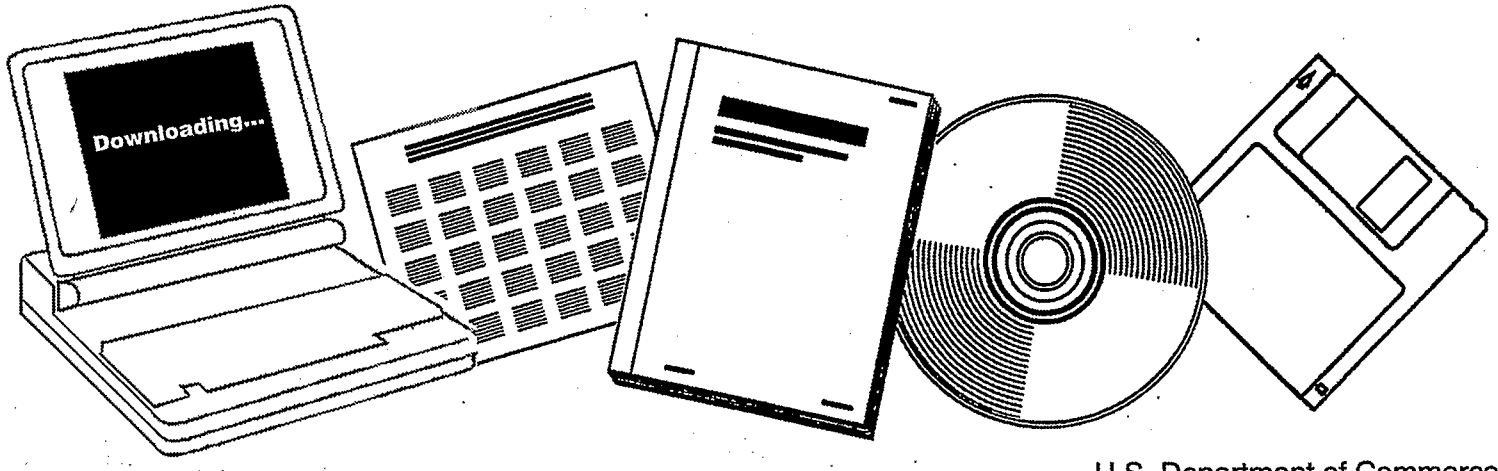
SLURRY FISCHER-TROPSCH/MOBILE TWO-STAGE PROCESS OF CONVERTING SYNGAS TO HIGH-OCTANE GASOLINE. QUARTERLY REPORT, 1 APRIL-30 JUNE 1981

MOBIL RESEARCH AND DEVELOPMENT CORP.
PAULSBORO, NJ

JUL 1981



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SLURRY FISCHER-TROPSCH/MOBIL TWO-STAGE PROCESS OF CONVERTING SYNGAS TO HIGH OCTANE GASOLINE

QUARTERLY REPORT FOR THE PERIOD
1 APRIL - 30 JUNE, 1981

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I. Abstract

The fabrication of major Bench-Scale Unit (BSU) components in the machine shop is essentially complete. Some refinements of BSU design have been made.

A simple mathematical model of the F-T bubble-column reactor was used to assist the planning of the operation of BSU. Three effects were examined, i.e., liquid-phase axial dispersion, axial catalyst distribution, and Space-Time-Yield as a variable of the catalyst loading and feed superficial gas velocity.

II. Objective and Scope of the Project

The overall objective of the contract is to develop a two-stage slurry Fischer-Tropsch/ZSM-5 process for direct conversion of syngas, of the type produced in a coal gasification system, to high octane gasoline. The immediate objective is to design, construct, and operate a bench-scale pilot plant so that the economic potential of this process concept can be evaluated. To accomplish these objectives, the following specific tasks will be undertaken:

Task 1 - Design of Bench-Scale Pilot Plant

A two-stage slurry F-T/ZSM-5 bench-scale pilot plant will be designed for conversion of syngas to high octane gasoline. The slurry F-T reactor will be 2" ID and 25' high. The fixed-bed ZSM-5 reactor will be 2" ID and 4-18" high. A distillation column will be designed to obtain stabilized gasoline products.

Task 2 - Construction and Shakedown of Pilot Plant

The pilot plant will be constructed in MRDC Paulsboro Laboratory. The unit will be shaken down when completed.

Task 3 - Operation of Pilot Plant

At least three slurry F-T catalysts will be tested in the bench-scale pilot plant. One of these catalysts may be provided by DOE's alternate catalyst development projects. The best first-stage catalyst together with a ZSM-5 class zeolite catalyst will be used for process variable studies and catalyst aging tests in the bench-scale unit. Products obtained from the unit will be evaluated to define their qualities.

Task 4 - Conceptual Design Study

A preliminary conceptual design of the process will be developed for a commercial size plant for the conversion of syngas to high octane gasoline. Scoping costs of the plant will be estimated.

III. Summary of Progress to Date

The construction phase of the BSU is progressing satisfactorily. The fabrication of major BSU components in the machine shop is essentially complete. The on-site construction of the BSU will be initiated soon. Some refinements of the BSU design were also undertaken during this period. Several methods for measuring the liquid-level and gas hold-up in the BSU bubble-column reactor were evaluated, and a method of measuring the axial pressure-drop in a bubble-column using pneumatic differential-pressure cells was adopted.

