	0.49	0.59	0.05	10.5
-50	36.8	0.05	0.36	0.49	0.59	0.06	12.1
-51	36.7	0.04	0.36	0.51	0.39	0.04	8.7
-52	30.5	0.04	0.41	0.45	0.31	0.06	10.0
-53	30.5	0.15	0.39	0.43	0.31	0.10	37.7
-54	30.5	0.17	0.40	0.41	0.32	0.11	42.9
-55	30.5	0.20	0.42	0.37	0.27	0.14	47.0
-56	42.9	0.0	0.33	0.54	0.61	0.0	0.0

CALCULATED HOLDUPS, RUN 206--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 85. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, ft/Sec	ELG	ELDP	EGG
206- 1	30.0	0.0	0.85	0.50	0.0
- 2	36.2	0.0	0.86	0.50	0.0
- 3	42.3	0.0	0.86	0.50	0.0
- 4	48.6	0.0	0.86	0.50	0.0
- 5	56.4	0.0	0.86	0.50	0.0
- 6	63.2	0.0	0.86	0.50	0.0
- 7	71.5	0.0	0.86	0.47	0.0
- 8	77.9	0.0	0.85	0.47	0.0
- 9	87.1	0.0	0.86	0.47	0.0
-10	29.8	0.07	0.82	1.19	0.03
-11	30.0	0.08	0.81	1.19	0.04
-12	30.0	0.08	0.81	1.19	0.04
-13	30.1	0.09	0.80	1.18	0.06
-14	30.0	0.12	0.78	1.16	0.08
-15	30.0	0.14	0.77	1.14	0.09
-16	30.0	0.19	0.75	1.25	0.11
-17	30.0	0.24	0.73	1.22	0.13
-18	30.0	0.29	0.71	1.21	0.16
-19	36.7	0.08	0.81	1.30	0.04
-20	36.4	0.09	0.80	1.29	0.06
-21	36.7	0.12	0.78	1.27	0.08
-22	36.3	0.14	0.77	1.26	0.09
-23	36.2	0.19	0.74	1.34	0.12

CALCULATED HOLDUPS, RUN 206--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 85. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	E6G
206-24	36.3	0.24	0.73	1.33	0.14
-25	36.7	0.29	0.71	1.31	0.16
-26	42.9	0.08	0.81	1.41	0.04
-27	43.0	0.03	0.84	1.04	0.00
-28	42.9	0.05	0.81	1.02	0.04
-29	30.0	0.14	0.77	0.98	0.09
-30	36.5	0.14	0.77	0.97	0.09
-31	43.0	0.08	0.79	0.98	0.06
-32	42.9	0.12	0.78	0.97	0.08
-33	43.3	0.15	0.75	0.97	0.11
-34	42.9	0.20	0.74	0.96	0.13
-35	42.9	0.25	0.72	0.95	0.15
-36	63.5	0.03	0.83	1.04	0.02
-37	63.7	0.05	0.80	1.00	0.05
-38	64.0	0.09	0.75	0.95	0.11
-39	30.0	0.07	0.81	1.01	0.04
-40	30.0	0.14	0.77	0.97	0.09
-41	36.7	0.14	0.77	0.97	0.09
-42	36.7	0.08	0.80	1.00	0.06
-43	63.7	0.15	0.73	0.93	0.14
-44	63.7	0.20	0.71	0.93	0.16
-45	63.7	0.34	0.68	0.90	0.20
-46	84.5	0.04	0.83	1.04	0.01
-47	84.3	0.04	0.81	1.01	0.04
-48	84.4	0.05	0.79	1.00	0.06
-49	36.1	0.04	0.83	1.00	0.01
-50	36.8	0.05	0.82	1.03	0.03
-51	36.7	0.04	0.85	0.93	-0.00
-52	30.5	0.04	0.84	0.91	0.00
-53	30.5	0.15	0.76	0.97	0.10
-54	30.5	0.17	0.74	0.96	0.13
-55	30.5	0.20	0.71	0.94	0.16
-56	42.9	0.0	0.89	0.61	0.0

% BED EXPANSION FOR RUN 207

CATALYST : HDS-2A *l/d = 3*
GAS : FREON-12
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 81. DEG F

Run No.	Liquid Flow Rate, GPM/Ft2	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
207- 1	37.0	0.03	48.	37.
- 2	38.2	0.06	52.	49.
- 3	38.1	0.07	53.	52.
- 4	43.3	0.03	57.	63.
- 5	43.4	0.05	55.	57.
- 6	43.2	0.07	58.	66.
- 7	62.4	0.03	69.	97.
- 8	63.0	0.05	68.	95.
- 9	62.9	0.07	67.	92.
-10	83.3	0.03	95.	172.
-11	82.9	0.05	98.	180.
-12	83.3	0.07	104.	198.
-13	82.6	0.09	105.	200.
-14	63.7	0.09	77.	120.
-15	42.5	0.09	61.	75.
-16	36.0	0.09	54.	54.
-17	35.8	0.05	49.	40.
-18	62.1	0.07	71.	103.
-19	36.2	0.17	61.	60.
-20	36.7	0.20	63.	65.
-21	36.3	0.25	65.	70.
-22	62.9	0.17	76.	99.
-23	62.9	0.20	71.	86.
-24	63.0	0.25	69.	81.
-25	70.5	0.17	82.	115.
-26	42.6	0.17	65.	70.
-27	42.4	0.25	67.	75.
-28	42.3	0.30	70.	89.
-29	37.2	0.30	68.	82.
-30	62.6	0.10	83.	124.
-31	83.4	0.05	109.	195.
-32	83.5	0.06	110.	197.
-33	63.2	0.05	78.	111.
-34	43.1	0.06	62.	68.

CALCULATED HOLDUPS, RUN 207: DENSE PHASE

CATALYST : HDS-2A
GAS : FREON-12
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 81. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGR	ELDPB	ELGR	Vcd (Mm/Sec)
207- 1	37.0	0.03	0.36	0.53	0.43	0.02	8.5
- 2	38.2	0.05	0.33	0.54	0.47	0.03	13.7
- 3	38.1	0.07	0.32	0.54	0.48	0.04	18.9
- 4	43.3	0.03	0.30	0.62	0.58	-0.03	11.1
- 5	43.4	0.05	0.31	0.60	0.55	-0.02	16.3
- 6	43.2	0.07	0.30	0.59	0.58	0.01	20.7
- 7	62.4	0.03	0.25	0.70	0.56	-0.08	14.7
- 8	63.0	0.05	0.25	0.69	0.55	-0.07	20.2
- 9	62.9	0.07	0.26	0.67	0.64	-0.04	24.7
-10	83.3	0.03	0.18	0.77	0.73	-0.09	16.7
-11	82.9	0.05	0.17	0.78	0.77	-0.10	23.1
-12	83.3	0.07	0.16	0.75	0.76	-0.06	26.3
-13	82.6	0.09	0.16	0.72	0.71	-0.01	28.8
-14	63.7	0.09	0.22	0.67	0.67	-0.01	28.6
-15	42.5	0.09	0.28	0.62	0.60	-0.02	28.5
-16	36.0	0.09	0.32	0.58	0.55	-0.00	27.7
-17	35.8	0.05	0.35	0.59	0.50	-0.05	17.8
-18	62.1	0.07	0.24	0.67	0.62	-0.03	23.6
-19	36.2	0.17	0.31	0.53	0.48	0.07	45.9
-20	36.7	0.20	0.30	0.53	0.49	0.08	53.2
-21	36.3	0.25	0.29	0.53	0.54	0.08	67.0
-22	62.9	0.17	0.25	0.55	0.54	0.11	39.8
-23	62.9	0.20	0.26	0.51	0.51	0.13	44.5
-24	63.0	0.25	0.27	0.50	0.50	0.14	56.6
-25	70.5	0.17	0.23	0.54	0.53	0.13	36.2
-26	42.6	0.17	0.29	0.55	0.54	0.06	46.0
-27	42.4	0.25	0.28	0.55	0.55	0.07	67.8
-28	42.3	0.30	0.26	0.54	0.54	0.10	78.6
-29	37.2	0.30	0.27	0.55	0.53	0.09	80.6
-30	62.6	0.10	0.22	0.57	0.53	0.11	21.0
-31	83.4	0.05	0.17	0.67	0.63	0.04	11.6
-32	83.5	0.06	0.16	0.65	0.61	0.06	14.3
-33	63.2	0.05	0.23	0.63	0.60	0.02	13.8
-34	48.1	0.06	0.29	0.57	0.53	0.03	18.1

CALCULATED HOLDUPS, RUN 207--DILUTE PHASE

CATALYST : HDS-2A
GAS : FREON-12
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 81. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	E8G
207- 1	37.0	0.03	0.83	0.83	0.01
- 2	38.2	0.05	0.81	0.81	0.04
- 3	38.1	0.07	0.78	0.78	0.07
- 4	43.3	0.03	0.86	0.87	-0.02
- 5	43.4	0.05	0.84	0.85	0.01
- 6	43.2	0.07	0.78	0.80	0.07
- 7	62.4	0.03	0.90	0.86	-0.07
- 8	63.0	0.05	0.88	0.86	-0.04
- 9	62.9	0.07	0.86	0.0	-0.01
-10	83.3	0.03	0.91	0.0	-0.08
-11	82.9	0.05	0.89	0.0	-0.06
-12	83.3	0.07	0.83	0.0	0.02
-13	82.6	0.09	0.75	0.0	0.11
-14	63.7	0.09	0.79	0.0	0.07
-15	42.5	0.09	0.80	0.0	0.05
-16	36.0	0.09	0.81	0.01	0.04
-17	35.8	0.05	0.86	0.0	-0.02
-18	62.1	0.07	0.80	0.0	0.05
-19	36.2	0.17	0.71	0.71	0.17
-20	36.7	0.20	0.69	0.70	0.18
-21	36.3	0.25	0.67	0.69	0.20
-22	62.9	0.17	0.69	0.71	0.18
-23	62.9	0.20	0.69	0.71	0.18
-24	63.0	0.25	0.68	0.69	0.20
-25	70.5	0.17	0.69	0.71	0.18
-26	42.6	0.17	0.73	0.74	0.13
-27	42.4	0.25	0.69	0.71	0.18
-28	42.3	0.30	0.65	0.67	0.23
-29	37.2	0.30	0.66	0.68	0.22
-30	62.6	0.10	0.70	0.71	0.17
-31	83.4	0.05	0.73	0.73	0.13
-32	83.5	0.06	0.71	0.70	0.16
-33	63.2	0.05	0.78	0.78	0.08
-34	43.1	0.06	0.78	0.78	0.07

Z BED EXPANSION FOR RUN 208

CATALYST : NDS-2A *l/d=3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.2 VOL %
TEMPERATURE : 148. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
208- 1	36.2	0.0	44.	26.
- 2	42.8	0.0	46.	31.
- 3	58.8	0.0	59.	59.
- 4	65.0	0.0	61.	65.
- 5	79.4	0.0	75.	103.
- 6	86.2	0.0	78.	111.
- 8	37.4	0.05	47.	27.
- 9	37.6	0.10	49.	32.
-10	37.0	0.15	52.	44.
-11	37.3	0.20	50.	39.
-12	36.5	0.25	55.	53.
-13	42.9	0.05	54.	50.
-14	42.8	0.10	60.	67.
-15	42.6	0.15	62.	72.
-16	43.5	0.20	60.	67.
-17	43.5	0.25	57.	58.
-18	63.7	0.05	63.	75.
-19	64.0	0.10	70.	94.
-20	64.1	0.15	76.	111.
-21	63.6	0.20	70.	94.
-22	63.9	0.25	68.	89.
-23	85.1	0.05	85.	136.
-24	86.3	0.10	82.	128.
-25	79.1	0.15	93.	158.
-26	85.8	0.0	78.	117.
-27	37.8	0.0	44.	22.
-28	63.7	0.0	63.	75.
-29	64.1	0.20	69.	92.
-30	43.3	0.10	58.	61.
-31	43.1	0.15	62.	72.

CALCULATED HOLDUPS, RUN 208: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.2 VOL %
TEMPERATURE : 148. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELCB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
208- 1	36.2	0.0	0.39	0.48	1.64	0.0	0.0
- 2	42.8	0.0	0.37	0.50	1.64	0.0	0.0
- 3	58.8	0.0	0.31	0.58	0.60	0.0	0.0
- 4	65.0	0.0	0.30	0.59	0.61	0.0	0.0
- 5	79.4	0.0	0.24	0.62	0.61	0.0	0.0
- 6	86.2	0.0	0.23	0.63	0.63	0.0	0.0
- 8	37.4	0.05	0.39	0.47	-1.49	0.06	11.9
- 9	37.6	0.10	0.37	0.45	-1.46	0.10	23.3
-10	37.0	0.15	0.34	0.48	-0.58	0.10	37.3
-11	37.3	0.20	0.35	0.46	-0.60	0.11	50.0
-12	36.5	0.25	0.32	0.48	-0.55	0.11	63.1
-13	42.9	0.05	0.33	0.55	-0.56	0.02	14.0
-14	42.8	0.10	0.29	0.53	-0.50	0.08	24.3
-15	42.6	0.15	0.28	0.53	-0.48	0.09	37.9
-16	43.5	0.20	0.29	0.51	-0.50	0.10	50.5
-17	43.5	0.25	0.31	0.49	-0.53	0.11	62.9
-18	63.7	0.05	0.28	0.56	-0.48	0.06	10.6
-19	64.0	0.10	0.25	0.54	-0.43	0.11	20.5
-20	64.1	0.15	0.23	0.52	-0.39	0.16	29.2
-21	63.6	0.20	0.25	0.50	-0.43	0.16	41.4
-22	63.9	0.25	0.26	0.50	-0.44	0.16	54.5
-23	85.1	0.05	0.21	0.67	-0.35	0.01	14.6
-24	86.3	0.10	0.22	0.59	-0.36	0.09	20.8
-25	79.1	0.15	0.19	0.58	-0.32	0.13	30.7
-26	85.8	0.0	0.23	0.62	1.63	0.0	0.0
-27	37.8	0.0	0.40	0.49	1.63	0.0	0.0
-28	63.7	0.0	0.28	0.59	1.63	0.0	0.0
-29	64.1	0.20	0.26	0.51	-0.43	0.14	43.9
-30	43.3	0.10	0.30	0.52	0.47	0.08	24.2
-31	43.1	0.15	0.28	0.53	0.49	0.09	37.4

CALCULATED HOLDUPS, RUN 208--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.2 VOL %
TEMPERATURE : 148. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
208- 1	36.2	0.0	0.88	1.60	0.0
- 2	42.8	0.0	0.86	1.60	0.0
- 3	58.8	0.0	0.87	0.80	0.0
- 4	65.0	0.0	0.87	0.80	0.0
- 5	79.4	0.0	0.87	0.80	0.0
- 6	86.2	0.0	0.87	0.80	0.0
- 8	37.4	0.05	0.83	0.85	0.02
- 9	37.6	0.10	0.76	0.79	0.11
-10	37.0	0.15	0.74	0.77	0.12
-11	37.3	0.20	0.72	0.75	0.15
-12	36.5	0.25	0.70	0.74	0.18
-13	42.9	0.05	0.82	0.85	0.03
-14	42.8	0.10	0.76	0.78	0.11
-15	42.6	0.15	0.75	0.77	0.12
-16	43.5	0.20	0.72	0.76	0.15
-17	43.5	0.25	0.70	0.74	0.18
-18	63.7	0.05	0.81	0.85	0.04
-19	64.0	0.10	0.74	0.78	0.12
-20	64.1	0.15	0.69	0.73	0.19
-21	63.6	0.20	0.69	0.74	0.18
-22	63.9	0.25	0.69	0.73	0.19
-23	85.1	0.05	0.82	0.85	0.03
-24	86.3	0.10	0.77	0.78	0.09
-25	79.1	0.15	0.70	0.72	0.17
-26	85.8	0.0	0.88	0.80	0.0
-27	37.8	0.0	0.87	0.80	0.0
-28	63.7	0.0	0.87	0.80	0.0
-29	64.1	0.20	0.72	0.74	0.15
-30	43.3	0.10	0.77	0.79	0.10
-31	43.1	0.15	0.74	0.78	0.12

Z BED EXPANSION FOR RUN 209

CATALYST : HDS-2A *2/d=3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 97. DEG F

Run No.	Liquid Flow Rate, BPM/Ft2	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
209- 1	36.0	0.0	48.	33.
- 2	42.5	0.0	52.	44.
- 3	62.8	0.0	72.	100.
- 4	83.7	0.0	100.	178.
- 5	36.3	0.05	52.	44.
- 6	36.5	0.10	58.	61.
- 7	36.1	0.20	62.	72.
- 8	36.3	0.15	52.	41.
- 9	36.6	0.15	56.	51.
-10	35.8	0.24	63.	70.
-11	42.2	0.05	62.	68.
-12	42.6	0.10	64.	73.
-13	42.4	0.15	65.	76.
-14	42.9	0.20	64.	73.
-15	42.1	0.25	65.	76.
-16	63.3	0.05	78.	111.
-17	63.2	0.10	84.	127.
-18	62.4	0.15	90.	143.
-19	62.6	0.20	78.	111.
-20	63.1	0.26	78.	111.
-21	82.9	0.05	99.	168.
-22	83.5	0.10	107.	189.
-23	83.8	0.15	105.	184.
-24	83.1	0.0	94.	154.
-25	63.3	0.0	70.	89.
-26	63.0	0.15	79.	114.
-27	62.5	0.15	85.	130.
-28	37.0	0.15	61.	65.
-29	83.1	0.05	106.	187.
-30	42.0	0.10	64.	73.

CALCULATED HOLDUPS, RUN 209: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 97. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Nm/Sec)
209- 1	36.0	0.0	0.37	0.56	0.49	0.0	0.0
- 2	42.5	0.0	0.34	0.58	0.46	0.0	0.0
- 3	62.8	0.0	0.24	0.68	0.50	0.0	0.0
- 4	83.7	0.0	0.18	0.72	0.55	0.0	0.0
- 5	36.3	0.05	0.34	0.47	0.61	0.10	9.5
- 6	36.5	0.10	0.30	0.46	0.62	0.15	20.2
- 7	36.1	0.20	0.28	0.49	0.56	0.14	47.3
- 8	36.3	0.15	0.35	0.41	0.64	0.16	31.7
- 9	36.6	0.15	0.32	0.46	0.68	0.13	34.5
-10	35.8	0.24	0.29	0.48	0.57	0.14	56.1
-11	42.2	0.05	0.29	0.52	0.59	0.10	9.7
-12	42.6	0.10	0.28	0.47	0.55	0.16	18.4
-13	42.4	0.15	0.28	0.47	0.48	0.17	30.7
-14	42.9	0.20	0.28	0.48	0.51	0.15	44.9
-15	42.1	0.25	0.28	0.47	0.51	0.17	55.3
-16	63.3	0.05	0.23	0.57	0.59	0.09	8.5
-17	63.2	0.10	0.22	0.52	0.54	0.17	15.6
-18	62.4	0.15	0.20	0.50	0.51	0.21	24.2
-19	62.6	0.20	0.23	0.47	0.49	0.22	34.8
-20	63.1	0.26	0.23	0.47	0.51	0.22	48.7
-21	82.9	0.05	0.18	0.62	0.61	0.08	8.3
-22	83.5	0.10	0.17	0.55	0.54	0.18	12.4
-23	83.8	0.15	0.17	0.51	0.50	0.22	19.1
-24	83.1	0.0	0.19	0.72	0.71	0.0	0.0
-25	63.3	0.0	0.26	0.67	0.67	0.0	0.0
-26	63.0	0.15	0.23	0.48	0.46	0.21	23.7
-27	62.5	0.15	0.21	0.48	0.48	0.21	23.6
-28	37.0	0.15	0.30	0.48	0.46	0.13	34.4
-29	83.1	0.05	0.17	0.63	0.62	0.08	8.5
-30	42.0	0.10	0.28	0.47	0.47	0.16	19.0

CALCULATED HOLDUPS, RUN 209--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 97. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
209- 1	36.0	0.0	0.83	0.80	0.0
- 2	42.5	0.0	0.83	0.81	0.0
- 3	62.8	0.0	0.83	0.80	0.0
- 4	83.7	0.0	0.83	0.80	0.0
- 5	36.3	0.05	0.76	0.82	0.11
- 6	36.5	0.10	0.70	0.76	0.17
- 7	36.1	0.20	0.66	0.73	0.22
- 8	36.3	0.15	0.69	0.76	0.18
- 9	36.6	0.15	0.69	0.76	0.18
-10	35.8	0.24	0.65	0.73	0.24
-11	42.2	0.05	0.75	0.82	0.11
-12	42.6	0.10	0.69	0.75	0.18
-13	42.4	0.15	0.67	0.74	0.20
-14	42.9	0.20	0.65	0.72	0.23
-15	42.1	0.25	0.64	0.72	0.24
-16	63.3	0.05	0.74	0.80	0.12
-17	63.2	0.10	0.67	0.72	0.20
-18	62.4	0.15	0.64	0.68	0.24
-19	62.6	0.20	0.65	0.70	0.23
-20	63.1	0.26	0.64	0.69	0.25
-21	82.9	0.05	0.76	0.81	0.10
-22	83.5	0.10	0.66	0.71	0.22
-23	83.8	0.15	0.60	0.65	0.29
-24	83.1	0.0	0.82	0.81	0.0
-25	63.3	0.0	0.82	0.81	0.0
-26	63.0	0.15	0.64	0.70	0.24
-27	62.5	0.15	0.64	0.70	0.24
-28	37.0	0.15	0.68	0.75	0.20
-29	83.1	0.05	0.75	0.80	0.12
-30	42.0	0.10	0.69	0.74	0.18

Z BED EXPANSION FOR RUN 210

CATALYST : WDS-2A *l/d=3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 11.9 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
210-01	36.4	0.0	63.	26.
-02	36.0	0.05	70.	40.
-03	37.7	0.10	74.	48.
-04	37.9	0.15	72.	44.
-05	37.7	0.20	70.	40.
-06	38.2	0.24	70.	40.
-07	43.2	0.0	69.	38.
-08	43.6	0.05	74.	48.
-09	43.5	0.10	77.	54.
-10	43.9	0.15	75.	50.
-11	42.5	0.20	72.	44.
-12	44.7	0.24	72.	44.
-13	63.8	0.0	92.	84.
-14	63.8	0.05	98.	96.
-15	63.9	0.10	105.	110.
-16	63.2	0.15	100.	100.
-17	63.9	0.20	99.	98.
-18	64.6	0.24	95.	90.
-19	84.0	0.0	128.	156.
-20	84.5	0.05	129.	158.
-21	84.9	0.10	135.	170.
-22	64.0	0.20	95.	90.
-23	44.9	0.10	78.	56.
-24	44.4	0.15	76.	52.
-25	44.5	0.10	77.	54.
-26	43.6	0.15	75.	50.
-27	44.7	0.10	77.	54.
-28	44.9	0.15	76.	52.

CALCULATED HOLDUPS, RUN 210: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 11.9 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	ELGB	Vcd (Mm/Sec)
210-01	36.4	0.0	0.39	0.56	0.59	0.0	0.0
-02	36.0	0.05	0.35	0.51	0.49	0.07	11.6
-03	37.7	0.10	0.33	0.49	0.48	0.11	22.5
-04	37.9	0.15	0.34	0.48	0.48	0.11	36.1
-05	37.7	0.20	0.35	0.50	0.51	0.09	52.2
-06	38.2	0.24	0.35	0.50	0.53	0.08	62.4
-07	43.2	0.0	0.36	0.52	0.54	0.0	0.0
-08	43.6	0.05	0.33	0.53	0.57	0.07	11.3
-09	43.5	0.10	0.32	0.50	0.53	0.12	21.3
-10	43.9	0.15	0.33	0.48	0.51	0.13	33.5
-11	42.5	0.20	0.34	0.46	0.51	0.13	46.5
-12	44.7	0.24	0.34	0.46	0.34	0.13	55.3
-13	63.8	0.0	0.27	0.66	0.64	0.0	0.0
-14	63.8	0.05	0.25	0.61	0.62	0.06	10.8
-15	63.9	0.10	0.23	0.55	0.54	0.15	17.2
-16	63.2	0.15	0.24	0.51	0.52	0.18	26.4
-17	63.9	0.20	0.25	0.53	0.54	0.16	41.9
-18	64.6	0.24	0.26	0.52	0.54	0.16	50.5
-19	84.0	0.0	0.19	0.71	0.70	0.0	0.0
-20	84.5	0.05	0.19	0.65	0.67	0.07	9.3
-21	84.9	0.10	0.18	0.59	0.59	0.15	14.8
-22	64.0	0.20	0.26	0.51	0.51	0.17	40.4
-23	44.9	0.10	0.31	0.51	0.07	0.10	22.4
-24	44.4	0.15	0.32	0.49	0.51	0.12	34.8
-25	44.5	0.10	0.32	0.51	0.55	0.11	22.1
-26	43.6	0.15	0.33	0.49	0.52	0.12	34.9
-27	44.7	0.10	0.32	0.51	0.52	0.11	22.3
-28	44.9	0.15	0.32	0.49	0.53	0.12	34.8

CALCULATED HOLDUPS, RUN 210--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 11.9 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
210-01	36.4	0.0	0.88	0.86	0.0
-02	36.0	0.05	0.81	0.87	0.08
-03	37.7	0.10	0.76	0.82	0.14
-04	37.9	0.15	0.74	0.80	0.16
-05	37.7	0.20	0.71	0.78	0.20
-06	38.2	0.24	0.69	0.76	0.21
-07	43.2	0.0	0.88	0.86	0.0
-08	43.6	0.05	0.81	0.86	0.09
-09	43.5	0.10	0.75	0.81	0.15
-10	43.9	0.15	0.73	0.79	0.18
-11	42.5	0.20	0.70	0.77	0.21
-12	44.7	0.24	0.69	0.70	0.22
-13	63.8	0.0	0.89	0.86	0.0
-14	63.8	0.05	0.79	0.84	0.10
-15	63.9	0.10	0.67	0.73	0.24
-16	63.2	0.15	0.69	0.74	0.22
-17	63.9	0.20	0.68	0.74	0.23
-18	64.6	0.24	0.67	0.74	0.24
-19	84.0	0.0	0.89	0.86	0.0
-20	84.5	0.05	0.79	0.85	0.11
-21	84.9	0.10	0.67	0.75	0.24
-22	64.0	0.20	0.68	0.75	0.23
-23	44.9	0.10	0.76	0.87	0.14
-24	44.4	0.15	0.74	0.79	0.16
-25	44.5	0.10	0.76	0.81	0.14
-26	43.6	0.15	0.74	0.79	0.16
-27	44.7	0.10	0.77	0.80	0.13
-28	44.9	0.15	0.74	0.79	0.16

X BED EXPANSION FOR RUN 211

CATALYST : MDS-2A *L/d: 3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 17.8 VOL %
TEMPERATURE : 80. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
211-01	36.1	0.0	51.	38.
-02	35.8	0.05	54.	46.
-03	36.7	0.10	57.	54.
-04	37.2	0.15	60.	62.
-05	36.0	0.20	62.	68.
-06	36.5	0.24	57.	54.
-07	42.3	0.0	57.	54.
-08	41.1	0.05	64.	73.
-09	42.5	0.10	65.	76.
-10	42.8	0.15	66.	78.
-11	43.0	0.20	64.	73.
-12	42.2	0.24	63.	70.
-13	62.0	0.0	79.	114.
-14	42.8	0.05	83.	124.
-15	63.9	0.10	89.	141.
-16	61.8	0.15	80.	116.
-17	62.9	0.20	80.	116.
-18	62.7	0.25	78.	111.