A simple mathematical model of the F-T bubble-column reactor was used to assist in planning the operation of the BSU F-T reactor. Three studies were concluded:

- Based on a preliminary study, the effect of the liquid phase axial dispersion in the BSU F-T reactor performance is expected to be very small. A non-mixing liquid phase mathematical model may be a good representation of the BSU F-T reactor operation.

- The effect of the axial catalyst distribution on the BSU F-T reactor performance was examined using a non-mixing liquid phase mathematical model coupled with an axial catalyst dispersion model. The catalyst size is the most important parameter. When the catalyst size is less than 40μ, less than 15% longer reactor length than that required for a reactor with a uniform catalyst distribution is needed to achieve a 90% hydrogen conversion. This may be acceptable. However, it is uncertain what the effect of the catalyst size is on catalyst pretreatment (the reduction of iron oxides to iron carbides).

- The Space-Time-Yield, i.e., gMole syngas converted/hr-mL expanded slurry, of the BSU F-T reactor was examined using a non-mixing liquid phase mathematical model. As expected, the STY depends strongly on the catalyst loading if one assumes the same hydrodynamic parameters for all catalyst loadings. The STY reaches an optimum with varying gas velocities. For catalyst loadings between 5-15 wt % Fe in the slurry, the feed gas superficial velocities at which the maximum STY occurs vary from 2.8 to 4.6 cm/s.

IV. Detailed Description of Technical Progress

A. Task 2 - Construction and Shakedown of Pilot Plant

1. Status of Task

As of this date, the fabrication of most BSU components -- preheaters, condensers, receivers, dropout pots, and fractionation columns -- is essentially complete. The construction of the two fixed-bed reactors and the slurry reactor is underway.

Drawings for the steel structure to house the slurry reactor have been finished, and the on-site construction of the structure will begin soon. The unit-arrangement drawings along with the drawings for the Unistrut-assemblies (steel structures) to mount BSU components are complete for two of the four sections; these have been sent to the machine-shop for fabrication.

2. Continual Refinements of Bench-Scale Pilot Plant Design

As was reported in the last Quarterly Report, the detailed designs for both the F-T and the ZSM-5 reactors were complete and the preliminary fabrication drawings for the slurry F-T reactor were attached. Continual refinements of these designs were done in this quarter and the updated fabrication drawings for both reactors are attached (Figures 1-3). On the drawings of the slurry F-T reactor (Figures 1 and 2), all physical dimensions and details of the filter, viewing ports on the bubble-column, the feed distributor plate, and supports for the reactor are now given. Figure 3 gives details of the fixed-bed ZSM-5 reactor. The reactor is 40" long with 1.939" inside-diameter. The maximum catalyst bed height is 23" (capacity - 68 $inch^3$), with the top 15" section acting as a preheater.

The liquid level and gas holdup in the slurry F-T reactor are important process variables. Considerable effort was spent earlier to evaluate methods for such measurements, i.e.:

- Axial pressure-drop measurment using pneumatic differential-pressure (DP) cells.
- Axial pressure-drop measurement using piezoresistive pressure transducers.
- Liquid level height measurement using a floating radioactive-source.

The decision has now been made to use the first method after thorough discussions with instrumentation experts and consultants. This method has the advantages of simplicity in both use and maintenance. The method of pressure-drop measurement using Peizo-resistive pressure-transducers has been ruled out because of difficulties with water-cooling to maintain the transducers at temperatures below a damaging 450°F. The method of measuring the liquid level height using a floating radioactive source in the bubble-column has been eliminated because the use of radioactive materials is cumbersome and does not give any indication of the axial profile of the gas holdup.

Figure 4 shows the schematic arrangement of a DP-cell with six N_2-purge lines along the reactor height to measure the pressure gradient. The pressure-drop between any two DP-cell lines, which gives the reactor pressure-drop between these two locations, can be measured by connecting these lines to a differential-pressure cell (DP-cell). In the Figure, Lines 1 and 6 are shown connected to the DP cell. Also shown is a pathway to connect Lines 1 and 4 to the DP cell.

A continuous N_2-purging through the lines is necessary to keep them free of any slurry from the reactor which may plug them. Based on a laboratory experiment using hexadecane in a glass bubble-column, a purge rate of 12 mL/min (at actual temperature and pressure) is sufficient. To be safe, a purge rate of 30 mL/min (actual) is recommended. The total N_2 purge rate will be less than 4% of the flow rate of syngas entering the reactor; thus, performance will not be affected.

Gas holdup is related to the densities of the slurry ($\rho_{s\ell}$), the expanded slurry ($\rho_{es\ell}$) and the gas (ρ_g) by the following equation:

$$\epsilon_g = (\rho_{s\ell}-\rho_{es\ell})/(\rho_{s\ell}-\rho_g) \qquad (1)$$

ρ_g and $\rho_{s\ell}$ can be easily calculated. $\rho_{es\ell}$ can be estimated from the measured pressure-drop between any two locations within the expanded slurry. For example, if the slurry level is between 15' and 20' levels, the pressure drop between Locations 1 and 4 can be used to estimate the average $\rho_{es\ell}$ between these two locations. In general, between Locations 1 and N:

$$\rho_{es\ell} = 2.309\ (P_N-P_1)/(\ell_N-\ell_1) \qquad (2)$$

where 2.309 is the factor for converting pressure (psi) into a hydraulic head (ft of water), P_i is the pressure at location i (psi), ℓ_i is the height of the reactor column at location i (Ft), and $\rho_{es\ell}$ is the average density of the expanded slurry (g/mL). An equation similar to Equation (2) can be used to calculate the average expanded slurry density between any two adjacent locations. An axial profile of the gas holdup can then be established. If the axial variation of the gas holdup is small,

the density calculated from Equation (2) can be used for the whole slurry column; otherwise, the axial variation of the bed density must be taken into account. Using the same example, the liquid level in the slurry bubble-column can be calculated using the following equation:

$$\ell = \ell_4 + (\ell_4 - \ell_1)(P_6 - P_4)/(P_4 - P_1) \qquad (3)$$

if the axial variation of the gas holdup is nil. With appreciable axial variation of the gas holdup, the axial profile can be taken into account to obtain a more accurate estimated of the liquid level.

Based on the information supplied by the DP-cell vendor, the absolute error of pressure drop measurements is 0.05 psi for a range of 10 psi. This translates to a maximum error of 3% for the average gas holdup calculation when the liquid level is above 10' and a maximum error of 6% in the liquid level estimate.

2. Conclusion

The fabrication of BSU components in the machine shop is essnetially complete and is on schedule. Continual refinements of BSU designs were undertaken. Updated fabrication drawings for both the slurry F-T and ZSM-5 reactors are included. A method of axial pressure drop measurement in the slurry F-T reactor using pneumatic differential-pressure cells was adopted for liquid level and gas holdup estimates.

3. Future Work

- The fabrication of the reactors will be completed in the next quarter.
- The on-site construction of the steel-structure to house the slurry reactor will begin soon.
- The fabrication and on-site construction of Unistrut assemblies to mount other BSU components will be initiated.

B. Task 3 - Operation of Pilot Plant

1. Fischer-Tropsch Bubble-Column Mathematical Model and Its Applications

a. Effect of Liquid Phase Axial Dispersion on BSU F-T Reactor Performance

Calculations in the last Quarterly Report (Section IV. A.3.b.) have shown that the state of the liquid phase axial mixing has a large effect on slurry F-T reactor performance.

That conclusion was drawn from calculations using two extreme state of the liquid phase, i.e., non-mixing (NM model) and perfect mixing (PM model). The actual state of the axial liquid mixing lies somewhere between these two extremes. The purpose of the current work is to investigate the effect of this liquid phase axial dispersion on BSU F-T reactor performance using a physical model similar to the one adopted by Deckwer, et.al. (1980) in which an axial dispersion coefficient is used.

Other than the axial mixing term in the liquid phase, the mathematical equations used follow exactly those given in the past Quarterly Report with one minor exception. The exception is that the molar contraction in the gas phase resulting from the F-T reactions is approximated by an average value given as a linear average of the values at the reactor entrance and exit. This approximation is a convenient way to investigate this axial mixing effect without using the complicated numerical computation method, since the resulting equations and boundary conditions can be solved analytically. The differential equations for this model were given in the first Quarterly Report (Equation 2) with the following expression for the axial dispersion in the liquid phase to be added to the RHS of Equation (2) there:

$$-A(1-\epsilon_g)(1-v)E_\ell \frac{d^2C_{H\ell}}{d\ell^2} d\ell \qquad (4)$$

where E_ℓ is the liquid phase axial dispersion coefficient (cm^2/s). A correlation of E_ℓ is recommended by Deckwer, et.al. (1980):

$$E_\ell = 3.676\, U_{gm}^{.32} d_r^{1.34} \qquad (5)$$

It is not clear what are the ranges of the parameters in which this correlation applies.

The BSU F-T bubble-column reactor has an inside diameter of 2" and a height of 10-25'. Using the axial dispersion model mentioned in the preceeding paragraph, calculations were done on the hydrogen conversion versus the reactor height for the NM liquid-phase case and the liquid phase axial dispersion case, as shown in Figure 5. The superficial gas velocity used in this calculation is 2 cm/s, the expected minimum velocity for BSU F-T reactor operation. The maximum deviation of the liquid phase axial dispersion effect from the description of the NM case is expected at this minimum gas velocity. As shown in Figure 5, the result from the NM case is practially identical with that of the liquid phase axial dispersion case for the BSU F-T reactor. Note that in order to make a proper comparison of these two cases, the same approximation of gas molar contraction, as mentioned earlier, was also used in the NM liquid-phase case. As a comparison, a curve for the PM liquid-phase case is also

included in the figure.

b. Effect of Axial Catalyst Distribution on BSU F-T Reactor Performance

In the F-T slurry process, fine catalyst particles are suspended in the liquid phase by the bubbling of the syngas. This uplifting force is balanced by the gravitational force on the particles. Therefore, the axial catalyst distribution is generally non-uniform. This non-uniform distribution will lower reactor performance. The objective of this study is to evaluate this effect on BSU F-T rector performance using a catalyst dispersion model coupled with the slurry F-T reactor mathematical model described in the last two Quarterly Reports. Specifically, the most important variable that affects catalyst distribution is the catalyst size. If the size is small enough, the axial catalyst distribution will be relatively uniform and good reactor performance can be ensured. The primary objective of this study is to determine the maximum catalyst size such that the deviation of reactor performance due to non-uniform axial catalyst distribution will be insignificant. Furthermore, Farley and Ray (1964), Schlesinger, et.al. (1954), and Koelbel and Ralek (1980) reported that F-T Fe-based catalysts disintegrate during normal operation. The former two reported that catalysts disintegrate to 1-3μ size. A secondary objective of this study is to insure that such stabilized catalyst size is small enough to ensure proper operation. It is expected that the satisfaction of the primary objective will automatically satisfy the secondary objective. For the present study, the NM liquid phase mathematical model is used.