CALCULATED HOLDUPS, RUN 211: DENSE PHASE

CATALYST : HDS-2A
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 17.8 VOL %
 TEMPERATURE : 80. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
211-01	36.1	0.0	0.36	0.57	0.53	0.0	0.0
-02	35.8	0.05	0.34	0.47	0.55	0.10	10.0
-03	36.7	0.10	0.32	0.46	0.52	0.12	22.3
-04	37.2	0.15	0.30	0.49	0.53	0.10	37.0
-05	36.0	0.20	0.29	0.49	0.54	0.11	49.8
-06	36.5	0.24	0.32	0.44	0.50	0.15	55.5
-07	42.3	0.0	0.32	0.61	0.54	0.0	0.0
-08	41.1	0.05	0.28	0.52	0.59	0.08	10.6
-09	42.5	0.10	0.28	0.48	0.55	0.13	20.9
-10	42.8	0.15	0.27	0.50	0.56	0.12	35.1
-11	43.0	0.20	0.28	0.48	0.55	0.14	46.5
-12	42.2	0.24	0.29	0.47	0.54	0.14	55.1
-13	62.0	0.0	0.23	0.66	0.61	0.0	0.0
-14	62.8	0.05	0.22	0.56	0.62	0.10	8.1
-15	63.9	0.10	0.20	0.52	0.57	0.17	15.6
-16	61.8	0.15	0.23	0.49	0.54	0.18	27.2
-17	62.9	0.20	0.23	0.48	0.56	0.18	38.8
-18	62.7	0.25	0.23	0.48	0.56	0.19	50.9

CALCULATED HOLDUPS, RUN 211--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 17.8 VOL %
TEMPERATURE : 80. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
211-01	36.1	0.0	0.81	0.76	0.0
-02	35.8	0.05	0.74	0.80	0.10
-03	36.7	0.10	0.70	0.76	0.15
-04	37.2	0.15	0.67	0.74	0.19
-05	36.0	0.20	0.66	0.73	0.20
-06	36.5	0.24	0.65	0.73	0.21
-07	42.3	0.0	0.81	0.75	0.0
-08	41.1	0.05	0.73	0.80	0.11
-09	42.5	0.10	0.70	0.76	0.14
-10	42.8	0.15	0.68	0.74	0.18
-11	43.0	0.20	0.67	0.73	0.19
-12	42.2	0.24	0.66	0.73	0.20
-13	62.0	0.0	0.81	0.75	0.0
-14	62.8	0.05	0.72	0.78	0.13
-15	63.9	0.10	0.64	0.70	0.22
-16	61.8	0.15	0.66	0.72	0.20
-17	62.9	0.20	0.65	0.71	0.21
-18	62.7	0.25	0.63	0.71	0.24

Z BED EXPANSION FOR RUN 212

CATALYST : NDS-2A *L/d=3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
212- 1	30.8	0.0	57.	14.
- 2	31.7	0.10	67.	34.
- 3	45.0	0.0	65.	30.
- 4	44.6	0.04	70.	40.
- 5	44.0	0.10	77.	54.
- 6	44.7	0.14	81.	62.
- 7	66.8	0.0	84.	68.
- 8	66.4	0.11	95.	90.
- 9	66.5	0.16	102.	104.
-10	88.7	0.0	109.	118.
-11	88.4	0.12	119.	138.
-12	45.2	0.19	82.	64.
-13	66.5	0.21	111.	122.
-14	88.3	0.23	134.	168.
-15	87.8	0.16	127.	154.

CALCULATED HOLDUPS, RUN 212: DENSE PHASE

M80-21
-152

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
212- 1	30.8	0.0	0.43	0.59	0.63	0.0	0.0
- 2	31.7	0.10	0.37	0.47	0.46	0.17	18.4
- 3	45.0	0.0	0.38	0.66	0.64	0.0	0.0
- 4	44.6	0.04	0.35	0.55	0.55	0.10	7.3
- 5	44.0	0.10	0.32	0.49	0.50	0.19	15.4
- 6	44.7	0.14	0.30	0.46	0.47	0.24	21.9
- 7	66.8	0.0	0.29	0.72	0.72	0.0	0.0
- 8	66.4	0.11	0.26	0.57	0.58	0.17	16.8
- 9	66.5	0.16	0.24	0.54	0.54	0.22	23.2
-10	88.7	0.0	0.22	0.80	0.77	0.0	0.0
-11	88.4	0.12	0.21	0.65	0.65	0.15	19.5
-12	45.2	0.19	0.30	0.42	0.45	0.28	27.8
-13	66.5	0.21	0.22	0.51	0.53	0.27	28.9
-14	88.3	0.23	0.18	0.57	0.58	0.25	32.3
-15	87.8	0.16	0.19	0.63	0.64	0.17	27.7

CALCULATED HOLDUPS, RUN 212--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGB
212- 1	30.8	0.0	0.97	0.94	0.0
- 2	31.7	0.10	0.79	0.87	0.21
- 3	45.0	0.0	0.98	0.94	0.0
- 4	44.6	0.04	0.88	0.96	0.12
- 5	44.0	0.10	0.76	0.85	0.24
- 6	44.7	0.14	0.69	0.78	0.31
- 7	66.8	0.0	0.99	0.94	0.0
- 8	66.4	0.11	0.77	0.86	0.23
- 9	66.5	0.16	0.71	0.79	0.29
-10	88.7	0.0	0.97	0.94	0.0
-11	88.4	0.12	0.79	0.88	0.21
-12	45.2	0.19	0.68	0.76	0.32
-13	66.5	0.21	0.65	0.73	0.35
-14	88.3	0.23	0.67	0.80	0.33
-15	87.8	0.16	0.75	0.83	0.25

% BED EXPANSION FOR RUN 213

CATALYST : MDS-2A *l/d=3*
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 70. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
213- 1	38.4	0.0	63.	26.
- 2	37.3	0.05	64.	28.
- 3	37.4	0.10	68.	36.
- 4	37.4	0.15	71.	42.
- 5	44.3	0.05	69.	38.
- 6	68.4	0.05	86.	79.
- 7	89.1	0.05	109.	127.
- 8	45.0	0.10	70.	46.
- 9	36.7	0.20	69.	44.
-10	44.4	0.20	76.	58.
-11	66.4	0.20	93.	94.
-12	37.5	0.25	69.	44.
-13	44.0	0.25	75.	56.
-14	66.5	0.25	99.	106.
-15	88.1	0.25	125.	160.
-16	88.4	0.20	120.	150.
-17	88.7	0.10	111.	131.
-18	88.7	0.15	117.	144.
-19	44.4	0.15	76.	58.
-20	66.6	0.15	96.	109.
-21	66.7	0.10	91.	90.
-22	109.5	0.05	135.	181.

CALCULATED HOLDUPS, RUN 213: DENSE PHASE

CATALYST : MDS-2A
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 70. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (Mm/Sec)
213- 1	38.4	0.0	0.39	0.63	0.62	0.0	0.0
- 2	37.3	0.05	0.38	0.54	0.55	0.07	10.9
- 3	37.4	0.10	0.36	0.52	0.53	0.12	21.5
- 4	37.4	0.15	0.35	0.48	0.50	0.17	30.3
- 5	44.3	0.05	0.36	0.56	0.58	0.09	9.7
- 6	68.4	0.05	0.27	0.69	0.68	0.04	12.0
- 7	89.1	0.05	0.22	0.75	0.76	0.04	11.8
- 8	45.0	0.10	0.34	0.58	0.56	0.09	23.5
- 9	36.7	0.20	0.34	0.55	0.54	0.11	49.8
-10	44.4	0.20	0.31	0.57	0.56	0.12	47.6
-11	66.4	0.20	0.25	0.60	0.62	0.15	42.2
-12	37.5	0.25	0.34	0.50	0.52	0.16	57.1
-13	44.0	0.25	0.31	0.49	0.53	0.20	51.4
-14	66.5	0.25	0.24	0.59	0.60	0.17	52.0
-15	88.1	0.25	0.19	0.64	0.66	0.17	49.9
-16	88.4	0.20	0.20	0.65	0.67	0.16	39.4
-17	88.7	0.10	0.21	0.70	0.73	0.08	21.3
-18	88.7	0.15	0.20	0.67	0.68	0.13	29.8
-19	44.4	0.15	0.31	0.56	0.55	0.13	33.8
-20	66.6	0.15	0.24	0.60	0.61	0.16	28.7
-21	66.7	0.10	0.26	0.63	0.64	0.12	19.5
-22	109.5	0.05	0.17	0.79	0.80	0.03	11.6

CALCULATED HOLDUPS, RUN 213--DILUTE PHASE

CATALYST : MDS-2A
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 70. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EOB
213- 1	38.4	0.0	0.97	0.94	0.0
- 2	37.3	0.05	0.94	1.01	0.06
- 3	37.4	0.10	0.88	0.95	0.12
- 4	37.4	0.15	0.80	0.89	0.20
- 5	44.3	0.05	0.90	0.99	0.10
- 6	68.4	0.05	0.93	0.98	0.07
- 7	89.1	0.05	0.91	1.03	0.09
- 8	45.0	0.10	0.88	0.93	0.12
- 9	36.7	0.20	0.80	0.93	0.20
-10	44.4	0.20	0.78	0.91	0.22
-11	66.4	0.20	0.76	0.89	0.24
-12	37.5	0.25	0.78	0.91	0.22
-13	44.0	0.25	0.76	0.89	0.24
-14	66.5	0.25	0.72	0.85	0.28
-15	88.1	0.25	0.73	0.86	0.27
-16	88.4	0.20	0.76	0.89	0.24
-17	88.7	0.10	0.84	0.98	0.16
-18	88.7	0.15	0.78	0.91	0.22
-19	44.4	0.15	0.77	0.90	0.23
-20	66.6	0.15	0.76	0.89	0.24
-21	66.7	0.10	0.81	0.95	0.19
-22	109.5	0.05	0.90	1.07	0.10

% BED EXPANSION FOR RUN 214

CATALYST : HDS-2A *l/d=3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
214- 1	45.8	0.0	74.	40.
- 2	43.4	0.0	69.	30.
- 3	39.3	0.0	63.	19.
- 4	31.2	0.0	58.	9.
- 5	53.9	0.0	81.	53.
- 6	60.3	0.0	86.	62.
- 7	70.0	0.0	90.	70.
- 8	71.4	0.0	96.	81.
- 9	79.7	0.0	105.	98.
-10	71.3	0.0	106.	112.
-11	88.3	0.0	128.	156.
-12	47.1	0.10	78.	56.
-13	46.0	0.14	74.	48.

CALCULATED HOLDUPS, RUN 214: DENSE PHASE

M80-21
-157

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Nm/Sec)
214- 1	45.8	0.0	0.35	0.63	0.63	0.0	0.0
- 2	43.4	0.0	0.38	0.63	0.62	0.0	0.0
- 3	39.3	0.0	0.41	0.58	0.59	0.0	0.0
- 4	31.2	0.0	0.45	0.56	0.59	0.0	0.0
- 5	53.9	0.0	0.32	0.65	0.64	0.0	0.0
- 6	60.3	0.0	0.30	0.67	0.65	0.0	0.0
- 7	70.0	0.0	0.29	0.68	0.66	0.0	0.0
- 8	71.4	0.0	0.27	0.69	0.68	0.0	0.0
- 9	79.7	0.0	0.25	0.70	0.69	0.0	0.0
-10	71.3	0.0	0.23	0.71	0.66	0.0	0.0
-11	88.3	0.0	0.19	0.72	0.67	0.0	0.0
-12	47.1	0.10	0.31	0.41	0.41	0.20	12.4
-13	46.0	0.14	0.33	0.41	0.41	0.19	24.6

CALCULATED HOLDUPS, RUN 214--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
214- 1	45.8	0.0	0.81	0.76	0.0
- 2	43.4	0.0	0.81	0.76	0.0
- 3	39.3	0.0	0.82	0.76	0.0
- 4	31.2	0.0	0.81	0.76	0.0
- 5	53.9	0.0	0.82	0.76	0.0
- 6	60.3	0.0	0.81	0.76	0.0
- 7	70.0	0.0	0.82	0.75	0.0
- 8	71.4	0.0	0.81	0.75	0.0
- 9	79.7	0.0	0.82	0.75	0.0
-10	71.3	0.0	0.82	0.80	0.0
-11	88.3	0.0	0.84	0.76	0.0
-12	47.1	0.10	0.70	0.78	0.17
-13	46.0	0.14	0.69	0.78	0.18

X BED EXPANSION FOR RUN 215

M80-21
-158

CATALYST : HDS-2A *lid = 3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
215- 1	40.4	0.05	84.	110.
- 2	43.5	0.09	85.	113.
- 3	44.1	0.14	83.	108.
- 4	67.3	0.04	128.	220.
- 5	67.8	0.10	139.	248.
- 6	68.1	0.14	158.	295.

CALCULATED HOLDUPS, RUN 215: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
215- 1	40.4	0.05	0.23	0.58	0.64	0.08	11.3
- 2	43.5	0.09	0.23	0.54	0.59	0.13	19.4
- 3	44.1	0.14	0.24	0.52	0.59	0.15	29.3
- 4	67.3	0.04	0.15	0.64	0.71	0.09	7.7
- 5	67.8	0.10	0.14	0.59	0.64	0.16	17.1
- 6	68.1	0.14	0.12	0.63	0.69	0.13	31.6

CALCULATED HOLDUPS, RUN 215--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 71. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
215- 1	40.4	0.05	0.75	0.84	0.11
- 2	43.5	0.09	0.71	0.81	0.16
- 3	44.1	0.14	0.69	0.78	0.18
- 4	67.3	0.04	0.74	0.81	0.12
- 5	67.8	0.10	0.65	0.75	0.23
- 6	68.1	0.14	0.51	0.77	0.40

% BED EXPANSION FOR RUN 216

CATALYST : HDS-2A
GAS : HELIUM *l/d: 3*
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 73. DEG F

Run No.	Liquid Flow Rate, GPM/Ft2	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
216- 1	44.5	0.10	85.	57.
- 4	39.6	0.10	77.	43.
- 5	39.9	0.05	75.	39.
- 6	47.3	0.05	86.	59.
- 7	47.6	0.15	79.	46.
- 8	67.4	0.15	124.	130.
- 9	66.8	0.10	123.	128.
-10	38.1	0.20	82.	52.
-11	37.7	0.25	75.	39.
-12	46.0	0.25	84.	56.
-13	46.2	0.20	92.	70.
-14	67.9	0.20	131.	143.
-15	69.5	0.25	127.	135.
-16	39.8	0.15	71.	32.

CALCULATED HOLDUPS, RUN 216: DENSE PHASE

M80-21
-160

CATALYST : HDS-2A
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 73. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
216- 1	44.5	0.10	0.31	0.47	0.56	0.13	20.7
- 4	39.6	0.10	0.34	0.46	0.53	0.12	21.7
- 5	39.9	0.05	0.35	0.45	0.52	0.11	8.5
- 6	47.3	0.05	0.31	0.50	0.57	0.10	9.1
- 7	47.6	0.15	0.33	0.44	0.53	0.15	31.0
- 8	67.4	0.15	0.21	0.54	0.64	0.15	30.1
- 9	66.8	0.10	0.22	0.58	0.66	0.10	21.1
-10	38.1	0.20	0.32	0.46	0.49	0.13	48.0
-11	37.7	0.25	0.35	0.42	0.45	0.15	58.6
-12	46.0	0.25	0.31	0.46	0.48	0.14	58.6
-13	46.2	0.20	0.29	0.51	0.52	0.11	48.9
-14	67.9	0.20	0.20	0.56	0.47	0.14	44.2
-15	69.5	0.25	0.21	0.53	0.72	0.16	53.7
-16	39.8	0.15	0.37	0.41	0.49	0.14	32.6

CALCULATED HOLDUPS, RUN 216--DILUTE PHASE

CATALYST : HDS-2A
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 15.5 VOL %
TEMPERATURE : 73. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
216- 1	44.5	0.10	0.76	0.85	0.09
- 4	39.6	0.10	0.77	0.85	0.09
- 5	39.9	0.05	0.78	0.87	0.07
- 6	47.3	0.05	0.77	0.86	0.08
- 7	47.6	0.15	0.76	0.83	0.10
- 8	67.4	0.15	0.73	0.81	0.14
- 9	66.8	0.10	0.74	0.81	0.13
-10	38.1	0.20	0.75	0.83	0.11
-11	37.7	0.25	0.73	0.83	0.13
-12	46.0	0.25	0.74	0.82	0.12
-13	46.2	0.20	0.74	0.83	0.13
-14	67.9	0.20	0.73	0.93	0.13
-15	69.5	0.25	0.72	0.82	0.14
-16	39.8	0.15	0.74	0.84	0.12

Z BED EXPANSION FOR RUN 217

CATALYST : HDS-2A *2/d=3*
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 15.5 VOL %
 TEMPERATURE : 73. DEG F

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Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate, Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
217- 1	44.2	0.05	93.	72.
- 2	46.0	0.10	91.	69.
- 3	44.2	0.14	88.	63.
- 4	67.3	0.05	131.	143.
- 5	68.4	0.10	138.	156.
- 6	67.9	0.15	127.	135.

CALCULATED HOLDUPS, RUN 217: DENSE PHASE

CATALYST : HDS-2A
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 15.5 VOL %
 TEMPERATURE : 73. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
217- 1	44.2	0.05	0.28	0.52	0.54	0.10	8.4
- 2	46.0	0.10	0.29	0.48	0.51	0.14	19.3
- 3	44.2	0.14	0.30	0.48	0.51	0.13	31.1
- 4	67.3	0.05	0.20	0.61	0.66	0.08	8.0
- 5	68.4	0.10	0.19	0.55	0.59	0.16	16.4
- 6	67.9	0.15	0.21	0.53	0.56	0.16	27.2

CALCULATED HOLDUPS, RUN 217--DILUTE PHASE

CATALYST : HDS-2A
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 15.5 VOL %
 TEMPERATURE : 73. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	E6G
217- 1	44.2	0.05	0.75	0.83	0.11
- 2	46.0	0.10	0.74	0.82	0.12
- 3	44.2	0.14	0.72	0.80	0.15
- 4	67.3	0.05	0.75	0.84	0.12
- 5	68.4	0.10	0.69	0.78	0.18
- 6	67.9	0.15	0.71	0.81	0.16

CALCULATED HOLDUPS, RUN 300: DENSE PHASE

CATALYST : NONE *l/d=0*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 84. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (MM/Sec)
300- 1	22.4	0.05	0.0	0.95	0.98	0.05	13.7
- 2	22.3	0.10	0.0	0.88	0.92	0.12	25.2
- 3	22.3	0.15	0.0	0.86	0.89	0.14	37.0
- 4	22.4	0.20	0.0	0.82	0.88	0.18	47.0
- 5	22.4	0.25	0.0	0.79	0.87	0.21	56.8
- 6	44.9	0.05	0.0	0.92	0.96	0.08	11.5
- 7	44.6	0.10	0.0	0.84	0.85	0.16	20.7
- 8	44.9	0.15	0.0	0.79	0.87	0.21	29.7
- 9	44.5	0.20	0.0	0.70	0.73	0.30	33.7
-10	45.2	0.23	0.0	0.68	0.73	0.32	38.1
-11	66.6	0.05	0.0	0.91	0.93	0.09	9.4
-12	66.9	0.10	0.0	0.79	0.80	0.21	14.8
-13	67.3	0.15	0.0	0.73	0.74	0.27	21.4
-14	66.6	0.21	0.0	0.68	0.70	0.32	29.7
-15	66.7	0.25	0.0	0.67	0.69	0.33	36.7
-16	88.7	0.05	0.0	0.93	0.94	0.07	10.1
-17	88.9	0.11	0.0	0.84	0.84	0.16	18.9
-18	89.7	0.16	0.0	0.78	0.81	0.22	25.8
-19	89.8	0.23	0.0	0.73	0.74	0.27	33.8
-20	112.3	0.05	0.0	0.95	0.98	0.05	9.8

CALCULATED HOLDUPS, RUN 300--DILUTE PHASE

CATALYST : NONE
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 84. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELBP	EOB
300- 1	22.4	0.05	0.94	0.96	0.06
- 2	22.3	0.10	0.85	0.89	0.15
- 3	22.3	0.15	0.79	0.84	0.21
- 4	22.4	0.20	0.77	0.79	0.23
- 5	22.4	0.25	0.73	0.78	0.27
- 6	44.9	0.05	0.91	0.94	0.09
- 7	44.6	0.10	0.82	0.86	0.18
- 8	44.9	0.15	0.73	0.76	0.27
- 9	44.5	0.20	0.64	0.70	0.36
-10	45.2	0.23	0.64	0.70	0.36
-11	66.6	0.05	0.89	0.95	0.11
-12	66.9	0.10	0.76	0.82	0.24
-13	67.3	0.15	0.71	0.77	0.29
-14	66.6	0.21	0.67	0.72	0.33
-15	66.7	0.25	0.64	0.70	0.36
-16	88.7	0.05	0.91	0.97	0.09
-17	88.9	0.11	0.82	0.88	0.18
-18	89.7	0.16	0.76	0.85	0.24
-19	89.8	0.23	0.70	0.77	0.30
-20	112.3	0.05	0.94	1.00	0.06

Z BED EXPANSION FOR RUN 301

CATALYST : HDS-2A *Q/d=6*
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 74. DEG F

Run No.	Liquid Flow Rate, GPH/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
301- 1	37.6	0.0	47.	7.
- 2	44.0	0.0	49.	11.
- 3	58.9	0.0	51.	16.
- 4	66.0	0.0	51.	16.
- 5	76.4	0.0	55.	25.
- 6	88.0	0.0	63.	43.
- 7	119.0	0.0	97.	121.
- 8	38.4	0.04	50.	22.
- 9	37.9	0.09	51.	24.
-10	40.4	0.14	51.	24.
-11	44.6	0.10	54.	32.
-12	44.2	0.14	55.	34.
-13	67.4	0.10	67.	63.
-14	66.6	0.15	71.	73.
-15	88.4	0.11	83.	102.
-16	88.3	0.16	93.	127.

CALCULATED HOLDUPS, RUN 301: DENSE PHASE

CATALYST : MDS-2A
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 74. DEG F

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Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (Mm/Sec)
301- 1	37.6	0.0	0.42	0.59	0.74	0.0	0.0
- 2	44.0	0.0	0.41	0.59	0.72	0.0	0.0
- 3	58.9	0.0	0.39	0.60	0.72	0.0	0.0
- 4	66.0	0.0	0.39	0.61	0.72	0.0	0.0
- 5	76.4	0.0	0.36	0.63	0.71	0.0	0.0
- 6	88.0	0.0	0.32	0.67	0.70	0.0	0.0
- 7	119.0	0.0	0.21	0.79	0.78	0.0	0.0
- 8	38.4	0.04	0.37	0.57	0.46	0.06	8.9
- 9	37.9	0.09	0.36	0.59	0.41	0.04	25.3
-10	40.4	0.14	0.36	0.56	0.38	0.08	34.4
-11	44.6	0.10	0.34	0.65	0.45	0.01	28.7
-12	44.2	0.14	0.34	0.65	0.42	0.01	40.8
-13	67.4	0.10	0.28	0.60	0.58	0.13	19.4
-14	66.6	0.15	0.26	0.54	0.55	0.20	23.4
-15	88.4	0.11	0.22	0.65	0.67	0.12	20.2
-16	88.3	0.16	0.20	0.63	0.65	0.17	27.6

CALCULATED HOLDUPS, RUN 301--DILUTE PHASE

CATALYST : MDS-2A
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 74. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EOG
301- 1	37.6	0.0	0.97	0.97	0.0
- 2	44.0	0.0	0.99	0.96	0.0
- 3	58.9	0.0	0.97	0.96	0.0
- 4	66.0	0.0	0.98	0.96	0.0
- 5	76.4	0.0	0.97	0.96	0.0
- 6	88.0	0.0	0.98	0.96	0.0
- 7	119.0	0.0	0.98	0.96	0.0
- 8	38.4	0.04	0.87	0.93	0.13
- 9	37.9	0.09	0.76	0.81	0.24
-10	40.4	0.14	0.70	0.76	0.30
-11	44.6	0.10	0.74	0.80	0.26
-12	44.2	0.14	0.67	0.72	0.33
-13	67.4	0.10	0.75	0.80	0.25
-14	66.6	0.15	0.71	0.76	0.29
-15	88.4	0.11	0.81	0.86	0.19
-16	88.3	0.16	0.76	0.81	0.24

% BED EXPANSION FOR RUN 310

CATALYST : HDS-2A *l/d: 3*
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, GPH/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
310- 1	40.0	0.0	44.	19.
- 2	44.6	0.0	45.	22.
- 3	37.2	0.14	51.	38.
- 4	88.5	0.0	71.	92.
- 5	114.1	0.0	91.	146.
- 6	128.2	0.0	115.	211.
- 7	38.6	0.04	47.	27.
- 8	38.8	0.09	50.	35.
- 9	43.8	0.04	44.	19.
-10	44.7	0.10	57.	54.
-11	66.0	0.05	63.	70.
-12	66.1	0.10	70.	89.
-13	66.8	0.0	59.	59.
-14	39.3	0.18	51.	38.
-15	38.5	0.21	49.	32.
-16	44.7	0.14	57.	54.
-17	44.5	0.19	55.	49.
-18	44.2	0.24	50.	35.
-19	67.0	0.15	70.	89.
-20	66.8	0.20	70.	89.

CALCULATED HOLDUPS, RUN 310: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (MM/Sec)
310- 1	40.0	0.0	0.47	0.41	0.59	0.0	0.0
- 2	44.6	0.0	0.46	0.43	0.60	0.0	0.0
- 3	37.2	0.14	0.40	0.47	0.37	0.13	29.8
- 4	88.5	0.0	0.29	0.66	0.65	0.0	0.0
- 5	114.1	0.0	0.23	0.73	0.71	0.0	0.0
- 6	128.2	0.0	0.18	0.77	0.81	0.0	0.0
- 7	38.6	0.04	0.44	0.56	0.40	0.00	12.6
- 8	38.8	0.09	0.41	0.61	0.37	-0.02	30.0
- 9	43.8	0.04	0.47	0.54	0.32	-0.01	13.8
-10	44.7	0.10	0.36	0.56	0.44	0.08	22.5
-11	66.0	0.05	0.33	0.62	0.60	0.05	10.3
-12	66.1	0.10	0.29	0.55	0.51	0.15	16.3
-13	66.8	0.0	0.35	0.59	0.60	0.0	0.0
-14	39.3	0.18	0.40	0.45	0.35	0.15	40.4
-15	38.5	0.21	0.42	0.43	0.31	0.15	47.4
-16	44.7	0.14	0.36	0.48	0.48	0.16	27.2
-17	44.5	0.19	0.37	0.45	0.44	0.18	37.3
-18	44.2	0.24	0.41	0.41	0.36	0.18	48.3
-19	67.0	0.15	0.29	0.56	0.62	0.15	28.7
-20	66.8	0.20	0.29	0.49	0.49	0.22	32.1

CALCULATED HOLDUPS, RUN 310--DILUTE PHASE

CATALYST : NDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 72. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	EL6	ELDP	EGG
310- 1	40.0	0.0	0.98	0.97	0.0
- 2	44.6	0.0	0.98	0.97	0.0
- 3	37.2	0.14	0.74	0.78	0.26
- 4	88.5	0.0	0.98	0.96	0.0
- 5	114.1	0.0	0.98	0.96	0.0
- 6	128.2	0.0	0.98	0.98	0.0
- 7	38.6	0.04	0.88	0.93	0.12
- 8	38.8	0.09	0.77	0.82	0.23
- 9	43.8	0.04	0.88	0.91	0.12
-10	44.7	0.10	0.74	0.77	0.26
-11	66.0	0.05	0.84	0.88	0.14
-12	66.1	0.10	0.70	0.73	0.30
-13	66.8	0.0	0.98	0.96	0.0
-14	39.3	0.18	0.72	0.76	0.28
-15	38.5	0.21	0.70	0.76	0.30
-16	44.7	0.14	0.71	0.77	0.29
-17	44.5	0.19	0.69	0.76	0.31
-18	44.2	0.24	0.67	0.75	0.33
-19	67.0	0.15	0.70	0.82	0.30
-20	66.8	0.20	0.66	0.72	0.34

Z BED EXPANSION FOR RUN 311

CATALYST : HDS-2A *2/d-3*
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 77. DEG F

Run No.	Liquid Flow Rate, GPH/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
311- 1	37.6	0.20	50.	35.
- 2	44.7	0.20	55.	49.
- 3	67.5	0.20	64.	73.
- 4	90.2	0.20	81.	119.
- 5	37.9	0.25	48.	30.
- 6	44.5	0.25	56.	51.
- 7	64.1	0.25	70.	89.
- 8	37.4	0.15	50.	35.
- 9	44.7	0.15	54.	46.
-10	67.0	0.15	65.	76.
-11	88.5	0.15	80.	116.
-12	37.9	0.05	47.	27.
-13	38.1	0.10	46.	24.
-14	44.8	0.10	51.	38.
-15	44.3	0.05	46.	24.
-16	67.9	0.05	59.	59.
-17	67.5	0.10	63.	70.
-18	89.9	0.10	76.	105.
-19	88.5	0.05	71.	92.

CALCULATED HOLDUPS, RUN 311: DENSE PHASE

CATALYST : NDS-2A
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 77. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mm/Sec)
311- 1	37.6	0.20	0.41	0.54	0.39	0.05	55.3
- 2	44.7	0.20	0.37	0.55	0.46	0.08	52.1
- 3	67.5	0.20	0.32	0.56	0.54	0.12	45.4
- 4	90.2	0.20	0.25	0.66	0.68	0.09	48.0
- 5	37.9	0.25	0.43	0.56	0.36	0.02	74.2
- 6	44.5	0.25	0.37	0.51	0.47	0.13	60.2
- 7	64.1	0.25	0.29	0.55	0.56	0.16	53.2
- 8	37.4	0.15	0.41	0.57	0.41	0.02	44.2
- 9	44.7	0.15	0.38	0.51	0.46	0.11	35.3
-10	67.0	0.15	0.32	0.57	0.55	0.12	31.9
-11	88.5	0.15	0.26	0.62	0.62	0.12	30.2
-12	37.9	0.05	0.44	0.57	0.45	-0.01	15.6
-13	38.1	0.10	0.45	0.52	0.39	0.03	28.0
-14	44.8	0.10	0.40	0.59	0.47	0.01	29.6
-15	44.3	0.05	0.45	0.54	0.44	0.01	14.4
-16	67.9	0.05	0.35	0.65	0.66	0.00	15.0
-17	67.5	0.10	0.33	0.63	0.67	0.04	26.4
-18	89.9	0.10	0.27	0.68	0.72	0.05	24.8
-19	88.5	0.05	0.29	0.70	0.73	0.01	14.3

CALCULATED HOLDUPS, RUN 311--DILUTE PHASE

CATALYST : HDS-2A
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 77. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
311- 1	37.6	0.20	0.80	0.87	0.20
- 2	44.7	0.20	0.77	0.84	0.23
- 3	67.5	0.20	0.76	0.83	0.24
- 4	90.2	0.20	0.79	0.86	0.21
- 5	37.9	0.25	0.76	0.84	0.24
- 6	44.5	0.25	0.75	0.82	0.25
- 7	64.1	0.25	0.71	0.78	0.29
- 8	37.4	0.15	0.80	0.87	0.20
- 9	44.7	0.15	0.78	0.84	0.22
-10	67.0	0.15	0.77	0.83	0.23
-11	88.5	0.15	0.77	0.84	0.23
-12	37.9	0.05	0.91	0.97	0.09
-13	38.1	0.10	0.86	0.92	0.14
-14	44.8	0.10	0.85	0.91	0.15
-15	44.3	0.05	0.91	0.97	0.09
-16	67.9	0.05	0.92	0.98	0.08
-17	67.5	0.10	0.85	0.92	0.15
-18	89.9	0.10	0.86	0.93	0.14
-19	88.5	0.05	0.92	0.98	0.08

Z. BED EXPANSION FOR RUN 320

CATALYST : MDS-2A *L/d = 2*
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 75. DEG F

Run No.	Liquid Flow Rate, GPM/ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
320- 1	38.2	0.0	63.	37.
- 2	47.0	0.0	65.	41.
- 3	69.1	0.0	86.	87.
- 4	89.7	0.0	106.	141.
- 5	109.3	0.0	141.	207.
- 6	38.1	0.20	67.	46.
- 7	44.9	0.20	71.	54.
- 8	67.3	0.20	94.	104.
- 9	89.0	0.20	124.	170.
-10	89.1	0.25	126.	174.
-11	67.1	0.25	101.	120.
-12	44.6	0.25	70.	52.
-13	38.1	0.25	66.	44.
-14	38.2	0.05	62.	38.
-15	38.3	0.10	66.	47.
-16	38.2	0.15	66.	47.
-17	43.0	0.05	65.	44.
-18	44.6	0.10	71.	58.
-19	42.9	0.15	75.	67.
-20	68.0	0.05	92.	104.
-21	66.2	0.10	95.	111.
-22	66.7	0.15	96.	113.
-23	89.7	0.05	111.	152.
-24	88.5	0.10	114.	153.
-25	89.1	0.15	121.	169.

CALCULATED HOLDUPS, RUN 320: DENSE PHASE

CATALYST : NDS-2A
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 75. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Mn/Sec)
320- 1	38.2	0.0	0.36	0.66	1.84	0.0	0.0
- 2	47.0	0.0	0.35	0.68	1.84	0.0	0.0
- 3	69.1	0.0	0.26	0.77	1.84	0.0	0.0
- 4	89.7	0.0	0.20	0.81	0.81	0.0	0.0
- 5	109.3	0.0	0.16	0.86	1.84	0.0	0.0
- 6	38.1	0.20	0.34	0.49	0.39	0.17	43.2
- 7	44.9	0.20	0.32	0.50	0.40	0.18	40.6
- 8	67.3	0.20	0.24	0.59	0.50	0.17	40.0
- 9	89.0	0.20	0.18	0.67	0.58	0.15	40.8
-10	89.1	0.25	0.18	0.64	0.53	0.18	48.0
-11	67.1	0.25	0.22	0.55	0.47	0.23	44.4
-12	44.6	0.25	0.32	0.47	0.38	0.21	49.7
-13	38.1	0.25	0.34	0.48	0.39	0.18	54.9
-14	38.2	0.05	0.36	0.57	0.55	0.08	10.8
-15	38.3	0.10	0.33	0.56	0.53	0.11	22.9
-16	38.2	0.15	0.33	0.50	0.46	0.17	30.6
-17	43.0	0.05	0.34	0.57	0.53	0.09	9.7
-18	44.6	0.10	0.31	0.55	0.52	0.14	19.7
-19	42.9	0.15	0.29	0.54	0.49	0.16	31.1
-20	68.0	0.05	0.24	0.68	0.65	0.08	8.6
-21	66.2	0.10	0.23	0.63	0.59	0.14	17.3
-22	66.7	0.15	0.23	0.60	0.56	0.17	27.1
-23	89.7	0.05	0.19	0.77	0.75	0.04	11.8
-24	88.5	0.10	0.19	0.72	0.71	0.09	21.3
-25	89.1	0.15	0.18	0.70	0.68	0.11	31.8

CALCULATED HOLDUPS, RUN 320--DILUTE PHASE

CATALYST : MDS-2A
GAS : HELIUM
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 75. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	EL8	ELBP	EGG
320- 1	38.2	0.0	0.98	1.84	0.0
- 2	47.0	0.0	0.98	1.84	0.0
- 3	69.1	0.0	0.98	1.84	0.0
- 4	89.7	0.0	0.98	1.00	0.0
- 5	109.3	0.0	0.97	1.84	0.0
- 6	38.1	0.20	0.83	0.85	0.17
- 7	44.9	0.20	0.82	0.84	0.18
- 8	67.3	0.20	0.79	0.84	0.21
- 9	89.0	0.20	0.82	0.86	0.18
-10	89.1	0.25	0.76	0.82	0.24
-11	67.1	0.25	0.73	0.78	0.27
-12	44.6	0.25	0.76	0.83	0.24
-13	38.1	0.25	0.78	0.85	0.22
-14	38.2	0.05	0.94	0.94	0.06
-15	38.3	0.10	0.88	0.89	0.12
-16	38.2	0.15	0.83	0.84	0.17
-17	43.0	0.05	0.92	0.93	0.08
-18	44.6	0.10	0.86	0.87	0.14
-19	42.9	0.15	0.80	0.81	0.20
-20	68.0	0.05	0.90	0.92	0.10
-21	66.2	0.10	0.82	0.83	0.18
-22	66.7	0.15	0.78	0.81	0.22
-23	89.7	0.05	0.95	0.96	0.05
-24	88.5	0.10	0.89	0.92	0.11
-25	89.1	0.15	0.84	0.87	0.16

Z BED EXPANSION FOR RUN 321

CATALYST : MDS-2A *l/d:2*
 GAS : NITROGEN
 LIQUID : KEROSENE
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 77. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
321- 1	36.9	0.04	66.	47.
- 2	39.4	0.09	71.	58.
- 3	42.9	0.05	70.	56.
- 4	43.2	0.10	76.	69.
- 5	67.6	0.05	91.	102.
- 6	38.4	0.14	66.	47.
- 7	38.1	0.19	67.	49.
- 8	38.9	0.24	66.	47.
- 9	44.7	0.14	76.	69.
-10	45.0	0.20	72.	60.
-11	44.4	0.26	71.	58.
-12	67.0	0.16	98.	118.
-13	67.5	0.11	91.	102.
-14	67.0	0.21	107.	138.
-15	67.3	0.27	109.	142.
-16	88.4	0.06	111.	147.
-17	88.5	0.12	123.	173.

CALCULATED HOLDUPS, RUN 321: DENSE PHASE

M80-21
-176

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 77. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELBB	ELDPB	EBB	Vcd (Mn/Sec)
321- 1	36.9	0.04	0.33	0.56	0.54	0.11	7.5
- 2	39.4	0.09	0.31	0.51	0.49	0.18	15.2
- 3	42.9	0.05	0.31	0.57	0.54	0.11	8.1
- 4	43.2	0.10	0.29	0.50	0.49	0.21	13.7
- 5	67.6	0.05	0.24	0.63	0.62	0.13	6.2
- 6	38.4	0.14	0.33	0.45	0.45	0.22	23.2
- 7	38.1	0.19	0.33	0.45	0.45	0.22	34.7
- 8	38.9	0.24	0.33	0.45	0.45	0.21	48.3
- 9	44.7	0.14	0.29	0.49	0.47	0.22	23.3
-10	45.0	0.20	0.31	0.45	0.44	0.24	32.5
-11	44.4	0.26	0.31	0.47	0.46	0.22	49.6
-12	67.0	0.16	0.22	0.54	0.59	0.23	22.3
-13	67.5	0.11	0.24	0.59	0.59	0.17	16.5
-14	67.0	0.21	0.21	0.54	0.52	0.25	33.0
-15	67.3	0.27	0.20	0.51	0.50	0.29	38.7
-16	88.4	0.06	0.20	0.73	0.74	0.07	10.1
-17	88.5	0.12	0.18	0.69	0.69	0.13	20.2

CALCULATED HOLDUPS, RUN 321--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : KEROSENE
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 77. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELD	ELDP	EGG
321- 1	36.9	0.04	0.87	0.91	0.13
- 2	39.4	0.09	0.77	0.81	0.23
- 3	42.9	0.05	0.87	0.89	0.13
- 4	43.2	0.10	0.76	0.79	0.24
- 5	67.6	0.05	0.84	0.88	0.16
- 6	38.4	0.14	0.76	0.80	0.24
- 7	38.1	0.19	0.75	0.72	0.25
- 8	38.9	0.24	0.71	0.75	0.29
- 9	44.7	0.14	0.73	0.76	0.27
-10	45.0	0.20	0.72	0.75	0.28
-11	44.4	0.26	0.70	0.74	0.30
-12	67.0	0.16	0.73	0.80	0.27
-13	67.5	0.11	0.79	0.83	0.21
-14	67.0	0.21	0.67	0.70	0.33
-15	67.3	0.27	0.62	0.66	0.38
-16	88.4	0.06	0.92	0.93	0.08
-17	88.5	0.12	0.82	0.85	0.18

Z BED EXPANSION FOR RUN 400

CATALYST : NONE *L/d = 0*
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 100. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (in.)	Z Bed Expansion
400-- 1	36.1	0.05	0.	0.
- 2	38.9	0.10	0.	0.
- 3	38.2	0.15	0.	0.
- 4	37.9	0.20	0.	0.
- 5	42.5	0.05	0.	0.
- 6	42.5	0.10	0.	0.
- 7	44.5	0.15	0.	0.
- 8	45.3	0.21	0.	0.
- 9	45.5	0.24	0.	0.
-10	38.2	0.24	0.	0.
-11	67.0	0.06	0.	0.
-12	72.1	0.12	0.	0.
-13	67.3	0.17	0.	0.

CALCULATED HOLDUPS, RUN 400: DENSE PHASE

M80-21
-178

CATALYST : NONE
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 100. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EBB	Vcd (Nm/Sec)
400- 1	36.1	0.05	0.0	0.92	0.95	0.08	11.7
- 2	38.9	0.10	0.0	0.90	0.95	0.10	24.2
- 3	38.2	0.15	0.0	0.90	0.93	0.10	37.2
- 4	37.9	0.20	0.0	0.87	0.92	0.13	48.9
- 5	42.5	0.05	0.0	0.93	0.95	0.07	11.8
- 6	42.5	0.10	0.0	0.90	0.93	0.10	23.8
- 7	44.5	0.15	0.0	0.88	0.92	0.12	37.9
- 8	45.3	0.21	0.0	0.86	0.91	0.14	51.7
- 9	45.5	0.24	0.0	0.86	0.91	0.14	60.0
-10	38.2	0.24	0.0	0.86	0.91	0.14	58.6
-11	67.0	0.06	0.0	0.93	0.97	0.07	12.5
-12	72.1	0.12	0.0	0.87	0.92	0.13	24.5
-13	67.3	0.17	0.0	0.85	0.89	0.15	36.5

CALCULATED HOLDUPS, RUN 400--DILUTE PHASE

CATALYST : NONE
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 100. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EBB
400- 1	36.1	0.05	0.91	0.89	0.09
- 2	38.9	0.10	0.90	0.88	0.10
- 3	38.2	0.15	0.90	0.88	0.10
- 4	37.9	0.20	0.86	0.86	0.14
- 5	42.5	0.05	0.90	0.88	0.10
- 6	42.5	0.10	0.89	0.88	0.11
- 7	44.5	0.15	0.88	0.87	0.12
- 8	45.3	0.21	0.86	0.86	0.14
- 9	45.5	0.24	0.86	0.85	0.14
-10	38.2	0.24	0.85	0.86	0.15
-11	67.0	0.06	0.91	0.88	0.09
-12	72.1	0.12	0.86	0.84	0.14
-13	67.3	0.17	0.86	0.84	0.14

% BED EXPANSION FOR RUN 401

CATALYST : NONE *l/d = 0*
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
401- 1	38.4	0.05	0.	0.
- 2	36.7	0.10	0.	0.
- 3	44.0	0.05	0.	0.
- 4	44.0	0.10	0.	0.
- 5	38.9	0.14	0.	0.
- 6	38.4	0.20	0.	0.
- 7	39.1	0.24	0.	0.
- 8	44.9	0.15	0.	0.
- 9	44.9	0.20	0.	0.
-10	44.7	0.25	0.	0.
-11	82.0	0.06	0.	0.
-12	78.4	0.11	0.	0.
-13	74.8	0.17	0.	0.

CALCULATED HOLDUPS, RUN 401: DENSE PHASE

M80-21
-180

CATALYST : NONE
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EEB	Vcd (Mn/Sec)
401- 1	38.4	0.05	0.0	0.89	0.96	0.11	10.7
- 2	36.7	0.10	0.0	0.87	0.94	0.13	22.5
- 3	44.0	0.05	0.0	0.89	0.80	0.11	9.9
- 4	44.0	0.10	0.0	0.84	0.92	0.16	20.7
- 5	38.9	0.14	0.0	0.85	0.79	0.15	33.5
- 6	38.4	0.20	0.0	0.84	0.78	0.16	45.9
- 7	39.1	0.24	0.0	0.82	0.78	0.18	55.8
- 8	44.9	0.15	0.0	0.84	0.78	0.16	33.3
- 9	44.9	0.20	0.0	0.82	0.76	0.18	44.9
-10	44.7	0.25	0.0	0.81	0.75	0.19	55.5
-11	82.0	0.06	0.0	0.92	0.82	0.08	11.7
-12	78.4	0.11	0.0	0.83	0.73	0.17	20.3
-13	74.8	0.17	0.0	0.80	0.76	0.20	30.2

CALCULATED HOLDUPS, RUN 401--DILUTE PHASE

CATALYST : NONE
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELB	ELDP	EEB
401- 1	38.4	0.05	0.88	0.85	0.12
- 2	36.7	0.10	0.87	0.84	0.13
- 3	44.0	0.05	0.86	0.81	0.14
- 4	44.0	0.10	0.84	0.82	0.16
- 5	38.9	0.14	0.85	0.81	0.15
- 6	38.4	0.20	0.82	0.82	0.18
- 7	39.1	0.24	0.81	0.80	0.19
- 8	44.9	0.15	0.85	0.82	0.15
- 9	44.9	0.20	0.81	0.79	0.19
-10	44.7	0.25	0.82	0.79	0.18
-11	82.0	0.06	0.90	0.86	0.10
-12	78.4	0.11	0.81	0.77	0.19
-13	74.8	0.17	0.79	0.75	0.21

Z BED EXPANSION FOR RUN 410

CATALYST : NONE *ℓ/d = 0*
 GAS : HELIUM
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
410- 1	38.5	0.20	0.	0.
- 2	38.4	0.25	0.	0.
- 3	45.1	0.25	0.	0.
- 4	44.8	0.20	0.	0.
- 5	71.3	0.20	0.	0.
- 6	71.2	0.25	0.	0.
- 7	38.8	0.05	0.	0.
- 8	39.0	0.10	0.	0.
- 9	38.2	0.15	0.	0.
-10	43.5	0.15	0.	0.
-11	44.3	0.10	0.	0.
-12	44.4	0.05	0.	0.
-13	71.2	0.05	0.	0.
-14	71.2	0.10	0.	0.
-15	70.5	0.15	0.	0.
-16	90.0	0.05	0.	0.
-17	90.1	0.10	0.	0.

CALCULATED HOLDUPS, RUN 410: DENSE PHASE

M80-21
-182

CATALYST : NONE
GAS : HELIUM
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELOB	ELDPB	EOB	Vcd (Mn/Sec)
410- 1	38.5	0.20	0.0	0.89	0.85	0.11	51.0
- 2	38.4	0.25	0.0	0.87	0.83	0.13	62.4
- 3	45.1	0.25	0.0	0.86	0.82	0.14	61.2
- 4	44.8	0.20	0.0	0.87	0.82	0.13	49.2
- 5	71.3	0.20	0.0	0.86	0.79	0.14	45.6
- 6	71.2	0.25	0.0	0.83	0.78	0.17	54.9
- 7	38.8	0.05	0.0	0.95	0.89	0.05	13.3
- 8	39.0	0.10	0.0	0.91	0.85	0.09	25.5
- 9	38.2	0.15	0.0	0.90	0.83	0.10	38.3
-10	43.5	0.15	0.0	0.87	0.84	0.13	36.3
-11	44.3	0.10	0.0	0.89	0.83	0.11	24.0
-12	44.4	0.05	0.0	0.92	0.86	0.08	11.4
-13	71.2	0.05	0.0	0.94	0.87	0.06	11.1
-14	71.2	0.10	0.0	0.89	0.84	0.11	21.5
-15	70.5	0.15	0.0	0.86	0.81	0.14	32.6
-16	90.0	0.05	0.0	0.95	0.90	0.05	11.7
-17	90.1	0.10	0.0	0.91	0.84	0.09	21.9

CALCULATED HOLDUPS, RUN 410--DILUTE PHASE

CATALYST : NONE
GAS : HELIUM
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELB	ELDP	EOB
410- 1	38.5	0.20	0.88	0.84	0.12
- 2	38.4	0.25	0.88	0.84	0.12
- 3	45.1	0.25	0.86	0.83	0.14
- 4	44.8	0.20	0.86	0.83	0.14
- 5	71.3	0.20	0.81	0.79	0.19
- 6	71.2	0.25	0.82	0.80	0.18
- 7	38.8	0.05	0.94	0.91	0.06
- 8	39.0	0.10	0.88	0.85	0.12
- 9	38.2	0.15	0.85	0.84	0.15
-10	43.5	0.15	0.88	0.85	0.12
-11	44.3	0.10	0.88	0.84	0.12
-12	44.4	0.05	0.90	0.86	0.10
-13	71.2	0.05	0.91	0.88	0.09
-14	71.2	0.10	0.86	0.82	0.14
-15	70.5	0.15	0.83	0.79	0.17
-16	90.0	0.05	0.93	0.90	0.07
-17	90.1	0.10	0.86	0.84	0.14

Z BED EXPANSION FOR RUN 420

M80-21
-183

CATALYST : HDS-2A *l/d = 2*
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 100. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
420- 1	22.0	0.0	46.	31.
- 2	39.4	0.0	71.	103.
- 3	41.3	0.0	87.	149.
- 4	11.5	0.0	39.	22.
- 5	9.3	0.0	36.	13.
- 6	53.3	0.0	108.	238.

CALCULATED HOLDUPS, RUN 420: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 100. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (Nm/Sec)
420- 1	22.0	0.0	0.42	0.63	0.64	0.0	0.0
- 2	39.4	0.0	0.27	0.74	0.67	0.0	0.0
- 3	41.3	0.0	0.22	0.77	0.69	0.0	0.0
- 4	11.5	0.0	0.45	0.55	0.76	0.0	0.0
- 5	9.3	0.0	0.49	0.49	0.70	0.0	0.0
- 6	53.3	0.0	0.16	0.81	2.18	0.0	0.0

CALCULATED HOLDUPS, RUN 420--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 100. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EOB
420- 1	22.0	0.0	0.98	1.03	0.0
- 2	39.4	0.0	0.99	1.05	0.0
- 3	41.3	0.0	0.99	1.05	0.0
- 4	11.5	0.0	1.00	1.04	0.0
- 5	9.3	0.0	1.00	1.05	0.0
- 6	53.3	0.0	0.99	0.05	0.0

Z BED EXPANSION FOR RUN 421

M80-21
-184

CATALYST : HDS-2A *l/d=3*
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, GPM/Ft2	Gas Flow Rate, Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
421- 1	10.3	0.0	37.	16.
- 2	22.6	0.0	45.	41.
- 3	38.9	0.0	63.	97.
- 4	44.1	0.0	69.	116.
- 5	71.2	0.0	119.	272.
- 6	11.4	0.05	36.	13.

CALCULATED HOLDUPS, RUN 421: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (Mm/Sec)
421- 1	10.3	0.0	0.48	0.49	0.67	0.0	0.0
- 2	22.6	0.0	0.39	0.61	0.64	0.0	0.0
- 3	38.9	0.0	0.28	0.71	0.64	0.0	0.0
- 4	44.1	0.0	0.26	0.73	0.67	0.0	0.0
- 5	71.2	0.0	0.15	0.83	0.76	0.0	0.0
- 6	11.4	0.05	0.49	0.46	0.32	0.04	13.5

CALCULATED HOLDUPS, RUN 421--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 125. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EOG
421- 1	10.3	0.0	0.99	1.06	0.0
- 2	22.6	0.0	0.98	1.06	0.0
- 3	38.9	0.0	0.99	1.05	0.0
- 4	44.1	0.0	0.99	1.05	0.0
- 5	71.2	0.0	0.96	1.05	0.0
- 6	11.4	0.05	0.93	0.90	0.07

Z BED EXPANSION FOR RUN 422

CATALYST : NDS-2A *L/d = 3*
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, GPH/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
422- 1	13.3	0.0	37.	16.
- 2	23.9	0.0	41.	28.
- 3	37.6	0.0	51.	59.
- 4	44.5	0.0	56.	75.
- 5	66.6	0.0	85.	166.
- 6	87.2	0.0	136.	325.
- 7	38.1	0.05	57.	78.
- 8	38.1	0.10	55.	72.
- 9	37.4	0.15	56.	75.
-10	43.8	0.05	63.	97.
-11	44.3	0.10	65.	103.
-12	44.6	0.15	68.	113.
-13	44.8	0.20	70.	119.
-14	69.1	0.06	92.	188.
-15	69.1	0.11	96.	200.
-16	69.1	0.16	104.	225.
-17	38.7	0.20	60.	88.
-18	38.9	0.0	49.	58.
-19	45.2	0.0	56.	81.
-20	24.3	0.0	36.	16.
-21	67.7	0.0	76.	145.
-22	89.4	0.0	120.	287.