The equation describing the axial catalyst concentration profile is given:

$$E_c \frac{dC_c'}{d\ell} + u_{cs} C_c' = 0 \qquad (6)$$

Catalyst Axial Dispersion — Catalyst Settling

This equation has the simple physical interpretation that there is no net catalyst movement at any axial position in the bubble-column. This catalyst concentration profile is constrained by the average catalyst loading:

$$\left(\int_0^L C_c'(\ell)d\ell\right)/L = C_{ca}' \qquad (7)$$

The solution of Equations (6) and (7) is:

$$C_c'(\ell)/C_{ca}' = Pe_c \exp(-Pe_c\, \ell/L)/(1-\exp(Pe_c)) \qquad (8)$$

The same solution was also obtained by Cova (1966) and

Deckwer, et.al. (1980), although they used a second-order differential equation and boundary conditions instead of Equations (6) and (7). This model describes very well experimental data, all for the air-water system, containing catalyst sizes of up to 180μ (Kato, et.al., (1972)). In treating these experimental data, a mean particle diameter defined as:

$$d_c = 6 \text{ (Volume of Particles)/Surface Area of Particles} \quad (9)$$

is used.

The equations that describe the hydrogen balance in an F-T bubble-column reactor with an axial NM liquid phase were given in the first Quarterly Report (Equation 2). For the present application, the only difference is that the kinetic rate is now proportional to $C_c'(\ell)$, which is given in Equation (8). The equations are, for the gas phase:

$$N_H^i \, dX_H = A \, k_\ell \, a_g (C_{H\ell}^* - C_{H\ell}) d\ell \quad (10)$$

and for the liquid phase:

$$A \, k_\ell \, a_g (C_{H\ell}^* - C_{H\ell}) d\ell = A \, r_H \, d\ell \quad (11)$$

where

$$r_H = k_H'' (1-\epsilon_g) f_{Fe} \, C_{H\ell} \, C_c'(\ell) \quad (12)$$

By substituting Equation (12) into Equation (11), Equations (10) and (11) can be solved in a way similar to that used to obtain the solution for the NM liquid phase case with uniform catalyst distribution as given in the first Quarterly Report (Section IV. A.3.d.). The solution is given implicitly as

$$L = -u_g^i \, R_d (\alpha^* \, X_H^e + (1+\alpha^*) \ln(1-X_H^e)) - E_c (\ln g(L))/u_{cs} \quad (13)$$

where

$$g(L) = (B_1 \, L + B_2(1-\exp(-B_1 \, L)))/(B_1 \, L - B_2(1-\exp(B_1 \, L))) \quad (14)$$

$$B_1 = u_{cs}/E_c \qquad B_2 = R_k/R_d \quad (15)$$

Table 1 gives the values of the parameters that were used in the current calculations. The correlations used to calculate the liquid and gas parameters are the same as those used in the last Quarterly Report. ϵ_g and a_g are determined at average gas velocity:

$$u_{gm} = u_g^i (1+0.5\alpha^* \, X_H^e) \quad (16)$$

The correlations used to estimate u_{cs} and E_c are those obtained

by Kato, et.al. (1972) from fitting particle settling data over a wide range of catalyst sizes and loadings, gas velocities, and reactor diameters for the air-water-glass beads system. These correlations are summarized below:

$$E_c = u_{gm}\, d_R (1+8\, Fr^{.85})/13\, Fr \qquad (17)$$

$$u_{cs} = 1.2\, u_{ct} (u_{gm}/u_{ct})^{.25} ((1-v)/(1-v^*))^{2.5} \qquad (18)$$

where

$$u_{ct} = \mu_\ell\, Ar/18\, \rho_\ell\, d_c \qquad (19)$$

$$v^* = v \text{ as } C_c' = 0.1 \qquad (20)$$

A preliminary study of this axial catalyst dispersion effect shows that the deviation of the F-T bubble-column reactor performance from that with a uniform catalyst distribution is the largest when:

1. The catalyst loading is low.

2. The feed gas superficial velocity is high.

Consequently, to study the same effect on the BSU F-T reactor performance, the maximum deviation is expected at the lowest catalyst loading of about 5 wt% Fe and the highest feed gas superficial velocity of about 7 cm/s. The effect of a non-uniform catalyst distribution can be represented as the percentage increase in reactor length required to achieve the same hydrogen conversion as that estimated using a uniform catalyst distribution. These results are given in Figure 6 for catalyst sizes from 10 to 50μ. If one considers that a 15% longer reactor length is an acceptable deviation, then the acceptable catalyst size is below 40μ for 90% hydrogen conversion. Note that this catalyst size-limit is not related to catalyst pretreatment. Only proper experiments can determine if the same size-limit is applicable to catalyst pretreatment.