CALCULATED HOLDUPS, RUN 422: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, Bpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (Mm/Sec)
422- 1	13.3	0.0	0.48	0.51	0.69	0.0	0.0
- 2	23.9	0.0	0.43	0.58	0.68	0.0	0.0
- 3	37.6	0.0	0.35	0.68	0.71	0.0	0.0
- 4	44.5	0.0	0.32	0.70	0.78	0.0	0.0
- 5	66.6	0.0	0.21	0.80	0.87	0.0	0.0
- 6	87.2	0.0	0.13	0.88	0.96	0.0	0.0
- 7	38.1	0.05	0.31	0.60	0.50	0.09	10.5
- 8	38.1	0.10	0.32	0.52	0.44	0.15	19.9
- 9	37.4	0.15	0.32	0.53	0.44	0.16	31.1
-10	43.8	0.05	0.28	0.62	0.54	0.10	9.6
-11	44.3	0.10	0.27	0.58	0.50	0.15	20.4
-12	44.6	0.15	0.26	0.59	0.52	0.14	32.7
-13	44.8	0.20	0.25	0.61	0.54	0.14	48.0
-14	69.1	0.06	0.19	0.72	0.62	0.09	9.9
-15	69.1	0.11	0.18	0.65	0.56	0.16	19.0
-16	69.1	0.16	0.17	0.64	0.54	0.19	28.5
-17	38.7	0.20	0.29	0.57	0.47	0.13	47.6
-18	38.9	0.0	0.35	0.67	0.76	0.0	0.0
-19	45.2	0.0	0.31	0.71	0.76	0.0	0.0
-20	24.3	0.0	0.48	0.51	0.79	0.0	0.0
-21	67.7	0.0	0.23	0.79	0.83	0.0	0.0
-22	89.4	0.0	0.14	0.84	0.91	0.0	0.0

CALCULATED HOLDUPS, RUN 422--DILUTE PHASE

CATALYST : MDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EBG
422- 1	13.3	0.0	0.98	1.07	0.0
- 2	23.9	0.0	0.98	1.07	0.0
- 3	37.6	0.0	0.98	1.07	0.0
- 4	44.5	0.0	0.99	1.07	0.0
- 5	66.6	0.0	0.99	1.06	0.0
- 6	87.2	0.0	0.99	1.04	0.0
- 7	38.1	0.05	0.87	0.83	0.13
- 8	38.1	0.10	0.82	0.79	0.18
- 9	37.4	0.15	0.82	0.78	0.18
-10	43.8	0.05	0.86	0.82	0.14
-11	44.3	0.10	0.79	0.78	0.21
-12	44.6	0.15	0.79	0.78	0.21
-13	44.8	0.20	0.79	0.78	0.21
-14	69.1	0.06	0.86	0.83	0.14
-15	69.1	0.11	0.76	0.74	0.24
-16	69.1	0.16	0.71	0.69	0.29
-17	38.7	0.20	0.78	0.78	0.22
-18	38.9	0.0	0.98	1.07	0.0
-19	45.2	0.0	0.99	1.06	0.0
-20	24.3	0.0	0.99	1.07	0.0
-21	67.7	0.0	0.98	1.07	0.0
-22	89.4	0.0	0.98	1.07	0.0

Z BED EXPANSION FOR RUN 423

CATALYST : MDS-2A *l/d=3*
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 175. DEG F

Run No.	Liquid Flow Rate, BPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
423- 1	38.0	0.0	44.	38.
- 2	44.9	0.0	50.	56.
- 3	67.0	0.0	66.	106.
- 4	89.3	0.0	100.	213.
- 5	23.6	0.0	36.	13.
- 6	23.9	0.05	40.	25.
- 7	38.4	0.05	51.	59.
- 8	38.5	0.10	51.	59.
- 9	38.4	0.15	50.	56.
-10	38.4	0.20	52.	63.
-11	45.0	0.05	56.	75.
-12	44.9	0.10	56.	75.
-13	45.1	0.15	55.	72.
-14	45.4	0.20	57.	78.
-15	45.2	0.25	50.	56.
-16	67.3	0.05	72.	125.
-17	67.7	0.11	81.	153.
-18	67.3	0.16	80.	150.
-19	67.8	0.22	77.	141.
-20	67.1	0.26	75.	134.
-21	89.5	0.05	102.	219.
-22	89.0	0.11	114.	256.
-23	38.7	0.23	50.	56.

CALCULATED MOLDUPS, RUN 423: DENSE PHASE

CATALYST : NDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 175. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELCB	ELDPB	EOB	Vcd (Nm/Sec)
423- 1	38.0	0.0	0.40	0.64	0.75	0.0	0.0
- 2	44.9	0.0	0.35	0.69	0.75	0.0	0.0
- 3	67.0	0.0	0.27	0.78	0.78	0.0	0.0
- 4	89.5	0.0	0.18	0.84	0.87	0.0	0.0
- 5	23.6	0.0	0.49	0.52	0.77	0.0	0.0
- 6	23.9	0.05	0.44	0.46	0.30	0.10	10.0
- 7	38.4	0.05	0.35	0.54	0.46	0.11	8.8
- 8	38.5	0.10	0.35	0.49	0.41	0.16	18.3
- 9	38.4	0.15	0.35	0.46	0.39	0.19	27.4
-10	38.4	0.20	0.34	0.47	0.42	0.19	40.4
-11	45.0	0.05	0.32	0.57	0.51	0.12	8.6
-12	44.9	0.10	0.32	0.50	0.46	0.19	16.0
-13	45.1	0.15	0.32	0.48	0.44	0.20	25.9
-14	45.4	0.20	0.31	0.51	0.45	0.18	40.6
-15	45.2	0.25	0.35	0.43	0.38	0.22	46.8
-16	67.3	0.05	0.25	0.65	0.62	0.11	7.3
-17	67.7	0.11	0.22	0.60	0.58	0.18	16.0
-18	67.3	0.16	0.22	0.54	0.51	0.23	22.6
-19	67.8	0.22	0.23	0.53	0.49	0.24	34.4
-20	67.1	0.26	0.24	0.53	0.49	0.23	45.3
-21	89.5	0.05	0.17	0.75	0.69	0.07	9.8
-22	89.0	0.11	0.16	0.69	0.64	0.15	17.6
-23	38.7	0.23	0.35	0.45	0.38	0.19	47.7

CALCULATED HOLDUPS, RUN 423--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 175. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
423- 1	38.0	0.0	0.98	1.08	0.0
- 2	44.9	0.0	0.97	1.08	0.0
- 3	67.0	0.0	0.97	1.07	0.0
- 4	89.5	0.0	0.98	1.07	0.0
- 5	23.6	0.0	0.98	1.07	0.0
- 6	23.9	0.05	0.88	0.85	0.12
- 7	38.4	0.05	0.85	0.82	0.15
- 8	38.5	0.10	0.79	0.77	0.21
- 9	38.4	0.15	0.78	0.76	0.22
-10	38.4	0.20	0.76	0.74	0.24
-11	45.0	0.05	0.85	0.81	0.15
-12	44.9	0.10	0.78	0.74	0.22
-13	45.1	0.15	0.76	0.74	0.24
-14	45.4	0.20	0.75	0.74	0.25
-15	45.2	0.25	0.73	0.73	0.27
-16	67.3	0.05	0.85	0.82	0.15
-17	67.7	0.11	0.75	0.73	0.25
-18	67.3	0.16	0.70	0.66	0.30
-19	67.8	0.22	0.71	0.69	0.29
-20	67.1	0.26	0.71	0.71	0.29
-21	89.5	0.05	0.88	0.85	0.12
-22	89.0	0.11	0.79	0.76	0.21
-23	38.7	0.23	0.74	0.74	0.26

Z BED EXPANSION FOR RUN 424

CATALYST : MDS-2A *l/d = 3*
 GAS : HELIUM
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 177. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
424- 1	38.2	0.20	51.	65.
- 2	38.3	0.25	52.	68.
- 3	45.0	0.25	56.	81.
- 4	45.0	0.20	57.	84.
- 5	66.9	0.20	79.	155.
- 6	67.4	0.25	81.	161.
- 7	87.8	0.25	116.	274.
- 8	88.1	0.20	117.	277.
- 9	38.4	0.05	49.	58.
-10	38.6	0.10	50.	61.
-11	38.0	0.15	51.	65.
-12	45.1	0.05	53.	71.
-13	45.2	0.10	55.	77.
-14	44.7	0.15	56.	81.
-15	66.4	0.05	70.	126.
-16	67.6	0.10	75.	142.
-17	66.9	0.15	78.	152.
-18	89.7	0.05	101.	226.
-19	89.4	0.10	108.	248.
-20	89.7	0.15	116.	274.

CALCULATED HOLDUPS, RUN 424: DENSE PHASE

CATALYST : NDS-2A
GAS : HELIUM
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 177. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELCB	ELDPB	EOB	Vcd (Hr/Sec)
424- 1	38.2	0.20	0.34	0.52	0.43	0.14	46.3
- 2	38.3	0.25	0.33	0.51	0.43	0.16	57.4
- 3	45.0	0.25	0.31	0.53	0.45	0.17	55.6
- 4	45.0	0.20	0.30	0.54	0.47	0.16	43.2
- 5	66.9	0.20	0.22	0.60	0.55	0.18	38.4
- 6	67.4	0.25	0.21	0.58	0.53	0.21	47.6
- 7	87.8	0.25	0.15	0.65	0.60	0.20	46.7
- 8	88.1	0.20	0.15	0.67	0.61	0.18	36.5
- 9	38.4	0.05	0.35	0.57	0.46	0.08	10.8
-10	38.6	0.10	0.34	0.54	0.43	0.11	22.1
-11	38.0	0.15	0.34	0.53	0.43	0.13	34.1
-12	45.1	0.05	0.32	0.61	0.52	0.07	10.9
-13	45.2	0.10	0.31	0.57	0.49	0.12	21.4
-14	44.7	0.15	0.31	0.54	0.47	0.15	31.6
-15	66.4	0.05	0.24	0.69	0.63	0.07	10.0
-16	67.6	0.10	0.23	0.65	0.60	0.12	19.3
-17	66.9	0.15	0.22	0.61	0.56	0.17	27.6
-18	89.7	0.05	0.17	0.76	0.71	0.07	9.1
-19	89.4	0.10	0.16	0.72	0.67	0.12	18.0
-20	89.7	0.15	0.15	0.67	0.63	0.18	24.3

CALCULATED HOLDUPS, RUN 424--DILUTE PHASE

CATALYST : HDS-2A
GAS : HELIUM
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 177. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
424- 1	38.2	0.20	0.80	0.78	0.20
- 2	38.3	0.25	0.81	0.78	0.19
- 3	45.0	0.25	0.78	0.76	0.22
- 4	45.0	0.20	0.79	0.76	0.21
- 5	66.9	0.20	0.76	0.72	0.24
- 6	67.4	0.25	0.73	0.70	0.27
- 7	87.8	0.25	0.73	0.70	0.27
- 8	88.1	0.20	0.75	0.72	0.25
- 9	38.4	0.05	0.90	0.86	0.10
-10	38.6	0.10	0.84	0.80	0.16
-11	38.0	0.15	0.83	0.79	0.17
-12	45.1	0.05	0.91	0.86	0.09
-13	45.2	0.10	0.84	0.80	0.16
-14	44.7	0.15	0.81	0.77	0.19
-15	66.4	0.05	0.91	0.87	0.09
-16	67.6	0.10	0.83	0.80	0.17
-17	66.9	0.15	0.78	0.74	0.22
-18	89.7	0.05	0.91	0.86	0.09
-19	89.4	0.10	0.83	0.79	0.17
-20	89.7	0.15	0.77	0.74	0.23

Z BED EXPANSION FOR RUN 425

CATALYST : MDS-2A *l/d = 3*
GAS : HELIUM
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
425- 1	38.4	0.05	54.	74.
- 2	38.4	0.10	55.	77.
- 3	39.0	0.15	56.	81.
- 4	45.4	0.05	57.	84.
- 5	45.4	0.10	63.	103.
- 6	44.6	0.15	61.	97.
- 7	67.3	0.05	81.	161.
- 8	66.8	0.10	91.	194.
- 9	67.4	0.15	91.	194.
-10	89.4	0.05	125.	303.
-11	89.5	0.10	129.	316.
-12	89.0	0.15	135.	336.
-13	39.7	0.20	51.	65.
-14	39.3	0.25	53.	71.
-15	45.9	0.25	53.	71.
-16	45.2	0.20	60.	94.
-17	67.7	0.20	91.	194.
-18	67.0	0.25	81.	161.

CALCULATED HOLDUPS, RUN 425: DENSE PHASE

CATALYST : MDS-2A
GAS : HELIUM
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Nm/Sec)
425- 1	38.4	0.05	0.32	0.61	0.52	0.07	11.3
- 2	38.4	0.10	0.31	0.59	0.50	0.10	23.4
- 3	39.0	0.15	0.31	0.58	0.50	0.12	35.8
- 4	45.4	0.05	0.30	0.63	0.55	0.07	10.7
- 5	45.4	0.10	0.27	0.62	0.56	0.11	22.5
- 6	44.6	0.15	0.28	0.59	0.53	0.13	33.8
- 7	67.3	0.05	0.21	0.71	0.65	0.08	9.6
- 8	66.8	0.10	0.19	0.68	0.63	0.13	19.0
- 9	67.4	0.15	0.19	0.67	0.61	0.14	30.9
-10	89.4	0.05	0.14	0.82	0.75	0.04	11.4
-11	89.5	0.10	0.13	0.76	0.69	0.10	20.1
-12	89.0	0.15	0.13	0.73	0.66	0.14	28.9
-13	39.7	0.20	0.34	0.54	0.45	0.12	48.2
-14	39.3	0.25	0.32	0.55	0.47	0.13	60.7
-15	45.9	0.25	0.32	0.53	0.45	0.14	58.1
-16	45.2	0.20	0.29	0.58	0.52	0.14	46.3
-17	67.7	0.20	0.19	0.66	0.64	0.16	42.2
-18	67.0	0.25	0.21	0.61	0.56	0.18	51.9

CALCULATED HOLDUPS, RUN 425--DILUTE PHASE

CATALYST : MDS-2A
GAS : HELIUM
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 150. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	EL8	ELDP	EO8
425- 1	38.4	0.05	0.91	0.87	0.09
- 2	38.4	0.10	0.86	0.83	0.14
- 3	39.0	0.15	0.85	0.83	0.15
- 4	45.4	0.05	0.90	0.86	0.10
- 5	45.4	0.10	0.84	0.81	0.16
- 6	44.6	0.15	0.84	0.81	0.16
- 7	67.3	0.05	0.89	0.85	0.11
- 8	66.8	0.10	0.81	0.78	0.19
- 9	67.4	0.15	0.80	0.76	0.20
-10	89.4	0.05	0.93	0.89	0.07
-11	89.5	0.10	0.86	0.84	0.14
-12	89.0	0.15	0.80	0.79	0.20
-13	39.7	0.20	0.84	0.81	0.16
-14	39.3	0.25	0.83	0.81	0.17
-15	45.9	0.25	0.83	0.80	0.17
-16	45.2	0.20	0.83	0.80	0.17
-17	67.7	0.20	0.78	0.74	0.22
-18	67.0	0.25	0.76	0.73	0.24

Z BED EXPANSION FOR RUN 426

CATALYST : MDS-2A *L/d=3*
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 154. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
426- 1	13.7	0.0	32.	3.
- 2	20.9	0.0	36.	16.
- 3	26.6	0.0	41.	32.
- 4	38.3	0.0	49.	63.
- 5	45.0	0.0	56.	87.
- 6	67.2	0.0	76.	153.
- 7	87.6	0.0	111.	270.
- 8	39.0	0.05	54.	80.
- 9	38.7	0.10	56.	87.
-10	38.7	0.15	54.	80.
-11	39.0	0.20	53.	77.
-12	44.8	0.05	60.	100.
-13	45.4	0.10	62.	107.
-14	44.7	0.15	61.	103.
-15	44.9	0.20	60.	100.
-16	67.4	0.05	82.	173.
-17	67.6	0.11	91.	203.
-18	67.4	0.16	92.	207.
-19	67.2	0.22	82.	173.

CALCULATED HOLDUPS, RUN 426: DENSE PHASE

CATALYST : HYS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 154. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELG2	ELDPB	EOB	Vcd (Nm/Sec)
426- 1	13.7	0.0	0.54	0.51	0.77	0.0	0.0
- 2	20.9	0.0	0.48	0.51	0.76	0.0	0.0
- 3	26.6	0.0	0.42	0.59	0.75	0.0	0.0
- 4	38.3	0.0	0.34	0.65	0.83	0.0	0.0
- 5	45.0	0.0	0.30	0.70	0.84	0.0	0.0
- 6	67.2	0.0	0.22	0.79	0.89	0.0	0.0
- 7	87.6	0.0	0.15	0.83	0.97	0.0	0.0
- 8	39.0	0.05	0.31	0.61	0.47	0.08	10.0
- 9	38.7	0.10	0.30	0.58	0.45	0.12	21.9
-10	38.7	0.15	0.31	0.57	0.44	0.12	34.0
-11	39.0	0.20	0.31	0.54	0.42	0.15	45.1
-12	44.8	0.05	0.28	0.63	0.52	0.09	9.7
-13	45.4	0.10	0.27	0.60	0.51	0.13	20.5
-14	44.7	0.15	0.27	0.59	0.50	0.14	32.0
-15	44.9	0.20	0.28	0.59	0.50	0.14	46.2
-16	67.4	0.05	0.20	0.71	0.63	0.09	9.2
-17	67.6	0.11	0.18	0.67	0.59	0.15	19.6
-18	67.4	0.16	0.18	0.62	0.55	0.20	27.1
-19	67.2	0.22	0.20	0.59	0.52	0.21	39.5

CALCULATED HOLDUPS, RUN 426--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 154. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EOG
426- 1	13.7	0.0	0.98	1.04	0.0
- 2	20.9	0.0	0.98	1.04	0.0
- 3	26.6	0.0	0.99	1.04	0.0
- 4	38.3	0.0	0.98	1.06	0.0
- 5	45.0	0.0	0.98	1.06	0.0
- 6	67.2	0.0	0.98	1.06	0.0
- 7	87.6	0.0	0.99	1.06	0.0
- 8	39.0	0.05	0.87	0.84	0.13
- 9	38.7	0.10	0.82	0.80	0.18
-10	38.7	0.15	0.81	0.79	0.19
-11	39.0	0.20	0.80	0.79	0.20
-12	44.8	0.05	0.86	0.83	0.14
-13	45.4	0.10	0.81	0.77	0.19
-14	44.7	0.15	0.81	0.78	0.19
-15	44.9	0.20	0.80	0.77	0.20
-16	67.4	0.05	0.87	0.83	0.13
-17	67.6	0.11	0.78	0.74	0.22
-18	67.4	0.16	0.73	0.70	0.27
-19	67.2	0.22	0.75	0.72	0.25

Z BED EXPANSION FOR RUN 427

CATALYST : NDS-2A *l/d=3*
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 99. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
427- 1	12.8	0.0	36.	20.
- 2	22.4	0.0	47.	57.
- 3	30.6	0.0	61.	103.
- 4	41.4	0.0	76.	153.
- 5	37.3	0.0	71.	137.
- 6	56.5	0.0	115.	283.
- 7	23.7	0.0	47.	57.
- 8	37.1	0.0	70.	133.
- 9	56.1	0.0	113.	277.

CALCULATED HOLDUPS, RUN 427: DENSE PHASE

M80-21
-201

CATALYST : NDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 99. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELCB	ELDPB	EOB	Vcd (Mm/Sec)
427- 1	12.8	0.0	0.46	0.50	0.84	0.0	0.0
- 2	22.4	0.0	0.35	0.59	0.82	0.0	0.0
- 3	30.6	0.0	0.27	0.70	0.82	0.0	0.0
- 4	41.4	0.0	0.22	0.76	0.86	0.0	0.0
- 5	37.3	0.0	0.23	0.73	0.85	0.0	0.0
- 6	56.5	0.0	0.14	0.82	0.92	0.0	0.0
- 7	23.7	0.0	0.35	0.61	0.74	0.0	0.0
- 8	37.1	0.0	0.24	0.74	0.76	0.0	0.0
- 9	56.1	0.0	0.15	0.81	0.84	0.0	0.0

CALCULATED HOLDUPS, RUN 427--DILUTE PHASE

CATALYST : NDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 99. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELC	ELDP	EOB
427- 1	12.8	0.0	0.99	1.04	0.0
- 2	22.4	0.0	1.00	1.04	0.0
- 3	30.6	0.0	0.99	1.04	0.0
- 4	41.4	0.0	0.99	1.04	0.0
- 5	37.3	0.0	0.99	1.03	0.0
- 6	56.5	0.0	0.99	1.03	0.0
- 7	23.7	0.0	0.99	1.04	0.0
- 8	37.1	0.0	1.00	1.04	0.0
- 9	56.1	0.0	0.99	1.04	0.0

Z BED EXPANSION FOR RUN 428

M80-21
-202

CATALYST : HDS-2A *l/d = 3*
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 127. DEG F

Run No.	Liquid Flow Rate, GPM/Ft ²	Gas Flow Rate, Ft/Sec	Catalyst Bed Height (In.)	Z Bed Expansion
428- 1	10.2	0.0	32.	7.
- 2	22.9	0.0	40.	33.
- 3	35.3	0.0	52.	73.
- 4	44.6	0.0	63.	110.
- 5	72.3	0.0	106.	253.
- 6	13.1	0.05	33.	10.

CALCULATED HOLDUPS, RUN 428: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 127. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EOB	Vcd (Hm/Sec)
428- 1	10.2	0.0	0.52	0.49	0.86	0.0	0.0
- 2	22.9	0.0	0.41	0.54	0.79	0.0	0.0
- 3	35.3	0.0	0.32	0.68	0.77	0.0	0.0
- 4	44.6	0.0	0.26	0.72	0.78	0.0	0.0
- 5	72.3	0.0	0.16	0.83	0.87	0.0	0.0
- 6	13.1	0.05	0.50	0.44	0.20	0.05	12.8

CALCULATED HOLDUPS, RUN 428--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 127. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELB	ELDP	EOB
428- 1	10.2	0.0	0.99	1.05	0.0
- 2	22.9	0.0	0.98	1.05	0.0
- 3	35.3	0.0	0.99	1.05	0.0
- 4	44.6	0.0	0.99	1.05	0.0
- 5	72.3	0.0	0.99	1.05	0.0
- 6	13.1	0.05	0.93	0.89	0.07

X BED EXPANSION FOR RUN 429

M80-21
-203

CATALYST : HDS-2A *2/d=3*
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 176. DEG F

Run No.	Liquid Flow Rate, GPM/Ft2	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
429- 1	45.7	0.0	50.	67.
- 2	45.8	0.06	55.	83.
- 3	44.9	0.10	57.	88.
- 4	45.0	0.15	55.	83.
- 5	45.1	0.20	54.	80.
- 6	44.0	0.26	54.	80.
- 7	67.8	0.0	69.	130.
- 8	66.9	0.05	72.	140.
- 9	68.1	0.10	81.	170.
-10	67.4	0.15	86.	187.
-11	67.3	0.22	84.	180.
-12	67.5	0.25	78.	160.

CALCULATED HOLDUPS, RUN 429: DENSE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 176. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Nm/Sec)
429- 1	45.7	0.0	0.33	0.67	0.83	0.0	0.0
- 2	45.8	0.06	0.30	0.60	0.49	0.10	12.9
- 3	44.9	0.10	0.29	0.54	0.45	0.17	18.0
- 4	45.0	0.15	0.30	0.51	0.43	0.19	27.9
- 5	45.1	0.20	0.31	0.51	0.42	0.18	41.3
- 6	44.0	0.26	0.31	0.50	0.42	0.19	55.2
- 7	67.8	0.0	0.24	0.77	0.85	0.0	0.0
- 8	66.9	0.05	0.23	0.67	0.61	0.10	7.9
- 9	68.1	0.10	0.20	0.64	0.56	0.16	15.9
-10	67.4	0.15	0.19	0.59	0.52	0.21	24.2
-11	67.3	0.22	0.20	0.55	0.49	0.25	34.0
-12	67.5	0.25	0.21	0.54	0.49	0.24	41.6

CALCULATED HOLDUPS, RUN 429--DILUTE PHASE

CATALYST : HDS-2A
GAS : NITROGEN
LIQUID : MIN-OIL
COAL CHAR CONC: 0.0 VOL %
TEMPERATURE : 176. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
429- 1	45.7	0.0	0.98	1.07	0.0
- 2	45.8	0.06	0.84	0.82	0.16
- 3	44.9	0.10	0.76	0.73	0.24
- 4	45.0	0.15	0.76	0.74	0.24
- 5	45.1	0.20	0.75	0.74	0.25
- 6	44.0	0.26	0.73	0.72	0.27
- 7	67.8	0.0	0.98	1.07	0.0
- 8	66.9	0.05	0.87	0.84	0.13
- 9	68.1	0.10	0.77	0.74	0.23
-10	67.4	0.15	0.71	0.68	0.29
-11	67.3	0.22	0.69	0.67	0.31
-12	67.5	0.25	0.71	0.69	0.29

CALCULATED HOLDUPS, RUN 430: DENSE PHASE

CATALYST : NONE *1/d: 6*
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 170. DEG F

M80-21
 -205

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Ma/Sec)
430- 1	45.3	0.10	0.0	0.80	0.74	0.20	18.5
- 2	38.3	0.10	0.0	0.82	0.76	0.18	20.1
- 3	66.5	0.10	0.0	0.80	0.72	0.20	15.3
- 4	66.9	0.05	0.0	0.89	0.81	0.11	8.6
- 5	45.4	0.05	0.0	0.88	0.81	0.12	9.6
- 6	38.6	0.05	0.0	0.88	0.82	0.12	10.3
- 7	38.3	0.15	0.0	0.79	0.73	0.21	30.8
- 8	45.4	0.15	0.0	0.79	0.72	0.21	29.4
- 9	66.6	0.16	0.0	0.74	0.67	0.26	23.8
-10	38.5	0.19	0.0	0.77	0.72	0.23	39.0
-11	45.0	0.20	0.0	0.76	0.71	0.24	37.9
-12	66.0	0.22	0.0	0.72	0.66	0.28	34.7
-13	66.5	0.25	0.0	0.71	0.65	0.29	40.4
-14	45.3	0.23	0.0	0.74	0.70	0.26	42.8
-15	38.5	0.23	0.0	0.75	0.71	0.25	46.7
-16	88.0	0.06	0.0	0.91	0.81	0.09	9.9
-17	88.0	0.12	0.0	0.81	0.73	0.19	17.7

CALCULATED HOLDUPS, RUN 430--DILUTE PHASE

CATALYST : NONE
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 170. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft ²	Gas Flow Rate, Ft/Sec	ELB	ELDP	EGG
430- 1	45.3	0.10	0.77	0.75	0.23
- 2	38.3	0.10	0.80	0.77	0.20
- 3	66.5	0.10	0.77	0.74	0.23
- 4	66.9	0.05	0.87	0.84	0.13
- 5	45.4	0.05	0.86	0.83	0.14
- 6	38.6	0.05	0.86	0.84	0.14
- 7	38.3	0.15	0.76	0.74	0.24
- 8	45.4	0.15	0.75	0.73	0.25
- 9	66.6	0.16	0.72	0.69	0.28
-10	38.5	0.19	0.75	0.73	0.25
-11	45.0	0.20	0.74	0.73	0.26
-12	66.0	0.22	0.69	0.68	0.31
-13	66.5	0.25	0.70	0.68	0.30
-14	45.3	0.23	0.73	0.71	0.27
-15	38.5	0.23	0.73	0.72	0.27
-16	88.0	0.06	0.89	0.86	0.11
-17	88.0	0.12	0.81	0.77	0.19

% BED EXPANSION FOR RUN 431

CATALYST : NONE */d = C*
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 149. DEG F

Run No.	Liquid Flow Rate, BPM/Ft2	Gas Flow Rate Ft/Sec	Catalyst Bed Height (In.)	% Bed Expansion
431- 1	38.9	0.05	0.	0.
- 2	39.0	0.10	0.	0.
- 3	38.9	0.14	0.	0.
- 4	39.2	0.20	0.	0.
- 5	45.3	0.05	0.	0.
- 6	45.0	0.10	0.	0.
- 7	45.3	0.15	0.	0.
- 8	45.7	0.20	0.	0.
- 9	45.0	0.24	0.	0.
-10	38.5	0.24	0.	0.
-11	67.5	0.05	0.	0.
-12	67.8	0.11	0.	0.
-13	67.6	0.16	0.	0.
-14	66.1	0.22	0.	0.
-15	69.3	0.25	0.	0.
-16	90.0	0.06	0.	0.

CALCULATED HOLDUPS, RUN 431: DENSE PHASE

CATALYST : NONE
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 149. DEG F

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 -207

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ECB	ELGB	ELDPB	EGB	Vcd (Nm/Sec)
431- 1	38.9	0.05	0.0	0.89	0.86	0.11	9.7
- 2	39.0	0.10	0.0	0.85	0.81	0.15	20.8
- 3	38.9	0.14	0.0	0.82	0.80	0.18	31.2
- 4	39.2	0.20	0.0	0.79	0.78	0.21	41.5
- 5	45.3	0.05	0.0	0.89	0.86	0.11	10.2
- 6	45.0	0.10	0.0	0.82	0.80	0.18	18.1
- 7	45.3	0.15	0.0	0.80	0.79	0.20	30.4
- 8	45.7	0.20	0.0	0.78	0.77	0.22	41.6
- 9	45.0	0.24	0.0	0.77	0.76	0.23	48.6
-10	38.5	0.24	0.0	0.78	0.77	0.22	51.2
-11	67.5	0.05	0.0	0.89	0.86	0.11	9.7
-12	67.8	0.11	0.0	0.81	0.78	0.19	18.5
-13	67.6	0.16	0.0	0.77	0.74	0.23	27.3
-14	66.1	0.22	0.0	0.75	0.73	0.25	39.3
-15	69.3	0.25	0.0	0.74	0.72	0.26	44.1
-16	90.0	0.06	0.0	0.91	0.87	0.09	9.9

CALCULATED HOLDUPS, RUN 431--DILUTE PHASE

CATALYST : NONE
 GAS : NITROGEN
 LIQUID : MIN-OIL
 COAL CHAR CONC: 0.0 VOL %
 TEMPERATURE : 149. DEG F

Run No.	Liquid Flow Rate, Gpm/Ft2	Gas Flow Rate, Ft/Sec	ELG	ELDP	EGG
431- 1	38.9	0.05	0.88	0.89	0.12
- 2	39.0	0.10	0.82	0.84	0.18
- 3	38.9	0.14	0.79	0.83	0.21
- 4	39.2	0.20	0.78	0.81	0.22
- 5	45.3	0.05	0.86	0.87	0.14
- 6	45.0	0.10	0.79	0.81	0.21
- 7	45.3	0.15	0.78	0.79	0.22
- 8	45.7	0.20	0.77	0.79	0.23
- 9	45.0	0.24	0.77	0.77	0.23
-10	38.5	0.24	0.77	0.76	0.23
-11	67.5	0.05	0.87	0.84	0.13
-12	67.8	0.11	0.78	0.75	0.22
-13	67.6	0.16	0.75	0.72	0.25
-14	66.1	0.22	0.74	0.72	0.26
-15	69.3	0.25	0.73	0.73	0.27
-16	90.0	0.06	0.90	0.87	0.10

Figure B.1

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Effect of Operating conditions on Catalyst holdup

Figure B.2

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Effect of Temperature on Catalyst holdup

Figure B.3

Effect of Viscosity on Catalyst holdup

Figure B.4

Effect of Particle size on Catalyst holdup

Figure B.5

Effect of Gas type on Catalyst holdup

Figure B.6

Effect of Particle size on Gas holdup

Figure B.7

Effect of Temperature on Gas holdup

APPENDIX C

TRACER DATA ANALYSIS

APPENDIX C

TRACER DATA ANALYSIS

Tracer Data Analysis

In analyzing tracer data, a minimum of two parameters is important for describing the residence time distribution of a pulse injected in a flow system (23,41): the first moment (μ) and the second moment (σ^2). The first moment is related to the mean residence time (τ) monitored between two reactor locations by the following equation:

$$\tau = \mu_2 - \mu_1 \quad (C-1)$$

The first and second moments are determined from the following equations:

$$\mu = \frac{\int_0^{\infty} t c(t) dt}{\int_0^{\infty} c(t) dt} \quad (C-2)$$

$$\sigma^2 = \frac{\int_0^{\infty} t^2 c(t) dt}{\int_0^{\infty} c(t) dt} - \mu^2 \quad (C-3)$$

The integrals were evaluated on the computer numerically using Simpson's Rule. Error due to this numerical procedure should be minimal because of the very small (less than 0.2 sec) interval size used. However, calculation of moments using Equations C-2 and C-3 can result in significant errors due to error in concentration values at long times (41). This error can be minimized by using a modified method of analysis of moments (41). This method utilizes a Laplace transform of a concentration distribution:

$$c(s) = \int_0^{\infty} c(t) \exp(-st) dt \quad (C-4)$$

Differentiating with respect to s gives:

$$\frac{d^n}{ds^n} c(s) = -1^n \int_0^{\infty} t^n c(t) \exp(-st) dt \quad (C-5)$$

The modified first moment is found by calculating the slope of a plot of $c(s)$ versus s as s approaches zero. The second moment is determined in a similar fashion.

Both methods were used to calculate the moments of the gas tracer concentration/time curves. The difference in the moments evaluated by the two different methods is an indication of the error in the concentration distribution at long times.

The gas linear velocity (V_g) between Locations 1 and 2 can be determined from the first moments by the following equation:

$$V_g = \frac{h}{\mu_2 - \mu_1} \quad (C-6)$$

where h is the distance between Locations 1 and 2.

In addition, both first and second moments can be used to estimate parameters of a mixing model. Mixing of the gas in the system studied is characterized by the Peclet number:

$$\frac{1}{Pe_g} = \frac{E_g}{V_g h} \quad (C-7)$$

where E_g is the dispersion coefficient.

A simple relationship between moments and the dispersion coefficient was given by Aris (43) as:

$$\frac{\Delta \sigma^2}{(\Delta \mu)^2} = \frac{2E_g}{V_g h} \quad (C-8)$$

Although the results reported in this work are not sufficient to extend the work of other investigators (44) to three-phase fluidized systems, it may be worthwhile to estimate the order of magnitude dispersion coefficients for the systems studied in this work.

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APPENDIX D

VISCOSITY MEASUREMENTS OF H-COAL LIQUIDS

VISCOSITY MEASUREMENTS OF H-COAL LIQUIDS

In this appendix, additional details are reported regarding the PDU sampling technique and the viscometer used by Battelle.

PDU Sampling Technique

The samples were obtained from the PDU using the sampling system shown in Figure D-1. The main two components of the system are a slop vessel and a sample vessel. The sampling procedure was quite simple. First the unit was connected to the PDU sampling point. The system was then purged with nitrogen and the system tubing and vessels were heated to approximately 300°F. After the purge was complete, the system was sealed to the atmosphere and the valves on the PDU were opened, allowing H-Coal liquids to flow into the slop bomb. After the slop bomb was filled, the valves leading to the sample vessel were opened and a sample was taken. After the sample was taken, the sample vessel was isolated and the valves connecting the system to the PDU were closed. The slop vessel and connecting lines were vented and purged with nitrogen. The sample vessel was removed from the unit and allowed to cool. After it had cooled, Valves A and B between the sample vessel and the outage vessel were opened to provide a void volume for shipment to Battelle. Both Battelle and HRI took precautions to see that the samples were not exposed to air.

Viscosity Measurement.--The viscometer used by Battelle is shown in Figure D-2. The viscometer consisted of:

- 1) Stainless steel body of a 300 cc Autoclave.
- 2) Hollow stainless steel bob of a diameter 1 mm less than the inside diameter of the Autoclave.
- 3) Stainless steel tube which supports the bob and serves to keep the pressure inside the bob the same as the Autoclave pressure.
- 4) Connector.
- 5) Alnico magnet.
- 6) Two coils.
- 7) Furnace supplied with the Autoclave.
- 8) Additional furnace for better temperature control.
- 9) Sheathed 1/16" thermocouple.
- 10) Three of the gas inlet-outlet connections.

Viscosity measurements are made by passing a current through one of the coils, thus exerting a force on the bob, moving it either up or down. As the bob moves through the sample, the magnet moves through the other coil, thus generating an emf in the coil. This emf can be related to the velocity of the bob. Thus, from the current applied in one coil and the emf generated in the other coil, the relationship between shear stress and shear rate of the fluid in the viscometer can be determined. A more complete description of the viscometer, the sample loading technique, and its operation is contained in Reference 51.

The measured values of viscosity η' and θ for the H-Coal samples (Amoco Nos. 1, 2, 4) and the char kerosene samples are listed in Table D-I, which has been reproduced from Battelle's report. It should be noted that the kerosene/char samples sent to Battelle were incorrectly identified; the actual coal concentrations were 5.1, 10.4, and 17.8 vol%. Inspection of the results indicates that the viscosity and yield stress of the char kerosene samples at ambient conditions are in the same range as those of the H-Coal samples at H-Coal reactor conditions (1 to 10 cp and 1000 to 2000 millinewtonians/m², respectively). There is considerable scatter in the yield stress data. A discussion of the scatter can be found in Battelle's report (51).

TABLE D-I

MEASURED VISCOUS PROPERTIES

Sample	Temperature		Pressure, psia	Plastic Viscosity, η' centipoises	Yield Stress, θ mN/m ²	Note
	C	F				
AMOCO #1	171	340	1982	48		
	140	284	1900	151		1
	292	558	2015	3.6		
	374	705	2012	2.1		2
	372	702	2030	2.1		3
	367	693	2500	2.2		
	403	757	2000	1.5		1,2
	404	759	2006	1.6		3
	402	756	2478	1.6		
	450	842	1990	1.2		
	453	847	2495	1.2		
	444	831	2990	1.2		
AMOCO #2	374		2010	3.5	640	2
	371		2000	3.4	900	3
	373		2490	3.2	930	
	373		3002	3.3	780	
	398	748	2010	2.3	340	
	397	747	2500	2.2	220	
	395	743	3003	2.5	<0	
	399	750	2020	2.0	700	1,2
	404	759	2008	1.9	390	3
	437	819	2000	1.5	1820	
	440	824	2500	1.5	1220	
	442	828	2995	1.6	1370	
	440	824	2000	1.5	1050	

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TABLE D-1

(Continued)

Sample	Temperature		Pressure, psia	Plastic Viscosity, η' centipoises	Yield Stress, θ mN/m ²	Note
	C	F				
AMOCO #2 (Cont'd)	450	842	2010	2.0	1500	
	452	846	-2500	1.8	1950	
	460	860	-3000	1.9	3140	
AMOCO #4	376	709	2515	3.5	770	
	366	691	2510	3.5	640	
	383	721	2518	2.9	510	
	413	775	2500	2.3	250	
	434	813	2510	1.7	430	
	461	862	2530	1.3	1190	
	352	666	2509	2.9	890	
Kerosene	30.2	86.4		1.38	50	
	44.8	112.6		1.05	80	
	65.7	150.3		.88	20	
5% Char	22.9	73.2		3.8	560	
	35.2	95.4		3.1	470	1
	61.3	142.3		2.2	340	
10% Char	22.3	72.1		4.2	990	
	37.5	99.5		3.2	820	1
	67.4	153.3		2.2	780	
16-1/2% Char	23.3	73.9		7.2	1100	
	45.1	113.2		3.6	1400	
	62.3	144.1		3.3	2000	

Notes: 1 -- Next day.
 2 -- Before mixing by passing hydrogen through slurry.
 3 -- After such mixing.

Figure D-1

Sampling system at HRI PDU

This assembly to be shipped
to battle for viscosity
determination

Notes:

(1) Piping spec's:

- (a) Tubing -3/8" O.D. x 0.065" wall 316 S.S.
- (b) Valves to be 3/8" Autoclave engineers 10V5071 w/graphite asbestos packing
- (c) Fittings to be swagelok 316 S.S.
- (d) To be furnished by HRI. all other material to be furnished by Amoco
- (e) Valves A & B to be opened only after sample has cooled

Figure D.2
Battelle Viscometer

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APPENDIX E

PREDICTIVE COMPUTER PROGRAM

APPENDIX E

PREDICTIVE COMPUTER PROGRAM

This appendix documents the Fortran program logic used to predict ebullated bed phase holdups from operating conditions.

Figures E-1 through E-4 are simplified flow charts of major subroutines whose purpose is not immediately apparent from the coding. Table E-I lists all subroutines and their functions.

The Fortran coding and a list of vector locations appear at the end of this appendix. Because this program shares several subroutines with the program for experimental data analysis, the vector list will not be continuous. Similarly, vectors such as EC, ECG, and ELG may be deleted if desired.

The Peebles and Garber correlation is presented in Table E-II to aid in tracing the logic of Subroutine PEEBLE.

TABLE E-I

PROGRAM SUBROUTINE FUNCTIONS

MAIN Program	Controls iteration sequence.
BHATIA	Calculates phase holdups from velocities.
CHGCD	Overlays vectors after normal cards read in.
DRIFT	Calculates V_{CD} , drift flux.
ELTAM	Calculates X_k .
FLWS	Converts ft/sec to volume units.
LINEAR	Calculates linear velocities of gas and liquid based on bed volume fractions.
PEEBLE	Calculates bubble diameter from terminal bubble velocity.
READ	Reads data cards.
UTB	Calculates bubble terminal velocity.
WAKE	Calculates ratio of wake to bubble holdup.

TABLE E-II

TERMINAL VELOCITY OF ISOLATED BUBBLES IN LIQUIDS

<u>Region</u>	<u>Terminal Velocity, U_t</u>	<u>Range of Applicability</u>
1	$\frac{2r_e^2(\rho_l - \rho_g)g}{9\mu_l}$	$Re_b < 2$ where Re_b bubble Reynolds number = $\frac{2\rho_l U_t r_e}{\mu_l}$
2	$0.33g^{0.73}\left(\frac{\sigma}{\mu_l}\right)^{0.52}r_e^{1.23}$	$2 < Re_b < 4.02M^{-0.214}$ where M Morton Number = $\frac{g\mu_l^4}{\rho_l\sigma^3}$
3	$1.35\left(\frac{\sigma}{\rho_l r_e}\right)^{1/2}$	$4.02M^{-0.214} < Re_b < 3.10M^{-1/4}$ or $16.32M^{0.144} < G < 5.75$ where $G = \frac{8r_e^4 U_t^4 \rho_l^3}{\sigma^3}$
4	$1.18\left(\frac{g\sigma}{\rho_l}\right)^{0.25}$	$3.10 G < Re_b$ $5.75 < G$

Source of Information: Peebles, F. N., and Garber, H. J., "Studies on the Motion of Gas Bubbles in Liquids," Chemical Engineering Progress, 49, 2 (1953).

REQUESTED OPTIONS: NOTERM

OPTIONS IN EFFECT: NAME(MAIN) NOOPTIMIZE LINECOUNT(60) SIZE(MAX) AUTODUM(MONE)

SOURCE FRCOIC NOLIST NUNECK OBJECT MAP NOFORMAT GUSTMI NOXREF NOALC NOANSF NOTERM IBM FLAG(I) XL

```

ISN 0002      COMMON EC(20),EGP(10),EFP(10),ELP(10),ESG(100),EFG(100),ELG(100)
ISN 0003      COMMON V(100),Z(100),E(100),F(100),C(100),GR(100),ZGR(100)
ISN 0004      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT
ISN 0005      KCOUNT=0
ISN 0006      KS(44) = 0
ISN 0007      C      CONSTANTS
ISN 0007      C(6) = 32.12
ISN 0008      C(7) = 32.12
ISN 0009      C      DEFAULT VALUES FOR XK OPTION
ISN 0009      DO 5 L = 47,50
ISN 0010      S KS (L) = -99
ISN 0011      KS(46) = 0
ISN 0012      C      WRITE HEADER FOR OUTPUT TABLE
ISN 0012      WRITE(6,1918)
ISN 0013      C      READ UNIT DATA
ISN 0013      10 CALL READ
ISN 0014      C      SET DEFAULT, MAX NUMBER OF ITERATIONS TO 20
ISN 0014      IF (KS(43), LE, 0) KS(43) = 20
ISN 0016      C      IF END OF DATA, QUIT
ISN 0016      IF (KS(44), NE, 0) GO TO 9999
ISN 0018      C      ZERO COUNTER FOR NUMBER OF ITERATIONS
ISN 0018      KS(42) = 0
ISN 0019      C
ISN 0019      C      CONVERT FLOWS TO ENGLISH UNITS
ISN 0019      C
ISN 0019      C      CALL FLOWS
ISN 0020      C
ISN 0020      C      INITIAL GUESSES FOR CAT, SLURRY, GAS HOLDUPS
ISN 0020      F(1) = 0.1
ISN 0021      F(9) = 0.7
ISN 0022      F(2) = 0.2
ISN 0023      C
ISN 0023      C      NOW ITERATE ON UP TO 5 OPTIONS TO CALCULATE XK
ISN 0023      C      THE RELATIVE SOLIDS HOLDUP IN THE RUBBLE WAKE
ISN 0023      DO 100 KL = 46,50
ISN 0024      KS(19) = KS (KL)
ISN 0025      IF (KS(19), LT, 0) GO TO 100
ISN 0027      C      A VALUE OF LESS THAN ZERO IMPLIES THERE IS NO MORE TO BE DONE
ISN 0027      C      MAIN LOOP TO CALCULATE HOLDUPS STARTS HERE
ISN 0027      15 E(21) = E(1)
ISN 0028      F(22) = F(2)
ISN 0029      E(23) = F(9)
ISN 0030      C      SEE IF TOO MANY ITERATIONS
ISN 0030      KS(42) = KS(42) + 1
ISN 0031      IF (KS(42), GT, KS(43)) GO TO 300
ISN 0033      C
ISN 0033      C      IF FINES PRESENT, APPORTION SLURRY TO LIQ + FINES
ISN 0033      F(25) = 100.0*V(5)*F(1)/(F(1)*V(5) + F(14)*(100.-V(5)))
ISN 0034      E(3) = E(4) * (100.-F(25))/100.
ISN 0035      F(4) = E(4)*F(25)/100.
ISN 0036      C      DETERMINE LINEAR VELOCITIES BASED ON THESE VOLUME FRACTIONS
ISN 0036      CALL LINEAR
ISN 0036      C
ISN 0036      C      DETERMINE RUBBLE TERMINAL VELOCITY-LEAVE HERE TO ALLOW

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C      POSSIBLE FUTURE CORRELATIONS INVOLVING VOLUME FRACTIONS
C
ISN 0037      CALL UTH
C
C      AND USE BUBBLE TERMINAL VELOCITY IN BHATTIA-EPSTEIN MODEL
C      TO CALCULATE PHASE HOLDUPS
C
ISN 0034      CALL BHATTIA
C
C      AND BACK SUBSTITUTE CALCULATED HOLDUPS
ISN 0039      E(2) = Z(13)
ISN 0040      E(9) = Z(19)
ISN 0041      E(3) = E(9)*(100.-E(25))/100.
ISN 0042      E(4) = E(9)*E(25)/100.
ISN 0043      E(1) = 1.-E(2)-E(9)
C
C      IS FURTHER ITERATION NECESSARY?
ISN 0044      DEL1 = ABS(E(1)-E(21))
ISN 0045      DEL2 = ABS(E(2)-E(22))
ISN 0046      DEL9 = ABS(E(9)-E(23))
ISN 0047      IF 1 (DEL1.GT.0.01). OR . (DEL2.GT.0.01). OR . (DEL9.GT.0.01))
C      GO TO 15
C      AND DETERMINE BUBBLE DIAMETER FROM BUBBLE TERMINAL VELOCITY
ISN 0049      CALL PEEHLE
C
C      DETERMINE DARTON-HARRISON DRIFT FLUX, VCD
C
ISN 0050      CALL DRIFT
C
C      PERCENT BED EXPANSION FROM CATALYST PROPERTIES AND EC
ISN 0051      E(40) = 100 * (((F(9)/F(10))/E(1))-1.)
C
ISN 0052      30 CONTINUE
ISN 0053      WRITE (6,1979) (C(I),I=1,43), E(14), V(21), (E(I),I=1,3), E(40),
C      F(17),E(15))
ISN 0054      1978 FORMAT (1H,6HRUN ID,2X,8HCOMMENTS,8X,2HUL,4X,2HUG,
C      1 5X,2HEC,4X,2HEG,4X,2HEL,2X,4H% RE,4X,2H%K,5X,3HULH//)
ISN 0055      1979 FORMAT (1H,3A2,3X,9HPREDICTED,2A,2F7.3,3F6.2,F6.0,2F7.3)
ISN 0056      100 CONTINUE
C      WRITE TRAILER
ISN 0057      WRITE(6,1980)
ISN 0058      1980 FORMAT(1H,90(1H_))
C
C      READ NEW SET OF OPERATING CONDITIONS
ISN 0059      300 GO TO 10
ISN 0060      9999 STOP
ISN 0061      END

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MAIN / SIZE OF PROGRAM 000640 HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.
C SF	C	H04	0008HH	F SFA	C	H04	000898	F F	C	H04	000A2H	I F	I04	000194	
L SF	C	I04	00010H	S	C	H04	NR	V F	C	H04	000578	Z F	C	H04	000708
EC	C	H04	NR	SD	C	H04	NR	IC	C	I04	NR	IV	C	I04	NR
I2	C	I04	NR	EL SF	C	I04	00019C	KS SF	C	I04	001068	TH	C	H04	NR
EPG	C	H04	NR	EIP	C	H04	NR	EGG	C	H04	NR	EGP	C	H04	NR
ELG	C	H04	NR	ELP	C	H04	NR	UTH SF	XF		000000	ZGR	C	H04	NR

REQUESTED OPTIONS: NOTERM

OPTIONS IN EFFECT: NAME(MAIN) ADJUSTSIZE LINECOUNT(40) SIZE(MAX) AUTODHL(NONE)
 SOURCE ERCDIC POLIST NODECK OBJECT MAP NOFORMAT GUSTMI NOXREF NOALC NOANSF NOTERM TRM FLAG(1) XI

```

ISN 0002      SUBROUTINE RHATIA
ISN 0003      COMMON EC(20),EGP(10),EFP(10),FLP(10),EGG(100),EFG(100),ELG(100)
ISN 0004      COMMON V(100),Z(100),E(100),F(100),G(100),GR(100),ZGR(100)
ISN 0005      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT
C
C          TEST TO ENSURE THAT THREE PHASES EXIST
C
ISN 0006      IF ( V(21) .LE. 0.00001) GO TO 9999
ISN 0008      IF ( E(14) .LE. 0.00001) GO TO 9999
C          ASSIGN MEASURED VALUES TO OLD VARIABLE LOCATIONS
ISN 0010      F(41) = E(1)
ISN 0011      E(42) = E(2)
ISN 0012      E(48) = E(18)
ISN 0013      E(43) = 1.0-E(2)- E(1)
C
C          LIMIT TO ITERATIONS, DEFAULT VALUE WILL BE 20
ISN 0014      KKLIM = KS(30)
ISN 0015      IF(KKLIM.LE.0) KKLIM=20
C
C          EQUATION 8  VGL = UTB + 2 UG
ISN 0017      Z(10)=E(51)+2.*V(21)
C          NUMBER OF ITERATIONS COUNTER
ISN 0018      KK=1
C          DETERMINE XK
ISN 0019      10 CALL ELTAM
C
C          NOW DETERMINE RATIO OF WAKE VOLUME TO BUBBLE VOLUME
C          SEE EL-TENTANY REFERENCE IN SUBROUTINE ELTAM
ISN 0020      CALL WAKE
C
ISN 0021      Z(11) = E(56) * EXP( -5.08* E(42))
ISN 0022      Z(12) = Z(11) * E(42)
C
C          GAS LINEAR VELOCITY IN BED
ISN 0023      E(48)=V(21)/E(42)
C
C          LIQUID HOLDUP (FLP) EQUATION 5-- TEST ARGUMENT FIRST      BEFORE
C          BEFORE RAISING TO EXPONENT 1/N
ISN 0024      20 Z(23)=((E(14)-E(48)*Z(12)* (1.-F(17))))/(F(21)*(1.-F(42)-Z(12)))
C          TEST TO SEE IF ZERO--IF SO, RETURN
ISN 0025      IF (Z(23).LE.0) RETURN
ISN 0027      Z(13) = Z(23) ** (1./E(52))
C
C          CALCULATE NEW LIQUID GAS VELOCITY, VG, EQUATION 4
ISN 0028      30 Z(16)=(E(14)+V(21)+Z(13)*(1.-F(42)-Z(12))*Z(10))/(1.-E(41))
C
C          AVERAGE GAS VELOCITY FOR NEW GUESS
ISN 0029      Z(17)=(Z(16)+E(48))/2.
C          NEW VALUE OF F0
ISN 0030      Z(18)=V(21)/Z(17)
C
C          CALCULATE NEW VALUE OF (EL+EF) FROM EQUATION 2
ISN 0031      Z(19)=Z(17)*(1.-F(17))+Z(13)*(1.-F(42)-Z(12)*F(17)*Z(12))