In open literature, many sizes of the fresh Fe-based catalysts have been mentioned. However, the slurry reactor diameters and the feed gas superficial velocities are not exactly those used in the current study. Koelbel and Ralek (1980) mentioned a 30μ catalyst for a 2" diameter bench-scale unit at a 3.5 cm/s feed gas superficial velocity, and the same size catalyst for the Rheinsprussen demonstration plant (1.29m reactor and 9.5 cm/s feed gas superficial velocity).
Schlesinger, et.al. (1951) used catalyst smaller than 60μ for a 3" diameter reactor and a feed gas superficial velocity of about 2.5 cm/s. Sakai and Kunugi (1974) used a 1μ catalyst for a 2" diameter reactor and a 3.8 cm/s feed gas superficial velocity.

As mentioned in the first paragraph of this sub-section, catalyst disintegration to sizes below 5μ may be expected during normal operation. Since this size is much smaller than the 40μ limit recommended for the fresh catalyst, no operational difficulties due to non-uniform catalyst distribution are expected.

c. Investigation of Space-Time-Yield of BSU F-T Reactor

One of the questions often raised in the discussion of the F-T reactor performance is the possible disadvantage due to the inherently low catalyst density in slurry reactors. In the other words, it requires a larger reactor volume to hold the same amount of catalyst than the conventional vapor-phase F-T reactors do. This larger reactor volume may pose a penalty as a higher cost for the final product. Consequently, it is important to examine the Space-Time-Yield behavior of a F-T slurry reactor in order to search for an optimal STY operation. The Space-Time-Yield used here is defined as:

$$\text{STY} = \text{gMol Syngas Converted/Hr-mL Expanded Slurry} \qquad (21)$$

Based on the current mathematical model, the yield is strongly dependent on the catalyst loading and the gas velocity. Using the NM mathematical model, Figure 7 shows results of calculations for the BSU F-T reactor operation. Only the results for the 90% hydrogen conversion are given. As expected, the STY depends strongly on the catalyst loading. Note that, in the current mathematical model, the hydrodynamic properties of the column are assumed to be independent of the catalyst loading. Deckwer, et.al. (1980) has mentioned that this independence exists up to 16 wt% of catalyst in the slurry. However, Koelbel and Ralek (1980) has indicated that the optimum catalyst loading is about 10 wt% in terms of the iron in the slurry. Higher catalyst loading increases the viscosity of the slurry and thus decreases the gas-liquid interfacial area. Based on these two references, the question of the optimal catalyst loading will need further investigation. The calculated results for the high catalyst loadings shown in Figure 7 can only be used as a guide for future studies.

The dependence of the STY on the gas velocity is very interesting because maximum STY's exist for each curve. The physical interpretation of this phenomenon is that, at low gas velocity, the STY is low because the gas-liquid interfacial area is low; while, at high gas velocity, the STY also became lower because the slurry-bed expansion becomes an overriding factor. In Figure 7, a dotted line indicates the locus of the maximum STY. For catalyst loadings between 5-15 wt% Fe, the feed gas superficial velocities at which the maximum STY occurs vary from 2.8 to 4.6 cm/s. Caution should be given in interpreting the results at the high end of the gas velocity, since, according to

Deckwer, et.al. (1980a), the flow in the BSU F-T column may approach "slug flow" regime at about 7 cm/s gas velocity.

2. Conclusions

The simple mathematical model of the F-T bubble-column reactor has been used to assist the planning of the operation of the BSU F-T reactor. Specifically, three studies were done:

- A preliminary study shows that the effect of the liquid phase axial dispersion on the BSU F-T reactor performance is very small. An NM liquid phase mathematical model may be a good representation of the BSU F-T reactor operation.

- The effect of the axial catalyst distribution on the BSU F-T reactor performance was examined using NM liquid phase mathematical model coupled with an axial catalyst dispersion model. The catalyst size is the most important parameter. When the catalyst size is less than 40μ, less than 15% longer reactor length than that required for a reactor with a uniform catalyst distribution is needed to achieve a 90% hydrogen conversion. This may be acceptable. However, it is uncertain what is the effect of the catalyst size on catalyst pretreatment (the reduction of iron oxides to iron carbides).

- The Space-Time-Yield, i.e., gMol syngas converted/hr-mL expanded slurry, of the BSU F-T reactor was examined using a NM liquid phase mathematical model. As expected, the STY depends strongly on the catalyst loading if one assumes the same hydrodynamic parameters for all catalyst loadings. The STY reaches a maximum value with varying gas velocities. For catalyst loadings between 5-15 wt % Fe in the slurry, the feed gas superficial velocities at which the maximum STY occurs vary from 2.8 to 4.6 cm/s.

3. Future Work

- An operating manual for the BSU will be prepared.

- A material balance program will be developed.

- An F-T bubble-column reactor mathematical model to include a detailed description of the axial liquid phase dispersion will be constructed and used to examine the effect of this dispersion on the reactor performance.

- The improvement of the mathematical model will be continued to provide better tool to aid in the planning of the operation of the BSU F-T reactor.

V. NOMENCLATURE

A	Column cross-sectional area, (cm^2)
a_g	Gas bubble interfacial area, $6\ \epsilon_g/d_B$, (cm^2 gas-liquid area/mL expanded slurry)
C'	Catalyst loading, (gCat/mL slurry)
C'_{ca}	Average catalyst loading, (gCat/mL slurry)
C_{Fe}	Iron loading, (gFe/mL liquid)
$C^*_{H\ell}$	$C_{H\ell}$ in equilibrium with C_{Hg}, (mol/mL liquid)
C_H	H_2 concentration, (mol/mL liquid)
d_B	Bubble diameter, (cm)
d_c	Catalyst particle diameter, (cm)
d_R	Reactor diameter, (cm)
E	Axial dispersion coefficient, (cm^2/s)
f	H_2/CO at reactor inlet
f_{Fe}	Weight fraction of Fe in catalyst
g	Gravitational constant, 981, (cm/s^2)
K	$f(1+U)/(1+f)U$
K_H	H_2 solubility coefficient, $C_{Hg}/C^*_{H\ell}$, (mL liquid/mL gas)
k''_H	Intrinsic kinetic rate constant for H_2 conversion, $r_H/(1-\epsilon_g)(1-v)C_{H\ell}\ C_{Fe}$, (mL liquid/s-gFe)
k_ℓ	Liquid side mass transfer coefficient, (mL liquid/s-(cm^2 gas-liquid area))
L	Bubble column height, (cm)
ℓ	Reactor vertical distance from its entrance, (cm)
N_H	H_2 molar flow rate, (mol/s)
R_d	H_2 transport resistance from gas-liquid interface to bulk liquid phase, $K_H/k_\ell\ a_g$, (s-mL expanded slurry/mL gas)

R_K	Kinetic resistance, $K_H/k''_H\ C_{Fe}(1-\epsilon_g)(1-v)$, (s-mL expanded slurry/mL gas)
r_H	H_2 conversion rate, (mol/s-mL expanded slurry)
STY	Space-Time-Yield, (gMol syngas converted/hr-mL expanded slurry)
U	H_2/CO usage ratio
u	Superficial velocity, (cm/s)
u_{cs}	Catalyst settling velocity in a catalyst swamp, (cm/s)
u_{ct}	Single catalyst particle settling velocity, (cm/s)
v	Volumetric fraction of catalysts in slurry, $\rho_\ell\ w/(\rho_s+w(\rho_\ell-\rho_s))$, (mL catalyst/mL slurry)
v*	v as C'_C=0.1 gCat/mL slurry, (mL catalyst/mL slurry)
w	Weight fraction of catalysts in slurry, (gCat/g slurry)
X_H	H_2 conversion
	Greek Letters
α	Contraction factor, $-\gamma U/(1+U)$
α^*	αK
γ	Moles product per mole of H_2 converted in F-T reaction
ϵ_g	Gas hold-up, (mL gas/mL expanded slurry)
ρ	Density, (g/mL)
μ	Viscosity, (g/s-cm)
	Dimensionless Numbers
Fr	Froude number, $u_{gm}/(g\ d_R)^{0.5}$
Re_c	Reynolds number (catalyst particle), $u_{ct}\ d_c\ \rho_\ell/\mu_\ell$
Ar	Archimedes number, $\rho_\ell(\rho_c-\rho_\ell)g\ d_c/\mu_\ell^2$
Pe_c	Peclet number (catalyst particle), $u_{cs}\ L/E_c$
	Superscripts

e	At reactor exit
i	At reactor inlet
	Subscripts
m	Arithmetic mean value of that at reactor entrance and that at reactor exit
g	Gas
ℓ	Liquid
sℓ	Slurry
esℓ	Expanded slurry
c	Catalyst

VI. Literature

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Table 1

PARAMETERS ADOPTED IN THE CURRENT
F-T REACTOR MATHEMATICAL MODEL CALCULATIONS

d_B, (cm)	.07
T, (°C)	265
P, (MPa)	1.38
f	.7
U	.645
α	-.5
k''_H, (mL liquid/s-gFe)	1.1
ρ_c, (gCat/mL catalyst particle)	2.1
ρ_s, (gCat/mL catalyst solid)	5.2
f_{Fe}, (gFe/gCat)	.67