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C
C   CALCULATE NEW VALUE OF EC FROM EQUATION 3
C
ISN 0032      40 Z(20)=1.-Z(19)-Z(18)
C
C   TEST CONVERGENCE
ISN 0033      125 TEG=ABS(Z(18)-E(42))
ISN 0034      TEL=ABS(Z(19)-E(43))
ISN 0035      TVG=ABS(Z(16)-E(48))
ISN 0036      IF (TEL.LE.0.01 .AND. TEG.LE.0.1 .AND. TVG.LE.0.01) GO TO 200
ISN 0038      IF (KK.GE.KKLM) GO TO 200
ISN 0040      KK=KK+1
ISN 0041      E(42)=Z(18)
ISN 0042      E(43)=Z(19)
ISN 0043      E(41)=Z(20)
ISN 0044      GO TO 10
ISN 0045      200 CONTINUE
ISN 0046      250 CONTINUE
ISN 0047      RETURN
C
C   ERROR SECTION- ZERO VELOCITY
ISN 0043      9999 CONTINUE
ISN 0049      WRITE (6,9998)
ISN 0050      9998 FORMAT (1H1, ' ZERO VELOCITY FOR BHATIA-EPSTEIN MODEL')
ISN 0051      STOP
ISN 0052      END

```

BHATIA / SIZE OF PROGRAM 000552 HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	
C	C	R*	NH	E SFA	C	R*	000838	F F	C	R*	000A2E	S	C	R*	NR	
V	F	C	R*	Z SFA	C	R*	000738	EC	C	R*	NR	GR	C	R*	NR	
IC	C	I*	NR	IV	C	I*	NR	IZ	C	I*	NR	KK SF	I*	I*	000128	
KS	F	C	I*	IM	C	R*	NR	EFG	C	R*	NR	EFP	C	R*	NR	
EGG	C	R*	NR	EGP	C	R*	NR	ELG	C	R*	NR	ELP	C	R*	NR	
EXP	F	XF	R*	TEG S	R*	R*	00012C	TEL S	R*	R*	000130	TVG S	R*	R*	000134	
ZGR	C	R*	NR	WAKE SF	XF	R*	000000	ELIAM SF	XF	R*	000000	KKLIM S	I*	I*	000138	
FRAPR#	XF	R*	000000	BHATIA		R*	00013C	IBCOM#	F	XF	I*	000000	KCOUNT	C	I*	NR

***** COMMON INFORMATION *****

NAME OF COMMON BLOCK * * SIZE OF BLOCK 001404 HEXADECIMAL BYTES

VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.
EC	R*	000000 NR	EFG	R*	000050 NR	EFP	R*	000078 NR	ELP	R*	0000A0 NR
EGG	R*	0000C0 NR	FFG	R*	000258 NR	ELG	R*	0003E8 NR	V	R*	000578 NR
Z	R*	000708 NR	F	R*	000498 NR	F	R*	000A28 NR	C	R*	000888 NR
IM	R*	000048 NR	ZGR	R*	000E08 NR	KS	I*	001068 NR	S	R*	001130 NR
TV	R*	001180 NR	IZ	I*	001310 NR	IV	I*	001360 NR	IC	I*	001380 NR
KCOUNT	I*	001400 NR									

SOURCE STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
-------	-----	------	-------	-----	------	-------	-----	------	-------	-----	------

REQUESTED OPTIONS: NONE

OPTIONS IN EFFECT: NAME(MAIN) NOOPTIMIZE LINECOUNT(50) SIZE(MAX) AUTODHL(NONE)
 SOURCE EXECIC NOLIST NOCHECK OBJECT MAP NOFORMAT GOSTMI NOXREF NOALC NOANSF NOTERM IBM FLAG(I) XL

```

ISN 0002      SUBROUTINE CHGCU
ISN 0003      COMMON EC(20),EGP(10),FFP(10),ELP(10),EGG(100),EFG(100),ELG(100)
ISN 0004      COMMON V(100),Z(100),E(100),F(100),C(100),GW(100),ZGR(100)
ISN 0005      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT

```

```

C
C      THIS SUBROUTINE ALLOWS INDIVIDUAL VECTOR VALUES
C      TO BE CHANGED AFTER ALL NORMAL INPUT IS READ IN
C

```

```

ISN 0006      DIMENSION VARNAM(20)
ISN 0007      DIMENSION IMX (20)
ISN 0008      DATA VARNAM/3H EC,3HEGP,3HEFP,3HELP,3HEGG,
1              3HEFG,3HELG,3H V,3H Z,3H E,
2              3H F,3H C,3H GW,3HZGR,3H KS,
3              3H S,3H TM,3H IZ,3H IV,3H IC/
ISN 0009      DATA HL/3H /
ISN 0010      DATA IMX/20,3*10,10*100,50,20,100,3*20/
ISN 0011      DATA DASH/1H-/

```

```

C      BEGIN
C      BYPASS      CARUS      WITH FORM 315-08

```

```

ISN 0012      28 READ(5,25,END=400) LC,VHL
ISN 0013      25 FORMAT(13,A1)

```

```

C      IS THIS THE LAST CARD?
IF(LC.GE.998) GO TO 400
IF(VHL.EQ.DASH) GO TO 28

```

```

C      NOT END CARD. NOT A DASH-IMPLIES THIS IS A CHANGE CARD

```

```

ISN 0018      BACKSPACE 5
ISN 0019      JNAMES=20
ISN 0020      50 READ(5,100,END=400) LC,VHL,IS,IF
ISN 0021      100 FORMAT(13,A3,2I3)

```

```

C      LAST CARD?
IF(LC.GE.998) GO TO 400

```

```

C      BYPASS BLANK CARD

```

```

ISN 0024      IF(VHL.EQ.HL) GO TO 50
C      TEST FOR PROPER SEQUENCE

```

```

ISN 0026      IF(IS.LE.0) GO TO 900
ISN 0028      IF(IF.LE.0) GO TO 900
ISN 0030      IF(IS.GT.1F) GO TO 900

```

```

C
C      MAIN LOOP TO IDENTIFY VARIABLE NAME
C

```

```

ISN 0032      DO 200 J=1,JNAMES
ISN 0033      JS=J
ISN 0034      IF(VHL.EQ.VARNAM(J)) GO TO 225
ISN 0036      200 CONTINUE

```

```

C      NO MATCH - ERROR
C

```

```

ISN 0037      GO TO 920

```

```

C      MATCH-OK. TEST FOR VALID UPPER LIMIT
C

```

```

ISN 0038      225 IF(IF.GT.IMX(JS)) GO TO 910
C

```

C READ THE DATA

C

```

ISN 0040      GO TO (1,2,3,4,5,6,7,8,9,10,11,12,13,14,15,16,17,18,19,20),JS
ISN 0041      1 READ(5,280) (EC(I),I=1S,IF)
ISN 0042      GO TO 50
ISN 0043      2 READ(5,280) (EGP(I),I=1S,IF)
ISN 0044      GO TO 50
ISN 0045      3 READ(5,280) (EFP(I),I=1S,IF)
ISN 0046      GO TO 50
ISN 0047      4 READ(5,280) (ELP(I),I=1S,IF)
ISN 0048      GO TO 50
ISN 0049      5 READ(5,280) (EGG(I),I=1S,IF)
ISN 0050      GO TO 50
ISN 0051      6 READ(5,280) (EFG(I),I=1S,IF)
ISN 0052      GO TO 50
ISN 0053      7 READ(5,280) (ELG(I),I=1S,IF)
ISN 0054      GO TO 50
ISN 0055      8 READ(5,280) ( V(I),I=1S,IF)
ISN 0056      GO TO 50
ISN 0057      9 READ(5,280) ( Z(I),I=1S,IF)
ISN 0058      GO TO 50
ISN 0059      10 READ(5,280) ( E(I),I=1S,IF)
ISN 0060      GO TO 50
ISN 0061      11 READ(5,280) ( F(I),I=1S,IF)
ISN 0062      GO TO 50
ISN 0063      12 READ(5,280) ( C(I),I=1S,IF)
ISN 0064      GO TO 50
ISN 0065      13 READ(5,280) ( GN(I),I=1S,IF)
ISN 0066      GO TO 50
ISN 0067      14 READ(5,280) (ZGR(I),I=1S,IF)
ISN 0068      GO TO 50
ISN 0069      15 READ(5,380) ( KS(I),I=1S,IF)
ISN 0070      GO TO 50
ISN 0071      16 READ(5,280) ( S(I),I=1S,IF)
ISN 0072      GO TO 50
ISN 0073      17 READ(5,280) ( TM(I),I=1S,IF)
ISN 0074      GO TO 50
ISN 0075      18 READ(5,380) ( IZ(I),I=1S,IF)
ISN 0076      GO TO 50
ISN 0077      19 READ(5,380) ( IV(I),I=1S,IF)
ISN 0078      GO TO 50
ISN 0079      20 READ(5,380) ( IC(I),I=1S,IF)
ISN 0080      GO TO 50
ISN 0081      280 FORMAT(8F4.0,8X)
ISN 0082      380 FORMAT(8I9,8X)
ISN 0083      400 RETURN

```

C

C ERROR SECTION

C

```

ISN 0084      900 WRITE (5,905)
ISN 0085      905 FORMAT(1H0,2HIMPROPER CHANGE CARD INDICES)
ISN 0086      GO TO 950
ISN 0087      920 WRITE (5,925)
ISN 0088      925 FORMAT(1H0,2HUNKNOWN VARIABLE NAME)
ISN 0089      GO TO 950
ISN 0090      930 WRITE (5,935)
ISN 0091      935 FORMAT(1H0,3HFINAL INDEX TOO HIGH FOR ARRAY)
ISN 0092      GO TO 950

```

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```

ISN 0093      950 4011F (6.955) LC-VBL+TS+IF
ISN 0094      955 FORMAT(1H+(13.43+213.10X+14H(CARD (IGNORED)))
ISN 0095      960 TO 50
ISN 0096      9999 RETURN
ISN 0097      END

```

/ CHRGD /				SIZE OF PROGRAM 000A4A HEXADECIMAL BYTES											
NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.
C	S	C	R04 000000	F	S	C	R04 000000	F	S	C	R04 000A28	I	F	I04 000180	
J	SF		I04 000184	S	S	C	R04 001130	V	S	C	R04 000578	Z	S	C	R04 000708
BL			R04 000188	EC	S	C	R04 000000	GR	S	C	R04 000048	IC	S	C	I04 001380
IF	SF		I04 00018C	IS	SF		I04 000190	IV	S	C	I04 001360	IZ	S	C	I04 001310
JS	SF		I04 000194	KS	S	C	I04 001068	LC	SF		I04 000198	TM	S	C	R04 001180
ELG	S	C	R04 000258	ELP	S	C	R04 000078	EGG	S	C	R04 0000C8	EGP	S	C	R04 000050
ELG	S	C	R04 0003F8	ELP	S	C	R04 0000A0	IMX			I04 0001AC	VBL	SF		R04 00019C
ZGR	S	C	R04 000ED8	DASH			R04 0001A0	CHRGD			R04 0001A4	TBCOMM	F	XF	I04 000000
JNAMES	SF		I04 0001AB	KCOUNT	C		I04 NR	VARNAM			R04 0001FC				

***** COMMON INFORMATION *****

NAME OF COMMON BLOCK * * SIZE OF BLOCK 001404 HEXADECIMAL BYTES

VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.
EC	R04	000000	FGP	R04	000050	ELP	R04	000078	ELP	R04	0000A0
EGG	R04	0000C8	FFG	R04	000258	ELG	R04	0003E8	V	R04	000578
Z	R04	000708	F	R04	000898	F	R04	000A28	C	R04	000888
GR	R04	000048	ZGR	R04	000ED8	KS	I04	001068	S	R04	001130
TM	R04	001180	IZ	I04	001310	IV	I04	001360	IC	I04	001380
KCOUNT	I04	001400 NR									

SOURCE STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
28	12	000300	50	20	000358	200	36	0003F6	225	38	000412
1	41	000442	2	43	000486	3	45	0004CA	4	47	00050E
5	49	000552	6	51	000596	7	53	0005DA	8	55	00061E
9	57	000662	10	59	0006A6	11	61	0006EA	12	63	00072E
13	65	000772	14	67	0007B6	15	69	0007FA	16	71	00083E
17	73	000882	18	75	0008C6	19	77	00090A	20	79	00094E
400	83	000992	900	84	00099A	920	87	0009B6	930	90	0009D2
950	93	0009EE	9999	96	000A2A NR						

COMPILER GENERATED LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
100002	12	000314	100003	16	000336	100004	18	000344	100005	20	00036C
100006	24	00039E	100007	26	0003AC	100008	28	0003BA	100009	30	0003C4
100010	32	0003D2	100011	33	0003DA	100012	37	00040C	100013	40	000428

FORMAT STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
25	13	000028	100	21	00002E	280	81	000038	380	82	000041

REQUESTED OPTIONS: NOTERM

OPTIONS IN EFFECT: NAME (MAIN) NOOPTIMIZE LINECOUNT (60) SIZE (MAX) AUTODHL (NONE)

SOURCE EXECIC NOLIST NOCHECK OBJECT MAP NOFORMAT GUSTMI NOXREF NOALC NOANSF NOTERM IBM FLAG(1) XL

```

ISN 0002      SUBROUTINE DRIFT
ISN 0003      COMMON EC(20),EGP(10),EFP(10),ELP(10),EGG(100),EFG(100),ELG(100)
ISN 0004      COMMON V(100),Z(100),E(100),F(100),C(100),GR(100),ZGR(100)
ISN 0005      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT

C
C          CALCULATE TWO OR THREE PHASE DRIFT-FLUX
C          UNITS WILL BE IN MM/SEC FOR COMPARISON WITH LIT.
C

ISN 0006      30 E(53)=((V(21)*(1.-E(2)))-(F(14)*E(2)*(1.-E(2))/(E(3)+F(4))))*304.8
ISN 0007      40 CONTINUE
ISN 0008      IF (V(21).LE.0.0)E(53)=0.0
ISN 0010      RETURN
ISN 0011      END

```

/ DRIFT / SIZE OF PROGRAM 000172 HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.
C	C	R*	NR	E SF	C	R*	000898	F	C	R*	NR	S	C	R*	NR
V	F	R*	000574	7	C	R*	NR	EC	C	R*	NR	GR	C	R*	NR
IC	C	I*	NR	IV	C	I*	NR	IZ	C	I*	NR	KS	C	I*	NR
TM	C	R*	NR	EFG	C	R*	NR	EFP	C	R*	NR	EGG	C	R*	NR
EGP	C	R*	NR	ELG	C	R*	NR	ELP	C	R*	NR	ZGR	C	R*	NR
DRIFT		R*	00009C	KCOUNT	C	I*	NR								

***** COMMON INFORMATION *****

NAME OF COMMON BLOCK * * SIZE OF BLOCK 001404 HEXADECIMAL BYTES

VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.
EC	R*	000000 NR	EGP	R*	000050 NR	EFP	R*	000078 NR	ELP	R*	0000A0 NR
EGG	R*	0000C8 NR	EFG	R*	000258 NR	ELG	R*	0003E8 NR	V	R*	00057A NR
Z	R*	000708 NR	E	R*	000898 NR	F	R*	000A28 NR	C	R*	0008BA NR
GR	R*	000048 NR	ZGR	R*	000ED8 NR	KS	I*	001068 NR	S	R*	001130 NR
TM	R*	001180 NR	IZ	I*	001310 NR	IV	I*	001360 NR	IC	I*	001380 NR
KCOUNT	I*	001400 NR									

SOURCE STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
30	5	0000CC	40	7	000136 NR						

COMPILER GENERATED LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
100002	9	000146	100003	10	000152						

*OPTIONS IN EFFECT:NAME (MAIN) NOOPTIMIZE LINECOUNT (60) SIZE (MAX) AUTODHL (NONE)

*OPTIONS IN EFFECT:SOURCE EXECIC NOLIST NOCHECK OBJECT MAP NOFORMAT GUSTMI NOXREF NOALC NOANSF NOTERM IBM FLAG(1) XL

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REQUESTED OPTIONS: NONE

OPTIONS IN EFFECT: NAME (NAME) OPTIMIZE LINKCOUNT (NO) SIZE (MAX) AUTODIAG (NONE)
 SOURCE EXECUTIC (LIST) NOFORN OBJECT MAP NOFORMAT GUSTHI NOXREF NOALC NOANSF NOTERM TBM FLAG(I) XL

```

ISN 0002      SUBROUTINE ELTAM
C
C      CALCULATES AN FOR FLUIDIZED BEDS BY CORRELATION OF EL-TAMTAMI
C      REF: INT. J. MULTIPHASE FLOW 4, 19 (1978)
C
ISN 0003      COMMON EC(20),EGP(10),EFP(10),ELP(10),EGG(100),EFG(100),ELG(100)
ISN 0004      COMMON V(100),Z(100),F(100),F(100),C(100),GP(100),ZGR(100)
ISN 0005      COMMON KS(50), S(20),TM(100),I7(20),IV(20),IC(20),KCOINT
C
C
C      CALCULATE VOL. VELOCITY OF GAS RELATIVE TO TOTAL LIQUID
C      BASED ON GAMMA RAY SCAN RESULTS
C
ISN 0006      Z(25) = V(21)/E(42) - E(14)/(1.0 - E(41) - E(42))
C
ISN 0007      Z(27) = Z(25)
C
C      ACTUAL CORRELATION FOR XK, FORCING IT TO BE >= 0
C      *****
ISN 0008      Z(26) = (1 - 0.877 * F(21)) / Z(27)
ISN 0009      IF (Z(26) .LE. 0.) Z(26) = 0.
C
C      SEE IF COMPUTED OR READ-IN VALUE TO BE USED
ISN 0011      IF (KS(19) .EQ. 3) GO TO 3500
ISN 0013      IF (KS(19) .EQ. 2) GO TO 1500
ISN 0015      IF (KS(19) .EQ. 0) GO TO 500
ISN 0017      F(17) = E(55)
ISN 0018      RETURN
C
C      FOR THIS PATH, ASSIGN EL-TEM TAMY VALUE TO F(17)
ISN 0019      500 F(17) = Z(26)
C
C
C      NOTE IN ABOVE EXPRESSION THAT VELOCITY USED IS CATALYST
C      TERMINAL VELOCITY AS USED FOR RICHARDSON/ZAKI EXPRESSION
ISN 0020      1000 CONTINUE
ISN 0021      RETURN
C
ISN 0022      1500 CONTINUE
C      HIGHEST LIQUID VELOCITY-USE HIGHEST VALUES ONLY, NOT EXTRAP.
ISN 0023      2700 Z(28) = TM(75)
ISN 0024      Z(29) = TM(74)
C
C      ACTUAL CORRELATION HERE
C      FORM: (1-X)^(1-1) = SLOPE * UTH + INTERCEPT
ISN 0025      3000 F(17) = Z(24) * F(51) + Z(29)
ISN 0026      F(17) = 1. - (1./F(17))
ISN 0027      RETURN
C
C      THIS SECTION FOR KS (19) = 3
C      AT LOW VELOCITIES, USE THE EL-TEM TAMY CORRELATION
C      AT HIGH VELOCITIES, USE THE TM-BASED CORRELATION
ISN 0028      3500 IF (E(14) .GT. 1.0000) GO TO 1500
ISN 0030      GO TO 500

```

ISN 0031 4000 RETURN
ISN 0032 END

/ ELTAM / SIZE OF PROGRAM 000246 HEXADECIMAL BYTES

NAME	TAG	TYPE	ADDR.	NAME	TAG	TYPE	ADDR.	NAME	TAG	TYPE	ADDR.	NAME	TAG	TYPE	ADDR.		
C	C	R04	NR	F	F	C	R04	000898	F	SF	C	R04	000A28	S	C	R04	NR
V	F	C	R04	000578	/	SF	C	R04	000708	EC	C	R04	NR	GR	C	R04	NR
IC	C	I04	NR	IV	C	I04	NR	IZ	C	I04	NR	KS	C	I04	NR	001068	
TM	F	C	R04	001180	ELG	C	R04	NR	EFP	C	R04	NR	EGG	C	R04	NR	
EGP	C	R04	NR	ELG	C	R04	NR	ELP	C	R04	NR	ZGR	C	R04	NR		
ELTAM		R04	000004	KCOUNT	C	I04	NR										

***** COMMON INFORMATION *****

NAME OF COMMON BLOCK * * SIZE OF BLOCK 001404 HEXADECIMAL BYTES

VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.
EC	R04	000000 NR	EGP	R04	000050 NR	EFP	R04	000078 NR	ELP	R04	0000A0 NR
EGG	R04	0000C8 NR	FFG	R04	000025A NR	ELG	R04	0000E8 NR	V	R04	000057A
/	R04	000108	E	R04	000089A	F	R04	0000A28	C	R04	000088A NR
GM	R04	000048 NR	ZGM	R04	0000ED8 NR	KS	I04	001068	S	R04	001130 NR
TM	R04	001180	IZ	I04	001310 NR	IV	I04	001360 NR	IC	I04	001380 NR
KCOUNT	I04	001400 NR									

SOURCE STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
500	19	0001F0	1000	20	0001FC NR	1500	22	000204	2700	23	000204 NR
3000	25	000224 NR	3500	28	000262	4000	31	000286 NR			

COMPILER GENERATED LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
100001	2	000120	100002	10	00013C	100003	11	0001A8	100004	13	0001BA
100005	15	0001CC	100006	17	0001DC	100007	30	000280			

*OPTIONS IN EFFECT*NAME(MAIN) NOOPTIMIZE LINECOUNT(60) SIZE(MAX) AUTOUNL(NONE)

*OPTIONS IN EFFECT*SOURCE EPICICNDLIST NODECK OBJECT MAP NOFORMAT GUSTMT NOXREF NOALC NOANSF NOTERM IBM FLAG(I) XL

STATISTICS SOURCE STATEMENTS = 11, PROGRAM SIZE = 678, SUBPROGRAM NAME = ELTAM

STATISTICS NO. DIAGNOSTICS GENERATED

***** END OF COMPILATION *****

3024 BYTES OF CORE NOT USED

REQUESTED OPTIONS: NOTERM

OPTIONS IN EFFECT: NAME (DATE) NOOPTIMIZE LINECOUNT (NO) SIZE (MAX) AUTODIAG (NONE)

SOURCE EXECIC NOLIST NODECK OBJECT MAP NOFORMAT GUSTMI NOXREF NOALC NOANSF NOTERM IBM FLAG(1) XL

```

      ISN 0002      SUBROUTINE FLOWS
      C             CONVERTS FEET/SECOND TO OTHER UNITS
      ISN 0003      COMMON EC(20),EGP(10),EFP(10),ELP(10),EGG(100),EFG(100),ELG(100)
      ISN 0004      COMMON V(100),Z(100),E(100),F(100),C(100),GR(100),ZGR(100)
      ISN 0005      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT
      C
      ISN 0006      IF (F(14),GT,0.0) GO TO 10
      C             CALCULATE SLURRY DENSITY
      ISN 0008      E(8)=F(1)
      ISN 0009      GO TO 11
      ISN 0010      10 E(8)=(100.*F(1)*F(14))/(V(5)*F(1)+(100.-V(5))*F(14))
      C             CALCULATE LIQUID FLOWRATES
      ISN 0011      11 V(19) = E(14)*60.*7.48*(3.1416*C(4)**2/4.)
      C
      C
      ISN 0012      E(17)=V(19)/(3.1416*C(4)**2/4.)
      C
      C
      C             CALCULATE GAS FLOW RATES,CFM
      ISN 0013      E(20) = (3.1416*C(4)**2/4.)*V(21)*3600.
      ISN 0014      RETURN
      ISN 0015      END

```

/ FLOWS / SIZE OF PROGRAM 00021A HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.
C F	C	R*	0008H	F SF	C	R*	00089A	F F	C	R*	000A28	S	C	R*	NR
V SF	C	R*	00057B	7	C	R*	NR	EC	C	R*	NR	GR	C	R*	NR
IC	C	I*	NR	IV	C	I*	NR	IZ	C	I*	NR	KS	C	I*	NR
TM	C	R*	NR	EFG	C	R*	NR	EFP	C	R*	NR	EGG	C	R*	NR
EGP	C	R*	NR	ELG	C	R*	NR	ELP	C	R*	NR	ZGR	C	R*	NR
FLOWS		R*	0000HC	KCOUNT	C	I*	NR								

***** COMMON INFORMATION *****

NAME OF COMMON BLOCK * * SIZE OF BLOCK 001404 HEXADECIMAL BYTES

VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.
EC	R*	000000 NR	EGP	R*	000050 NR	EFP	R*	000078 NR	ELP	R*	0000A0 NR
EGG	R*	0000C0 NR	EFG	R*	000258 NR	ELG	R*	0003E8 NR	V	R*	00057A NR
Z	R*	000708 NR	F	R*	00089A NR	F	R*	000A28 NR	C	R*	00088A NR
GR	R*	00004A NR	ZGR	R*	000ED8 NR	KS	I*	001068 NR	S	R*	001130 NR
TM	R*	001180 NR	IZ	I*	001310 NR	IV	I*	001360 NR	IC	I*	001380 NR
KCOUNT	I*	001400 NR									

SOURCE STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
10	10	000112	11	11	00016A						

REQUESTED OPTIONS: NOTERM

OPTIONS IN EFFECT: NAME(MAIN) OPTIMIZE LINECOUNT(50) SIZE(MAX) AUTODHL(NONE)

SOURCE FRC01C MODLIST MODPCK OBJECT MAP NOFORMAT GOSTMT NOXREF NOALC NOAMSF NOTERM THM FLAG(1) XL

```

ISN 0002      SUBROUTINE LINEAR
ISN 0003      COMMON EC(20),EGP(10),EFP(10),ELP(10),EGG(100),EFG(100),ELG(100)
ISN 0004      COMMON V(100),Z(100),E(100),F(100),S(100),GR(100),ZGR(100)
ISN 0005      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT
                C
                C      CALCULATE LINEAR VELOCITIES FT/SEC
                C
ISN 0006      E(15)=E(14)/(F(3)+E(4))
                C      CALCULATE GAS FLOWS
ISN 0007      IF(V(21).LE.0.) GO TO 12
ISN 0009      E(18) = V(21)/E(2)
ISN 0010      GO TO 15
ISN 0011      12 E(18) = 0.
ISN 0012      15 CONTINUE
ISN 0013      RETURN
ISN 0014      END

```

/ LINEAR / SIZE OF PROGRAM 000142 HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.
C	C	R*	NR	F SF	C	R*	000898	F	C	R*	NR	S	C	R*	NR
V	F	C	R*	7	C	R*	NR	EC	C	R*	NR	GR	C	R*	NR
IC	C	I*	NR	IV	C	I*	NR	IZ	C	I*	NR	KS	C	I*	NR
TM	C	R*	NR	EFG	C	R*	NR	EFP	C	R*	NR	EGG	C	R*	NR
EGP	C	R*	NR	ELG	C	R*	NR	ELP	C	R*	NR	ZGR	C	R*	NR
KCOUNT	C	I*	NR	LINEAR		I*	000098								