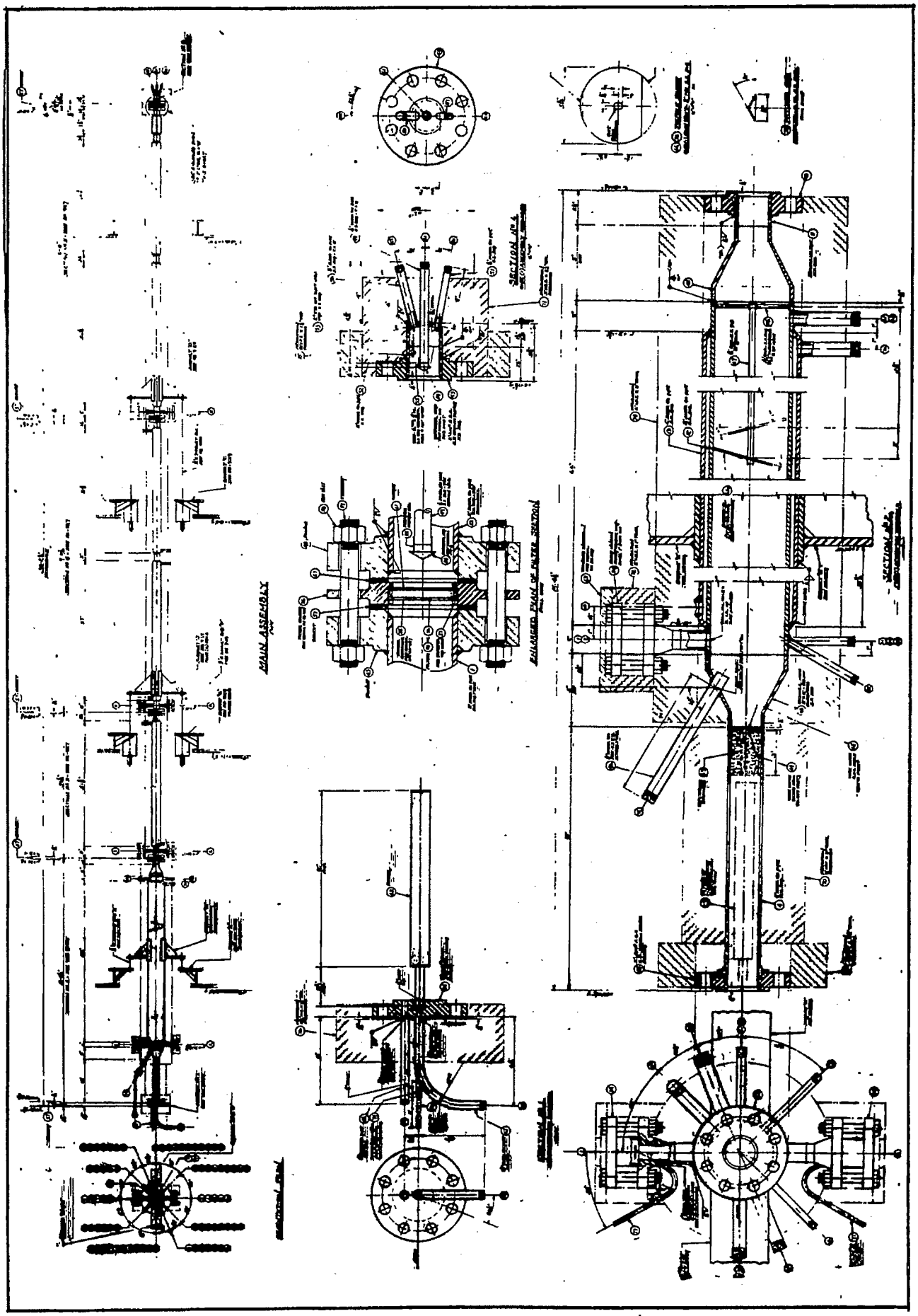


Figure 1: SLURRY F-T REACTOR ASSEMBLY AND DETAILS

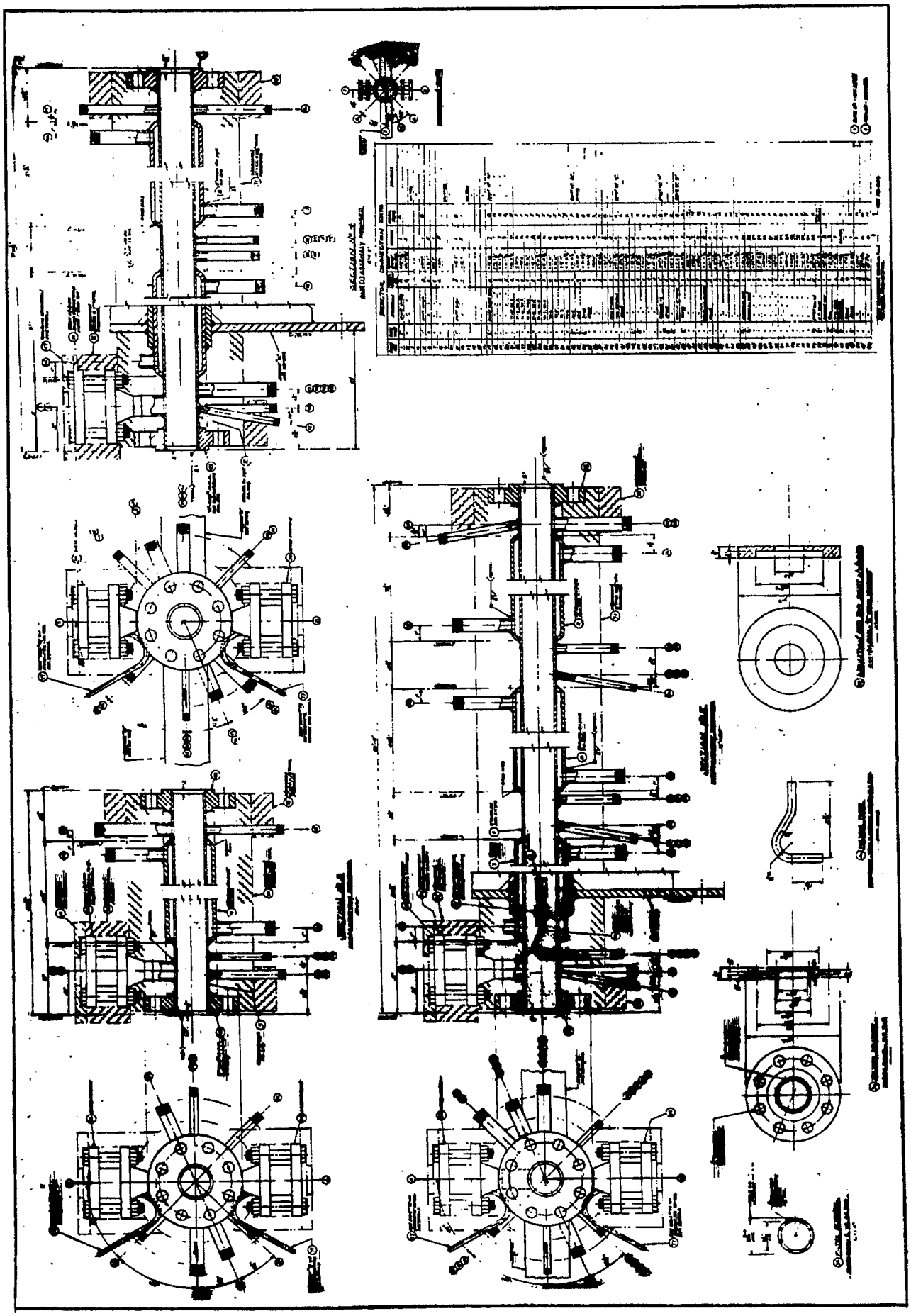