***** COMMON INFORMATION *****

NAME OF COMMON BLOCK * * SIZE OF BLOCK 001404 HEXADECIMAL BYTES

VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.
EC	R*	000000 NR	EGP	R*	000050 NR	EFP	R*	000078 NR	ELP	R*	0000A0 NR
EGG	R*	0000C8 NR	EFG	R*	000258 NR	ELG	R*	0003E8 NR	V	R*	000578 NR
Z	R*	000708 NR	E	R*	000498 NR	F	R*	000A28 NR	C	R*	000888 NR
GR	R*	000D48 NR	ZGR	R*	000E08 NR	KS	I*	001068 NR	S	R*	001130 NR
TM	R*	001180 NR	IZ	I*	001310 NR	IV	I*	001360 NR	IC	I*	001380 NR
KCOUNT	I*	001400 NR									

SOURCE STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
12	11	000116	15	12	000122						

COMPILER GENERATED LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
100001	2	0000C0	100002	4	0000F8						

OPTIONS IN EFFECT: NAME(MAIN) OPTIMIZE LINECOUNT(50) SIZE(MAX) AUTODHL(NONE)

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REQUESTED OPTIONS: NOTERM

OPTIONS IN EFFECT: NAME(MAIN) NOOPTIMIZE LINCOUNT(60) SIZE(MAX) AUTODHL(NONE)

SOURCE FPCDIC NOLIST NODECK OBJECT MAP NOFORMAT GUSTMI NOXREF NUALC NOANSF NOTERM IBM FLAG(I) XL

```

      ISN 0002      SUBROUTINE PEEBLE
      C              THIS SUBROUTINE DETERMINES THE HUBBLE DIAMETER FROM THE
      C              TERMINAL VELOCITY. IT USES THE INVERSE OF THE PEEBLES
      C              AND GARBER CORRELATIONS.
      C
      C              REF: PEEBLES, F.M. AND GARBER, M.J.,
      C              CHEM. ENG. PROGRESS 49 2 (1953)
      ISN 0003      COMMON EC(20),EGP(10),EFP(10),ELP(10),EGG(100),EFG(100),ELG(100)
      ISN 0004      COMMON V(100),Z(100),F(100),F(100),C(100),GR(100),ZGR(100)
      ISN 0005      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT
      C
      C              TRANSLATE VISCOSITY AND SURFACE TENSION TO ENGLISH UNITS
      ISN 0006      F(23) = F(3)* 2.24809E-6/0.03280833
      C
      C              USE LIQUID OR SLURRY VISCOSITY DEPENDING ON PCT. FINES
      C
      ISN 0007      IF (V(5).LT. 0.01) GO TO 10
      ISN 0009      F(22) = F(5)
      ISN 0010      GO TO 15
      ISN 0011      10 F(22) = F(2)
      ISN 0012      15 F(22) = F(22) * 6.7197E-4
      C
      C              REGION ONE-HUBBLE RADIUS WILL BE OBTAINED IN FEET.
      C
      ISN 0013      KS(41) = 1
      ISN 0014      F(20) = SQRT((9.*E(51)*F(22))/(2.*(E(8)-F(6))*C(6)))
      C
      C              CALCULATE REYNOLDS NUMBER TO SEE IF REGION IS APPLICABLE
      C
      ISN 0015      E(35) = 2.* E(8) *E(51) *F(20)/F(22)
      ISN 0016      IF (E(35).LE.2.) GO TO 1000
      C
      C              REGION 2
      C
      ISN 0018      KS(41) = 2
      ISN 0019      F(20)=(E(51)/(0.33*(C(6)**0.76)*((E(8)/F(22))**0.52))**((1./1.29)
      C
      C              REYNOLDS NUMBER AND MORTON (WEHER) NUMBER
      ISN 0020      E(35) = 2.*E(8)*E(51)*F(20)/F(22)
      ISN 0021      E(36) = C(6)*F(22)**4/(E(8)*F(23)**3*C(7)**3)
      ISN 0022      IF (E(35).LT.2.) GO TO 5000
      ISN 0024      IF (E(35).LT.(4.02*(F(36)**(-0.214)))) GO TO 1000
      C
      C              REGION 3
      C
      ISN 0026      KS(41) = 3
      ISN 0027      F(20) = (F(23)*C(7)*1.35**2)/(E(8)*E(51)**2)
      C
      C              REYNOLDS NUMBER
      C
      ISN 0028      F(35) = 2.*E(8)*E(51) * F(20)/F(22)
      C
      C              62 NUMBER = 22.002/16 FM

```

```

C
ISN 0029      E(37) = C(6) * F(20)**4 * F(51)**4 * E(8)**3/(F(23)**3 * C(7)**3)
ISN 0030      E(36) = C(6) * F(22)**4/(F(8) * F(23)**3 * C(7)**3)

C
C      TEST 1
ISN 0031      IF ((E(35).GT.(4.02*(E(36)**(-0.214))))). AND .
1      (E(35).LT. 3.10 * E(36)**(-0.25)) )    GO TO 1000

C
C      TEST 2
ISN 0033      IF ( (E(37).GT.(16.32 * E(36)**0.144) ) . AND .
1      (F(37).LT. 5.75) )    GO TO 1000

C
C      REGION 4
ISN 0035      KS(41) = 4
ISN 0036      GO TO 5000
ISN 0037      1000 F(38) = 2. * F(20)**30.48
ISN 0038      RETURN

C
C      IF UTR IS TOO HIGH, ENTER A NEGATIVE NUMBER
C      SINCE THE CORRELATION IS FOR SINGLE BUBBLES
C      THIS MAY WELL OCCUR IN THE CHURN-TURBULENT REGION
ISN 0039      5000 CONTINUE
ISN 0040      E(35) = -99.
ISN 0041      E(38) = -95.
ISN 0042      KS(41) = -KS(41)
ISN 0043      RETURN
ISN 0044      END

```

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/ PEFBLE / SIZE OF PROGRAM 00074C HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.				
C	FA	C	R#4	000388	F	SFA	C	R#4	000898	F	SFA	C	R#4	000A28	S	C	R#4	NR	
V		C	R#4	000578	?		C	R#4	NR	EC		C	R#4	NR	GR	C	R#4	NR	
IC		C	I#4	NR	I?		C	I#4	NR	IZ		C	I#4	NR	KS	SF	C	I#4	001068
TM		C	R#4	NR	EFG		C	R#4	NR	EFP		C	R#4	NR	EGG		C	R#4	NR
EGP		C	R#4	NR	ELG		C	R#4	NR	ELP		C	R#4	NR	ZGR		C	R#4	NR
SHORT	F	XF	R#4	000000	FRXPR#		XF	R#4	000000	KCOUNT		C	I#4	NR	PEEBLE		C	R#4	000148

***** COMMON INFORMATION *****

[illegible][illegible]

SOURCE STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR

REQUESTED OPTIONS: NOTERM

OPTIONS IN EFFECT: NAME(MAIN) NOOPTIMIZE LINECOUNT(40) SIZE(MAX) AUTODHL(NONE)
 SOURCE EXECIC NOLIST NODECK OBJECT MAP NOFORMAT GUSTMI NOXREF NOALC NOANSF NOTERM IBM FLAG(1) XL

```

ISN 0002      SUBROUTINE READ
ISN 0003      COMMON EC(20),EGP(10),EFP(10),ELP(10),EGG(100),EFG(100),ELG(100)
ISN 0004      COMMON V(100),Z(100),F(100),F(100),C(100),GR(100),ZGR(100)
ISN 0005      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT
ISN 0006      C      START READING DATA
ISN 0006      C      IF (KCOUNT.GE.1) GO TO 200
ISN 0006      C      READ DATA
ISN 0006      C      READ LIQUID, GAS, CATALYST, AND FINES PROPERTIES
ISN 0006      C      2 CARDS FOR CATALYST PROPERTIES NOW
ISN 0008      READ(5,110) C(1), (C(I), I=21,40)
ISN 0009      C      LIQUID PROPERTIES
ISN 0009      READ(5,120) F(1), F(2), F(5), F(3)
ISN 0010      C      GAS PROPERTIES
ISN 0010      READ(5,130) F(6), F(7)
ISN 0011      C      CATALYST PROPERTIES
ISN 0011      READ(5,140) F(8), F(9), F(10), F(11), F(12)
ISN 0012      C      CATALYST TERMINAL VELOCITY, RICHARDSON/ZAKI INDEX
ISN 0012      READ(5,140) F(21), E(52)
ISN 0013      C      FINES PROPERTIES
ISN 0013      READ(5,150) F(14), F(15)
ISN 0014      C      READ REACTOR AND MISCELLANEOUS
ISN 0014      READ(5,140) C(4), V(5), V(7)
ISN 0015      C      READ COEFFICIENTS FOR CORRELATIONS
ISN 0015      READ(5,140) (TM(I), I=40,42)
ISN 0016      C      COEFFICIENTS FOR WAKE VOLUME RATIO
ISN 0016      READ(5,140) (TM(I), I=43,46)
ISN 0017      C      VELOCITIES FOR UTH CORRELATION
ISN 0017      READ(5,140) (TM(I), I=51,54)
ISN 0018      C      COEFFICIENTS FOR UTH CORRELATION
ISN 0018      READ(5,140) (TM(I), I=55,60)
ISN 0019      C      XK CORRELATION VARIABLES-CROSSOVER VELOCITY + COFFFS
ISN 0019      READ(5,140) TM(65), TM(74), TM(75)
ISN 0020      C      READ TEST DATA
ISN 0020      C
ISN 0020      KCOUNT = 1
ISN 0021      C      READ REACTOR FLOWS (FT/SEC)
ISN 0021      200 CONTINUE
ISN 0022      READ(5,230,END=9999) (C(I), I=41,50), E(14), V(21)
ISN 0023      C
ISN 0023      300 CONTINUE
ISN 0023      C
ISN 0023      C      READ ANY CHANGE CARDS, INCLUDING THE 9999
ISN 0024      400 CALL CHGCI)
ISN 0025      500 FORMAT(13)
ISN 0026      101 FORMAT(2413)
ISN 0027      110 FORMAT(5X,A3,12X,4(542),6X)
ISN 0028      120 FORMAT(10X,5F10.5)
ISN 0029      130 FORMAT(10X,3F10.5)
ISN 0030      140 FORMAT(10X,6F10.5)
ISN 0031      150 FORMAT(10X,3F10.5)
ISN 0032      160 FORMAT(10X,2F10.5,110)
ISN 0033      170 FORMAT(10X,6F10.5)
ISN 0034      180 FORMAT(10X,6F10.5,71)

```

```

ISN 0035      190  FORMAT(20X,110,F10.5,2110)
ISN 0036      210  FORMAT(20X,5F10.5)
ISN 0037      220  FORMAT(20X,5F10.5)
ISN 0038      230  FORMAT(10A2,6F10.5)
ISN 0039      RETURN
ISN 0040      9999 KS(44) = 1
ISN 0041      RETURN
ISN 0042      END

```

/ READ / SIZE OF PROGRAM 00045C HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	
C S	C	R*	000898	F S	C	R*	000898	F S	C	R*	000A28	I F	I*	000180		
S	C	R*	NR	V S	C	R*	000578	Z	C	R*	NR	EC	C	R*	NR	
GH	C	R*	NR	IC	C	I*	NR	IV	C	I*	NR	IZ	C	I*	NR	
KS S	C	I*	001068	TM S	C	R*	001180	ELG	C	R*	NR	ELP	C	R*	NR	
EGG	C	R*	NR	EGP	C	R*	NR	ELG	C	R*	NR	ELP	C	R*	NR	
ZGR	C	R*	NR	READ		R*	000184	CHGCU SF	XF		000000	IBCOM#	F	XF	I*	000000
KCOUNT S	C	I*	001400													

***** COMMON INFORMATION *****

NAME OF COMMON BLOCK * * SIZE OF BLOCK 001404 HEXADECIMAL BYTES

VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.
EC	R*	000000 NR	EGP	R*	000050 NR	ELP	R*	000078 NR	ELP	R*	0000A0 NR
EGG	R*	0000C8 NR	FFG	R*	000258 NR	ELG	R*	0003E8 NR	V	R*	000578
Z	R*	000708 NR	E	R*	000B98	F	R*	000A28	C	R*	000898
GH	R*	000048 NR	ZGR	R*	000E08 NR	KS	I*	001068	S	R*	001130 NR
TM	R*	001180	IZ	I*	001310 NR	IV	I*	001360 NR	IC	I*	001380 NR
KCOUNT	I*	001400									

SOURCE STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
200	21	00030C	300	23	00041C NR	400	24	00041C NR	9999	40	000430

COMPILER GENERATED LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
100001	2	000184	100002	8	0001CA	200001	15	000304	100013	22	00030C
100014	22	0003F0									

FORMAT STATEMENT LABELS

LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR	LABEL	ISN	ADDR
500	25	000028 NR	101	26	00002C NR	110	27	000032	120	28	000043
130	29	00004C	140	30	000055	150	31	00005F	160	32	000067 NR
170	33	000072 NR	180	34	000078 NR	190	35	00008A NR	210	36	000095 NR
220	37	00009E NR	230	38	0000A7						

OPTIONS IN EFFECT: NAME (NAME) SOURCE LINE COUNT (60) SIZE (MAX) AUTODIAG (NONE)

OPTIONS IN EFFECT: SOURCE PROLOG COLIST: SOURCE OBJECT MAP: INFORMATION: GUSINT: NOREF: NOALC: NOANSF: NOTERM: TRM FLAG(1) XL

REQUESTED OPTIONS: NONE

OPTIONS IN EFFECT: NAME(MAIN) NOOPTIMIZE LINECOUNT(60) SIZE(MAX) AUTODBL(NONE)

SOURCE EGGGIC MOLLIST NOCHECK OBJECT MAP NOFORMAT GUSTMI NOXREF NOALC NOANSF NOTERM IBM FLAG(1) XL

```

ISN 0002      SUBROUTINE UTH
C
C      CALCULATES BUBBLE TERMINAL VELOCITY FOR
C      KHATIA EPSTEIN MODEL
C
ISN 0003      COMMON EC(20),EGP(10),FFP(10),ELP(10),EGG(100),EFG(100),ELG(100)
ISN 0004      COMMON V(100),Z(100),E(100),F(100),C(100),GR(100),ZGR(100)
ISN 0005      COMMON KS(50),S(20),TM(100),IZ(20),IV(20),IC(20),KCOUNT
C      USE READ IN VALUE OR CORRELATION ONE?
ISN 0006      IF (KS(20) .EQ. 1) GO TO 100
C
C      CALCULATE DELTA V, DIFFERENCE OF GAS AND LIQUID
C      SUPERFICIAL VELOCITIES
ISN 0008      Z(5) = V(21) - E(34)
ISN 0009      IF (Z(5) .GT. TM(52) ) GO TO 40
ISN 0011      E(51) = TM(55) + TM(56) * Z(5)
ISN 0012      RETURN
ISN 0013      40 IF (Z(5) .GT. TM(53) ) GO TO 60
ISN 0015      E(51) = TM(57) + TM(58) * Z(5)
ISN 0016      RETURN
ISN 0017      60 E(51) = TM(59) + TM(60) * Z(5)
ISN 0018      RETURN
ISN 0019      100 E(51) = E(54)
ISN 0020      RETURN
ISN 0021      END

```

/ UTH / SIZE OF PROGRAM 00020A HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.
C	C	R*	NR	F	SF	C	R*	F	C	R*	NR	S	C	R*	NR
V	F	C	R*	7	SF	C	R*	EC	C	R*	NR	GR	C	R*	NR
IC	C	I*	NR	IV	C	I*	NR	IZ	C	I*	NR	KS	C	I*	001068
TM	F	C	R*	EFG	C	R*	NR	ELP	C	R*	NR	EGG	C	R*	NR
EGP	C	R*	NR	ELG	C	R*	NR	ELP	C	R*	NR	UTH		R*	000084
ZGR	C	R*	NR	KCOUNT	C	I*	NR								

***** COMMON INFORMATION *****

NAME OF COMMON BLOCK * * SIZE OF BLOCK 001404 HEXADECIMAL BYTES

VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.	VAR. NAME	TYPE	REL. ADDR.
EC	R*	000000 NR	EGP	R*	000050 NR	ELP	R*	000078 NR	ELP	R*	0000A0 NR
EGG	R*	0000C8 NR	EFG	R*	000258 NR	ELG	R*	0003E8 NR	V	R*	000578 NR
Z	R*	000708 NR	E	R*	000898 NR	F	R*	000A28 NR	C	R*	0008B8 NR
GR	R*	000148 NR	7GR	R*	000ED8 NR	KS	I*	001068 NR	S	R*	001130 NR
TM	R*	001180 NR	IZ	I*	001310 NR	IV	I*	001360 NR	IC	I*	001380 NR
KCOUNT	I*	001400 NR									

SOURCE STATEMENT LABELS

REQUESTED OPTIONS: NONE

OPTIONS IN EFFECT: NAME (MAIN) NOOPTIMIZE LINESCOUNT (60) SIZE (MAX) AUTODBL (NONE)

SOURCE FBCDIC NOLIS NOCHECK OBJECT MAP NOFORMAL GOSTMI NOXREF NOALC NOANSF NOTERM IBM FLAG(I) XL

```

1SN 0002      SUBROUTINE WAKE
1SN 0003      COMMON EC(20),EGP(10),EFP(10),FLP(10),FGG(100),EFG(100),ELG(100)
1SN 0004      COMMON V(100),Z(100),E(100),F(100),C(100),GR(100),ZGR(100)
1SN 0005      COMMON KS(SO),S(20),TM(100),I7(20),IV(20),IC(20),XCOUNT

```

```

C
C      CALCULATES THE RATIO OF THE WAKE TO THE BUBBLE VOLUME
C      CORRELATION USING LETAN-KEMAT APPROACH
C      SEE LITERATURE REFERENCE IN SUBROUTINE ELTAM

```

```

ISN 0006      100 CONTINUE
C             SEE IF IN RANGE OF CORRELATION
ISN 0007      IF ( E(14) .LT. TM(40)) GO TO 300
ISN 0009      IF ( E(14) .GT. TM(42)) GO TO 300
C             IT'S IN RANGE, SEE IF REGION 1 OR 2
ISN 0011      IF ( E(14) .GE. TM(41)) GO TO 250
C             LOWER RANGE
ISN 0013      200 E(56) = EXP(TM(43) + TM(44)*E(14))
ISN 0014      GO TO 400
ISN 0015      250 E(56) = EXP(TM(45) + TM(46)*E(14))
ISN 0016      GO TO 400
ISN 0017      300 E(56) = -9.87E+65
ISN 0018      400 CONTINUE
ISN 0019      IF ( E(57) .LT. 0.001) RETURN
C             IF A VALUE OF E(57) READ IN, USE IT INSTEAD
ISN 0021      E(56) = E(57)
ISN 0022      RETURN
ISN 0023      END

```

/ WAKE / SIZE OF PROGRAM 00022E HEXADECIMAL BYTES

NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.	NAME	TAG	TYPE	ADD.
C	C	R04	NH	F	SFA	C	R04	F	C	R04	NR	S	C	R04	NR
V	C	R04	NH	Z	C	R04	NH	EC	C	R04	NR	GR	C	R04	NR
IC	C	I04	NH	IV	C	I04	NR	IZ	C	I04	NR	KS	C	I04	NR
TA	FA	C	R04	EFG	C	R04	NH	EFP	C	R04	NR	EGG	C	R04	NR
EGP	C	R04	NH	ELG	C	R04	NH	ELP	C	R04	NR	EXP	F	XF	R04
ZGR	C	R04	NR	WAKF	R04	0000H0	KCOUNT	C	I04	NR					000000

***** COMMON INFORMATION *****

[illegible][illegible]

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C VECTOR SYSTEM CONSTANTS

C(1)	Run no.
C(2)	Test no. for base case data.
C(3)	
C(4)	Reactor diameter, ft.
C(6)	$g \ 32.12 \text{ ft/sec}^2$.
C(7)	$g_c \ 32.12 \text{ lb}_m\text{-ft/sec}^2\text{-lb}_f$.
C(8)	
C(9)	
C(10)	
C(11)	
C(12)	
C(13)	
C(14)	
C(15)	
C(16)	
C(17)	
C(18)	
C(19)	
C(20)	
C(21)-C(25)	Liquid name.
C(26)-C(30)	Gas name.
C(31)-C(40)	Catalyst name.
C(41)-C(50)	Problem ID.

E VECTOR CALCULATED VALUES

E(1)	Catalyst holdup, ϵ_c .
E(2)	Average gas holdup in bed, ϵ_g .
E(3)	Average liquid holdup in bed, ϵ_l .
E(4)	Average fines holdup in bed, ϵ_f .
E(8)	Slurry density, lb/ft ³ .
E(9)	Combined ($\epsilon_l + \epsilon_f$).
E(14)	Superficial liquid flow rate, ft/sec.
E(15)	Liquid linear flow in bed, ft/sec.
E(17)	Total liquid flow, gpm/ft ² .
E(18)	Gas linear velocity in bed, ft/sec.
E(20)	Gas volumetric flow, cfm.
E(21)	Old value of E(1).
E(22)	Old value of E(2).
E(23)	Old value of E(9).
E(25)	Vol% fines in liquid.
E(35)	Bubble Reynolds number.
E(36)	Bubble Morton number.
E(37)	Bubble G_2 number.
E(38)	Bubble diameter from U_{tB} (E_{si}).
E(40)	% bed expansion.
E(41)	Old value of E(1) in Bhatia-Epstein model.
E(42)	Old value of E(2) in Bhatia-Epstein model.
E(43)	Old value of [E(3) + E(4)] " " .
E(48)	Gas linear velocity in bed, ft/sec, " " .
E(49)	
E(50)	
E(51)	Single bubble terminal velocity, U_{tB} , ft/sec.
E(52)	Richardson-Zaki index, n.
E(53)	Drift flux, V_{CD} , mm/sec.
E(54)	Read-in value of U_{tB} , ft/sec.
E(55)	Read-in value of X_k .
E(56)	K_0 for wake holdup.
E(57)	Read-in value of K_0 .

F VECTOR PHYSICAL PROPERTIES

F(1)	Liquid density, lb/ft ³ .
F(2)	Liquid viscosity, cp.
F(3)	Liquid surface tension, dynes/cm.
F(5)	Slurry viscosity, cp.
F(6)	Gas density, lb/ft ³ .
F(7)	Gas viscosity, cp.
F(8)	Soaked catalyst density, lb/ft ³ .
F(9)	Bulk density, lb/ft ³ .
F(10)	Particle density, lb/ft ³ .
F(11)	Catalyst diameter, in.
F(12)	Catalyst length, in.
F(14)	Fines density, lb/ft ³ .
F(15)	Average fine size, μ .
F(17)	Ratio of solids in wake to solids in particulate phase, X_k .
F(20)	Bubble radius, ft.
F(21)	Catalyst terminal velocity, ft/sec.
F(22)	Slurry viscosity, lbm/ft-sec.
F(23)	Surface tension, lbf/ft.

KS VECTOR SWITCHES

KS(19)	Option to calculate X_k : 0--Use El-Temtamy correlation. 1--Use read-in value of X_k . 3--Calculate from U_{tB} correlation if $U_1 > U_1^*$; otherwise, use El-Temtamy correlation.
KS(30)	Maximum number of iterations for Bhatia-Epstein routine (default 20).
KS(41)	Region for Peebles/Garber correlation.
KS(42)	Iteration counter for loop in main program.
KS(43)	Maximum limit for KS(42) (default 20).
KS(44)	Read indicator: 0--Normal. 1--End of input deck.
KS(46)	First option for KS(19) to calculate X_k (default 0).
KS(47)-KS(50)	KS(19) option to calculate X_k (default -99).

TM VECTOR CORRELATION VALUES

TM(40)	U_{11}	Wake volume correlation.
TM(41)	U_{12}	" "
TM(42)	U_{13}	" "
TM(43)	a_{12}	" "
TM(44)	b_{12}	" "
TM(45)	a_{23}	" "
TM(46)	b_{23}	" "
TM(51)	Δ_1	Bubble terminal velocity correlation.
TM(52)	Δ_2	" "
TM(53)	Δ_3	" "
TM(54)	A_4	" "
TM(55)	a_{12}	" "
TM(56)	b_{12}	" "
TM(57)	a_{23}	" "
TM(58)	b_{23}	" "
TM(59)	a_{34}	" "
TM(60)	b_{34}	" "
TM(65)	U_1^*	Crossover liquid velocity, ft/sec.
TM(74)	a	X_k correlation.
TM(75)	b	X_k correlation.

V VECTOR OPERATING VARIABLES

V(5)	Fines concentration in liquid, wt%.
V(7)	Initial bed heights.
V(19)	Liquid flow rate, gpm.
V(21)	Gas superficial velocity, ft/sec.

Z VECTOR INTERMEDIATE CALCULATIONS

Z(5) $U_g - U_l$, ft/sec.
 Z(10) Relative velocity between bubble phase and liquid
 in particulate phase. (V_{g1}''') ft/sec.
 Z(11) $\epsilon_k''/\epsilon_g''$ two-phase wake/two-phase gas holdup.
 Z(12) Wake holdup (ϵ_k).
 Z(13) Liquid holdup in two-phase particulate phase.
 Z(16) New linear gas V_g velocity in Subroutine BHATIA.
 Z(17) Average linear gas velocity used as new value.
 Z(18) New value of gas holdup in BHATIA (ϵ_g).
 Z(19) New value of liquid holdup in BHATIA (ϵ_l).
 Z(20) New value of ϵ_c .
 Z(23) Value in expression for Z(13).
 Z(25) V_{g1} for El-Temtamy calculation of X_k , ft/sec.
 Z(26) X_k from El-Temtamy calculation.
 Z(27) Identical with Z(25).
 Z(28) Slope for X_k correlation.
 Z(29) Intercept for X_k correlation.

Figure E-1

Main program

Figure E-2

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Subroutine BHATIA

Solves BHATIA/Epstein model

Figure E-3

Subroutine ELTAM

(Computes X_k , the ratio of solids holdup)

Figure E-4

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APPENDIX F

CIRCULATION MODEL FOR GAS RESIDENCE
TIME DISTRIBUTION IN THREE-PHASE FLUIDIZATION

APPENDIX F

CIRCULATION MODEL FOR GAS RESIDENCE TIME DISTRIBUTION IN THREE-PHASE FLUIDIZATION

Gas flow in three-phase (gas, liquid, solid) fluidization has been shown to be generally characterized by an upflow of gas bubbles in the center region of the bed and a downflow of gas bubbles near the wall region. In order to describe the various degrees of gas mixing which may occur in each region, a circulation model consisting of n and m completely mixed tanks is proposed. This is shown schematically in Figure F-1.

Mathematical Development

A material balance at the node where the recirculation stream and input stream meet leads to the equation:

$$vC_1 + v_2C_2 = v_1C_1 \quad (1)$$

To derive the response equation to the system shown in Figure F-1, consider the transfer function in the Laplace domain for the n and m tanks, respectively (55).

$$\frac{C_o(S)}{C_1(S)} = \frac{1}{[S \frac{V_n}{nv_1} + 1]^n} \quad (2)$$

$$\frac{C_2(S)}{C_o(S)} = \frac{1}{[S \frac{V_m}{mv_2} + 1]^m} \quad (3)$$

Taking the Laplace transform of Equation 1 and rearranging:

$$\frac{v}{v_1} C_1(S) + \frac{v_2}{v_1} C_2(S) = C_1(S) \quad (4)$$

Substituting the transfer functions into Equation 4 and rearranging:

$$\frac{C_o(S)}{C_1(S)} = \frac{\frac{v}{v_1} [S\bar{t}_2 + 1]^m}{\{[S\bar{t}_1 + 1]^n [S\bar{t}_2 + 1]^m - \frac{v_2}{v_1}\}} \quad (5)$$

where $\bar{t}_1 = V_n / nv_1$ and $\bar{t}_2 = V_m / mv_2$, assuming equal sized tanks in each stream.

Letting $A = v / (v_1 \bar{t}_1^n)$ and $B = v_2 / v_1 (1 / \bar{t}_1^n \bar{t}_2^m)$ gives a more workable form of Equation 5 as:

$$\frac{C_o(S)}{C_1(S)} = \frac{A(S + 1/\bar{t}_2)^m}{[(S + 1/\bar{t}_1)^n (S + 1/\bar{t}_2)^m - B]} \quad (6)$$

Equation 6 is the general transfer function for the recirculation model shown in Figures F-1 and 36.

Consider now the analytical response equation of the system to an impulse input of tracer.

The first output signal will be for the n tanks alone--i.e., $C_2(S) = 0$:

$$C_1(S) = \frac{v}{v_1} C_i(S) \quad (7)$$

$$\frac{C_o(S)}{C_1(S)} = \frac{1}{(\bar{S}t_1 + 1)^n} \quad (8)$$

Substituting Equation 7 into Equation 8 gives the transfer function for the first output signal as:

$$\frac{C_o(S)}{C_i(S)} = \frac{v/v_1}{(\bar{S}t_1 + 1)^n} \quad (9)$$

The second output signal, corresponding to one complete loop, may be obtained by first substituting Equation 9 into Equation 3:

$$\frac{C_2(S)}{C_1(S)} = \frac{v/v_1}{(\bar{S}t_1 + 1)^n} \left\{ \frac{1}{(\bar{S}t_2 + 1)^m} \right\} \quad (10)$$

A material balance at the first node gives, $C_1(S) = 0$,

$$C_1(S) = \frac{v_2}{v_1} C_2(S) \quad (11)$$

Substituting Equation 11 into Equation 10:

$$\frac{C_1(S)}{C_i(S)} = \frac{v/v_1}{(\bar{S}t_1 + 1)^n} \left\{ \frac{v_2/v_1}{(\bar{S}t_2 + 1)^m} \right\} \quad (12)$$

Substituting Equation 12 into Equation 2 yields the transfer function for one complete loop as:

$$\frac{C_o(S)}{C_i(S)} = \frac{v/v_1}{(\bar{S}t_1 + 1)^n} \left\{ \frac{v_2}{v_1} \frac{1}{(\bar{S}t_2 + 1)^m} \frac{1}{(\bar{S}t_1 + 1)^n} \right\} \quad (13)$$

Continuing this procedure, it can easily be shown that the overall transfer function with an impulse input to the system may be generally written:

$$\frac{C_0(S)}{C_1(S)} = \frac{v}{v_1} \sum_{R=1}^{R^*} \left\{ \left(\frac{v_2}{v_1} \right)^{R-1} \frac{1}{(S\bar{t}_1 + 1)^{Rn}} \frac{1}{(S\bar{t}_2 + 1)^{(R-1)m}} \right\} \quad (14)$$

Letting $a = Rn$ and $b = (R-1)m$:

$$\frac{C_0(S)}{C_1(S)} = \frac{v}{v_1} \sum_{R=1}^{R^*} \left\{ \left(\frac{v_2}{v_1} \right)^{R-1} \frac{1}{\bar{t}_1^a \bar{t}_2^b} \frac{1}{(S + 1/\bar{t}_1)^a (S + 1/\bar{t}_2)^b} \right\} \quad (15)$$

In order to obtain the inverse function, first consider the term

$$\frac{1}{(S - a_1)^a (S - a_2)^b}$$

where $a_1 = -1/\bar{t}_1$

and $a_2 = -1/\bar{t}_2$

Now consider a Heaviside expansion formula for repeated roots (56):

$$\begin{aligned} f(S) &= \frac{1}{(S - a_1)^a (S - a_2)^b} \\ &= \frac{A_1}{(S - a_1)^a} + \frac{A_2}{(S - a_1)^{a-1}} + \dots + \frac{A_a}{(S - a_1)} \\ &\quad + \frac{C_1}{(S - a_2)^b} + \frac{C_2}{(S - a_2)^{b-1}} + \dots + \frac{C_b}{(S - a_2)} \\ &= \sum_{r=1}^a \frac{A_r}{(S - a_1)^{a-r+1}} + \sum_{r=1}^b \frac{C_r}{(S - a_2)^{b-r+1}} \end{aligned} \quad (16)$$

In order to obtain the values of the coefficients, first multiply both sides of Equation 16 by $(S - a_1)^a$:

$$\begin{aligned} f(S)(S - a_1)^a &= A_1 + A_2(S - a_1) + \dots + A_a(S - a_1)^{a-1} \\ &\quad + (S - a_1)^a \left\{ \sum_{r=1}^b \frac{C_r}{(S - a_2)^{b-r+1}} \right\} \end{aligned}$$

or:

$$f(S)(S - a_1)^a = \sum_{r=1}^a (S - a_1)^{r-1} A_r + (S - a_1)^a h_1(S) \quad (17)$$

$$\text{where } h_1(S) = \sum_{r=1}^b \frac{C_r}{(S - a_2)^{b-r+1}}$$

Differentiating Equation 17 successively $(M - 1)$ times, where $1 \leq M \leq r$, all terms in the summation from $r = 1$ to $r = M - 1$ disappear, leaving:

$$\frac{d^{M-1}}{ds^{M-1}} [(S - a_1)^a f(S)] = \sum_{r=M}^a (r - 1)(r - 2) \dots (r - M + 1) \left[(S - a_1)^{r-M} A_r \right] + \frac{d^{M-1}}{ds^{M-1}} (S - a_1)^a h_1(S)$$

Setting $S = a_1$, all terms will disappear except that in which $r = M$. In the remaining term, replace M by r :

$$A_r = \frac{1}{(r - 1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_1)^a f(S)] \Big|_{S = a_1} \quad (18)$$

where $1 \leq r \leq a$.

The coefficients C_r can be determined in a similar fashion as:

$$C_r = \frac{1}{(r - 1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_2)^b f(S)] \Big|_{S = a_2} \quad (19)$$

where $1 \leq r \leq b$.

Making use of the Laplace transform pairs:

$$\mathcal{L}^{-1} \left\{ \frac{1}{(S - a_1)^k} \right\} = \frac{t^{k-1}}{(k-1)!} e^{a_1 t}$$

$$\mathcal{L}^{-1} \{f_1(S) + f_2(S)\} = \mathcal{L}^{-1} \{f_1(S)\} + \mathcal{L}^{-1} \{f_2(S)\} = f_1(t) + f_2(t)$$

This leads to:

$$f(t) = \sum_{r=1}^a A_r \frac{t^{a-r}}{(a-r)!} e^{a_1 t} + \sum_{r=1}^b C_r \frac{t^{b-r}}{(b-r)!} e^{a_2 t}$$

Substituting Equations 18 and 19 into the above gives:

$$f(t) = e^{a_1 t} \sum_{r=1}^a \frac{1}{(r-1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_1)^a f(S)] \Big|_{S = a_1} \cdot \frac{t^{a-r}}{(a-r)!} + e^{a_2 t} \sum_{r=1}^b \frac{1}{(r-1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_2)^b f(S)] \Big|_{S = a_2} \cdot \frac{t^{b-r}}{(b-r)!}$$

The dimensionless exit age distribution may now be obtained from the inverse of Equation 15 and by noting that $E(\theta) = \bar{t}E(t)$.

$$E(\theta) = \frac{\bar{t}}{v_1} \sum_{R=1}^{R^*} \left\{ \left(\frac{v_2}{v_1} \right)^{R-1} \frac{1}{\bar{t}_1 a_2^b} \left[e^{a_1 t} \sum_{r=1}^a \frac{1}{(r-1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_1)^a f(S)] \Big|_{S=a_1} \cdot \frac{t^{a-r}}{(a-r)!} + e^{a_2 t} \sum_{r=1}^b \frac{1}{(r-1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_2)^b f(S)] \Big|_{S=a_2} \cdot \frac{t^{b-r}}{(b-r)!} \right] \right\} \quad (20)$$

$$E(\theta) = \left(\frac{\bar{t}}{1+\lambda} \right) \sum_{R=1}^{R^*} \left\{ \left(\frac{\lambda}{1+\lambda} \right)^{R-1} \frac{1}{\bar{t}_1 a_2^b} \left[e^{-\tau_1 \theta} \sum_{r=1}^a \frac{1}{(r-1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_1)^a f(S)] \Big|_{S=a_1} \cdot \frac{t^{a-r}}{(a-r)!} + e^{-\tau_2 \theta} \sum_{r=1}^b \frac{1}{(r-1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_2)^b f(S)] \Big|_{S=a_2} \cdot \frac{t^{b-r}}{(b-r)!} \right] \right\} \quad (21)$$

Method of Solution

Although Equation 20 appears to be somewhat complex, solving the general transfer function, Equation 6, numerically can be even more tedious, time-consuming, and often yields unreliable results (57,58). By making several transformations, the analytical solution developed can be easily programmed:

Consider the term:

$$\sum_{r=1}^a \frac{1}{(r-1)!} \frac{d^{r-1}}{ds^{r-1}} [(S - a_1)^a f(S)] \Big|_{S=a_1} \cdot \frac{t^{a-r}}{(a-r)!}$$

Rewrite the derivative term to give:

$$\sum_{r=1}^a \frac{1}{(r-1)!} \frac{d^{r-1}}{ds^{r-1}} \left[\frac{1}{(S - a_2)^b} \right] \Big|_{S=a_1} \cdot \frac{t^{a-r}}{(a-r)!}$$

A generalized recursion formula for the derivative term may be readily derived:

Consider the term:

$$\left. \frac{d^{r-1}}{ds^{r-1}} \left[\frac{1}{(s - a_2)^b} \right] \right|_{s = a_1}$$

For $r = 2$:

$$\left. \frac{d}{ds} \left[\frac{1}{(s - a_2)^b} \right] \right|_{s = a_1} = \frac{(-b)}{(a_1 - a_2)^{b+1}}$$

For $r = 3$:

$$\left. \frac{d^2}{ds^2} \left[\frac{1}{(s - a_2)^b} \right] \right|_{s = a_1} = \frac{(-b)[(-b + 1)]}{(a_1 - a_2)^{b+2}}$$

For $r = 4$:

$$\left. \frac{d^3}{ds^3} \left[\frac{1}{(s - a_2)^b} \right] \right|_{s = a_1} = \frac{(-b)[-b + 1][-b + 2]}{(a_1 - a_2)^{b+3}}$$

For $r = a$, then:

$$\left. \frac{d^{a-1}}{ds^{a-1}} \left[\frac{1}{(s - a_2)^b} \right] \right|_{s = a_1} = \frac{(-b)[-b + 1] \dots [-b + a - 2]}{(a_1 - a_2)^{b+a-1}}$$

Generalizing, replacing a by r leads to the general recursion formula.

$$\begin{aligned} & \left. \frac{d^{r-1}}{ds^{r-1}} \left[\frac{1}{(s - a_2)^b} \right] \right|_{s = a_1} \\ &= \frac{(-b)[-b + 1][-b + 2] \dots [-b + r - 2]}{(a_1 - a_2)^{b+r-1}} \quad \text{for } r > 1. \end{aligned}$$

A similar equation for the remaining derivative term may be written as:

$$\begin{aligned} & \left. \frac{d^{r-1}}{ds^{r-1}} \left[\frac{1}{(s - a_1)^a} \right] \right|_{s = a_2} \\ &= \frac{(-a)[-a + 1][-a + 2] \dots [-a + r - 2]}{(a_2 - a_1)^{a+r-1}} \quad \text{for } r > 1. \end{aligned}$$

For $R = 1$, corresponding to the first pass through the upflow branch, these formulas are not needed. Equation 15 will collapse to:

$$c(s) = \frac{V}{V_1} \cdot \frac{1}{t_1^n} \cdot \frac{1}{(s + 1/t_1)^n}$$

Then:

$$E(\theta) = \frac{V}{V_1} \cdot \frac{\bar{t}}{t_1^n} \cdot \frac{t^{n-1}}{(n-1)!} e^{-t/t_1}$$

which is the response of n CSTR's in series (59), modified for a volume fraction of less than one.

Figure F-2 shows the flow diagram for the computational procedures based on the proposed model. Calculations of the exit age distribution function begin with specifications of n , m , λ , and P_1 . Assuming that the n tanks are all of equal size, V_n/n , and that the m tanks are all of equal size, V_m/m , the following relationships can be easily shown to hold:

$$\frac{\bar{t}_1}{\bar{t}_2} = \frac{P_1}{P_2} \left(\frac{m}{n} \right) \left(\frac{\lambda}{1 + \lambda} \right)$$

$$\bar{t} = \bar{t}_1 n (1 + \lambda) + \bar{t}_2 m \lambda$$

where: $P_1 = V_n/V$
 $\frac{P_2}{P_1} = \frac{V_m/V}{1 - P_1}$

The exit age distribution is calculated from Equation 21 and the appropriate recursion formula from the derivatives. Once the exit age distribution for each output has been determined, a summation over all outputs is made to give the final exit age distribution for the specified parameters. The Fortran program used for the calculation appears in Table F-1.

Figures F-3 and F-4 show some typical response curves obtained from the model. For most regions of interest, the initial response of such a system is controlled by the number of tanks in the upflow stream and the upflow volume fraction. The influence of the recycle branch will largely affect the later decay of the response signal.

A comparison may be made between this model and that recently published by Mann, et al. (58). Figure F-5 contrasts the model predictions (note that different definitions of recycle ratio are used). The agreement is excellent, indicating that the two approaches--summation by convolution in the time domain, or Heaviside expansion in the Laplace domain--yield equivalent solutions.

Application to Experimental Radiotracer Data

This model was then used to describe flow patterns observed during the radiotracer tests. Figure F-6 shows the results of two tests conducted with no coal fines present in the kerosene. At low gas rates, the upflow branch consists of five CSTR's in series; this flow regime is intermediate between backmix and plug flow. A small amount of recycle is present through the one downflow CSTR. Increasing the superficial gas velocity from 0.10 ft/sec to 0.20 ft/sec decreased the number of tanks from five to two, and also slightly diminished the volume fraction occupied by the upflow stream.

The addition of fines to the kerosene changed the residence time distribution as indicated in Figure F-7. For the two experiments shown, the behavior is closely represented by one backmix reactor with a small recycle probability (0.05-0.10). The apparent time delay of 0.2θ (corresponding to 2-3 seconds) may be related to the rise time for a large gas bubble. This is especially probable for Test 15; a better fit to the response curve could be made by assuming the initial peak of $E(\theta)$ to be a bubble leaving the bed; other gas remaining in the bed would then recirculate to produce the dispersed response seen experimentally.

NOMENCLATURE FOR APPENDIX F

a	Rn
a_1	$-1/\bar{t}_1$
a_2	$-1/\bar{t}_2$
A	$v/(v_1 \bar{t}_1^n)$
A_r	First term coefficients in partial fraction expansion
b	$(R - 1)m$
B	$(v_2/v_1)(1/\bar{t}_1^n \bar{t}_2^m)$
C_0	Outlet tracer concentration from system (arbitrary units)
C_1	Inlet tracer concentration to the n tanks (arbitrary units)
C_2	Outlet tracer concentration from the m tanks (arbitrary units)
C_r	Second term coefficients in partial fraction expansion
$E(t)$	Exit age distribution function (1/time)
$E(\theta)$	Dimensionless exit age distribution function
m	Number of tanks in downflow region
n	Number of tanks in upflow region
P_1	Volume fraction in upflow stream
P_2	Volume fraction in downflow stream
R	Output signal number
R^*	Maximum number of cycles
$\bar{t}$	Mean residence time, $V/$
$\bar{t}_1$	Mean residence time in the n th tank
$\bar{t}_2$	Mean residence time in the m th tank

NOMENCLATURE FOR APPENDIX F

-2-

v	Total gas volumetric flow rate
v_1	Gas volumetric flow rate in upflow stream
v_2	Gas volumetric flow rate in downflow stream
V	Total gas volume in system
V_m	Total volume of the m tanks
V_n	Total volume of the n tanks

Greek

θ	Dimensionless time
λ	Recycle ratio
$\bar{\theta}_p$	Initial delay in plug flow element

TABLE F-I

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MAIN

DATE = 80032

10/12/00

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```

0001      IMPLICIT REAL*(A-H, O-Z)
0002      DIMENSION Z(115, 700)
-----
0003      C INPUT PARAMETERS FOR CIRCULATION MODEL
0004      99 IFLAG3=1
0005      N=1
0006      P1=0.700
0007      RLAMB=0.100
0008      WRITE(6, 100) N, 4, RLAMB, P1
0009      100 FORMAT(' THE INPUT PARAMETERS ARE: /, N= ', I3, ' M= ', I3,
1      ' LAMBDA= ', F7.4, ' P1= ', F7.4)
-----
0010      C INPUT MEAN RESIDENCE TIME FROM EXPERIMENTS
0011      TBAR=51.000
0012      C COMPUTE MODEL PARAMETERS FROM INPUT PARAMETERS
0013      P2=1.000-0.1
0014      T2BAR=TBAR/(M*RLAMB*((P1/P2)+1.000))
0015      T1BAR=T2BAR*(P1/P2)*(FLOAT(M)/FLOAT(N))*(RLAMB/(RLAMB+1.000))
0016      RT1=1.000/T1BAR
0017      RT2=1.000/T2BAR
0018      TAU1=TBAR/T1BAR
0019      TAU2=TBAR/T2BAR
0020      COEF=(1.000/(1.000+RLAMB))*TAU2
0021      PT=3.1415900
0022      L=2
0023      C BEGIN EXTERNAL SUMMATION
0024      DO 4 I=1, L
0025      C SET THETA
0026      THETA=0.000
0027      K=1
0028      INA=0.1
0029      INB=4*(I-1)
0030      IF(I.EQ.1) GO TO 7
0031      Z=((RLAMB/(1.000+RLAMB))*((I-1))*(1.000/((T1BAR**INA)*(T2BAR**
1      INB))))
0032      IF(I.GT.1) GO TO 9
0033      7 Z=1.000/((T1BAR**INA)
0034      8 U1=1.000/DEXP(THETA*TAU1)
0035      C IF I EQUAL TO 1 SKIP FIRST INTERNAL SUMMATION
0036      IF(I.EQ.1) GO TO 51
0037      C COMPUTE FIRST INTERNAL SUMMATION
0038      SUM01=0.000
0039      NUM10=1
0040      DO 9 J=1, INA
0041      JN1=J-1
0042      IF(J.EQ.1) GO TO 10
0043      CALL IFACT(JN1, JN1F)
0044      IF(J.GT.1) GO TO 11
0045      10 JN1F=1
0046      11 IF(J.EQ.1) GO TO 12
0047      NUM1N=JN1N*(JN1F-1)
0048      NUM1N=NUM1N*NUM10
0049      DERS1=NUM1N/((RT2-RT1)**(INB+J-1))
0050      IF(J.GT.1) GO TO 14
0051      12 NUM1N=1
0052      DERS1=NUM1N/((RT2-RT1)**INB)
0053      14 INANJ=INA-J
0054      IF(J.EQ.INA) GO TO 15
0055      CALL IFACT(INANJ, INANF)

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TABLE F-I

-2-

FORTRAN IV G1 RELEASE 2.0

MAIN

DATE = 93032

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0050      V1=(THETA*TRAR)**INAMJ
0051      IF(J.LI.INA) GO TO 16
0052      INAMF=1
0053      V1=1.000
0054      16 SUM41=(1.000/J41F)*DERS1*(V1/INAMF)
0055      SJ41=SUM41+SJ41
0056      SJ41=SUM41
0057      NU410=NU41N
0058      9 CONTINUE
      C COMPUTE SECOND INTERNAL SUMMATION
0059      SUM42=0.000
0060      NU420=1
0061      DO 17 J=1,INR
0062      JM1=J-1
0063      IF(J.EQ.1) GO TO 18
0064      CALL IFACT(JM1,JM1F)
0065      IF(J.GT.1) GO TO 19
0066      18 JM1F=1
0067      19 IF(J.EQ.1) GO TO 20
0068      NUM2N=(-(INA+J-2))
0069      NU42N=NU42N*NU420
0070      DERS2=NUM2N/((RT1-RT2)**(INA+J-1))
0071      IF(J.GT.1) GO TO 21
0072      NUM2N=1
0073      DERS2=NU42N/((RT1-RT2)**INA)
0074      21 INRMJ=INR-J
0075      IF(J.EQ.INR) GO TO 22
0076      CALL IFACT(INRMJ,INRMF)
0077      V2=(THETA*TRAR)**INRMJ
0078      IF(J.LI.INR) GO TO 23
0079      22 INRMF=1
0080      V2=1.000
0081      23 SUMN2=(1.000/J41F)*DERS2*(V2/INRMF)
0082      SUMN2=SUMN2+SUM02
0083      SUM02=SUMN2
0084      NUM20=NUM2N
0085      17 CONTINUE
      C COMPUTE EXTERNAL SUMMATION
0086      U2=1.000/DEXP(THETA*TAU2)
0087      SUMTN=2*((U1*SUMN1)+(U2*SUMN2))
0088      IF(I.GT.1) GO TO 52
0089      51 NM1=N-1
0090      IF(N.EQ.1) GO TO 66
0091      CALL IFACT(NM1,NM1F)
0092      SUMTN=2*J1*((THETA*TRAR)**NM1)/NM1F
0093      IF(N.GT.1) GO TO 52
0094      66 SUMTN=2*J1
0095      C COMPUTE E(THETA)
      52 E(I,K)=CDEF*SUMTN
      C INCREMENT THETA
0096      THETA=THETA+0.0100
0097      K=K+1
0098      IF(THETA.LE.6.0) GO TO 8
0099      C PRINT OUT E(THETA) FOR EACH OUTPUT
      WRITE(6,100) (E(I,K),K=1,600)
0100      100 FORMAT(15F7.4,15F7.4)
0101      C GO TO NEXT OUTPUT
      A CONTINUE

```

TABLE F-I

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C CALCULATE AND PRINT TOTAL OUTPUT SIGNAL FOR L TERMS AT EACH THETA

```

0102      K=1
0103      SUMED=0.000
0104      DO 33 I=1,L
0105      SUMEN=((I,K)
0106      SUMED=SUMEN+SUMED
0107      SUMED=SUMED
0108      33 CONTINUE
0109      WRITE(6,1)5K,SUMEN
0110      105 FORMAT(1X,F7.4)
0111      K=K+1
0112      IF(K.LE.600) GO TO 40
0113      STOP
0114      END

```

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TABLE F-1

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```
0001 SUBROUTINE IFACT(IX,IXF)
0002 IF(IX.EQ.1) IXN=1
0003 IF(IX.EQ.2) IXN=2
0004 IF(IX.LT.3) GO TO 94
0005 IXN=IX
0006 IX41=IX-1
0007 D1 92 I1=1,IX41
0008 IXN=IX*(IX-11)
0009 IXN=IXN
0010 92 CONTINUE
0011 96 IXF=IXN
0012 RETURN
0013 END
```

Figure F-1
Gas mixing model

Figure F-2

FLOW DIAGRAM FOR
COMPUTATION PROCEDURE

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Figure F-3
Typical response curves based on the proposed model

Figure F-4

Typical response curves based on the proposed model

Figure F-5
Comparison of Laplace transform and
integral models

Figure F-6

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Comparison of gas mixing model with experiments: Kerosene/catalyst/no fines

Figure F-7

**Comparison of gas mixing model with experiments:
Kerosene/catalyst/15.5 vol % coal fines**

APPENDIX G
MODEL DEVELOPMENT

APPENDIX G

MODEL DEVELOPMENT

I) Correlations for Wake Volume Ratio, K_o

$$K_o = \exp(a_{12} + b_{12} U_1) \quad U_{11} < U_1 < U_{12}$$

$$K_o = \exp(a_{23} + b_{23} U_1) \quad U_{12} \leq U_1 \leq U_{13}$$

Run	211	212	423
Fluid	Kerosene	Kerosene	Mineral Oil
Temperature, °F	80	71	175
% Fines	17.8	0	0
U_{11} , Ft/Sec	0.0001	0.0001	0.0001
U_{12} , "	0.5000	0.5000	0.1496
U_{13} , "	0.5001	0.5001	0.5000
a_{12} , Sec/Ft	-0.03074	0.5441	0.13995
b_{12} , "	8.55736	0.0	6.56681
a_{23} , "	--	--	1.28003
b_{23} , "	--	--	-1.05323

II) Correlation of Bubble Terminal Velocity

Define $\Delta = U_g - U_1$.

$$U_{tB} = a_{12} + b_{12} \Delta \quad \Delta_1 \leq \Delta \leq \Delta_2$$

$$= a_{23} + b_{23} \Delta \quad \Delta_2 \leq \Delta \leq \Delta_3$$

$$= a_{34} + b_{34} \Delta \quad \Delta_3 \leq \Delta \leq \Delta_4$$

Run	211	212	423
Fluid	Kerosene	Kerosene	Mineral Oil
Temperature, °F	80	71	175
% Fines	17.8	0	0
Δ_1 , Ft/Sec	-99	-99	-99
Δ_2 , "	-0.0476	-0.02190	-0.04280
Δ_3 , "	0.1254	100	0.00534
Δ_4 , "	100	--	100
a_{12}	0.259	0.24652	0.10885
b_{12}	0	-1.45387	-2.54940
a_{23}	0.6005	0.26270	0.27151
b_{23}	7.1713	-0.71516	1.25074
a_{34}	1.50	--	0.25183
b_{34}	0	--	4.93799

III) Correlation of X_k Relative Solids Holdup in Wake

$$X_k = 1 - \frac{1}{(a + b U_{tB})} \quad \text{if } U_1 \geq U_1^*$$

$$X_k = 1 - 0.877 \frac{U_t}{\frac{U_g}{\epsilon_g} - \frac{U_1}{\epsilon_1}} \quad \text{using predicted } \epsilon_g, \epsilon_1 \text{ if } U_1 \leq U_1^*$$

Run	211	212	423
Fluid	Kerosene	Kerosene	Mineral Oil
Temperature, °F	80	71	175
% Fines	17.8	0	0
U_1^* , Ft/Sec	0.135	100	0.095
a	0.5038	--	0.36174
b, Sec/Ft	2.582	--	2.83710