Figure 2: SLURRY F-T REACTOR DETAILS

Figure 3 : SECOND-STAGE FIXED-BED REACTOR ASSEMBLY AND DETAILS

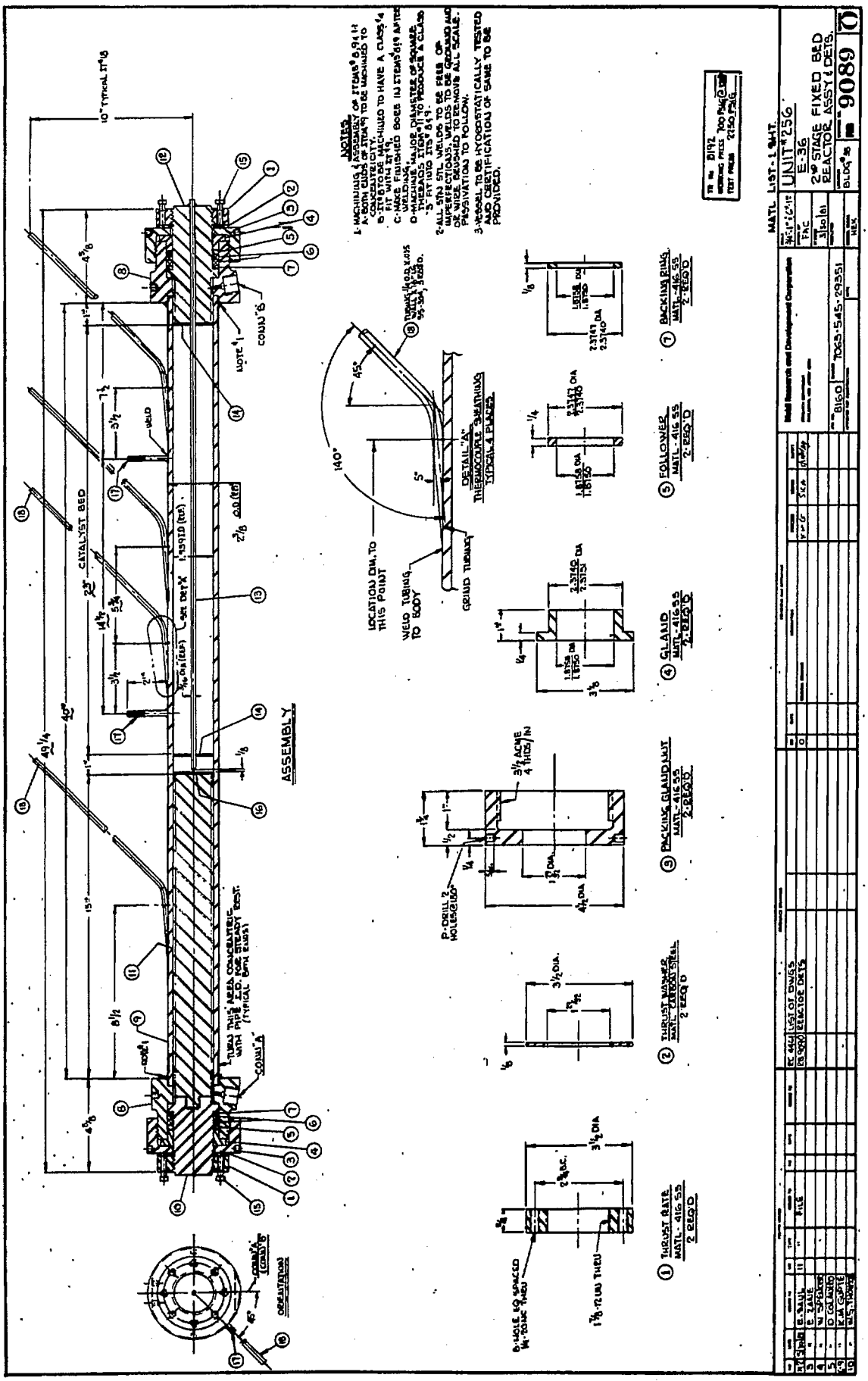
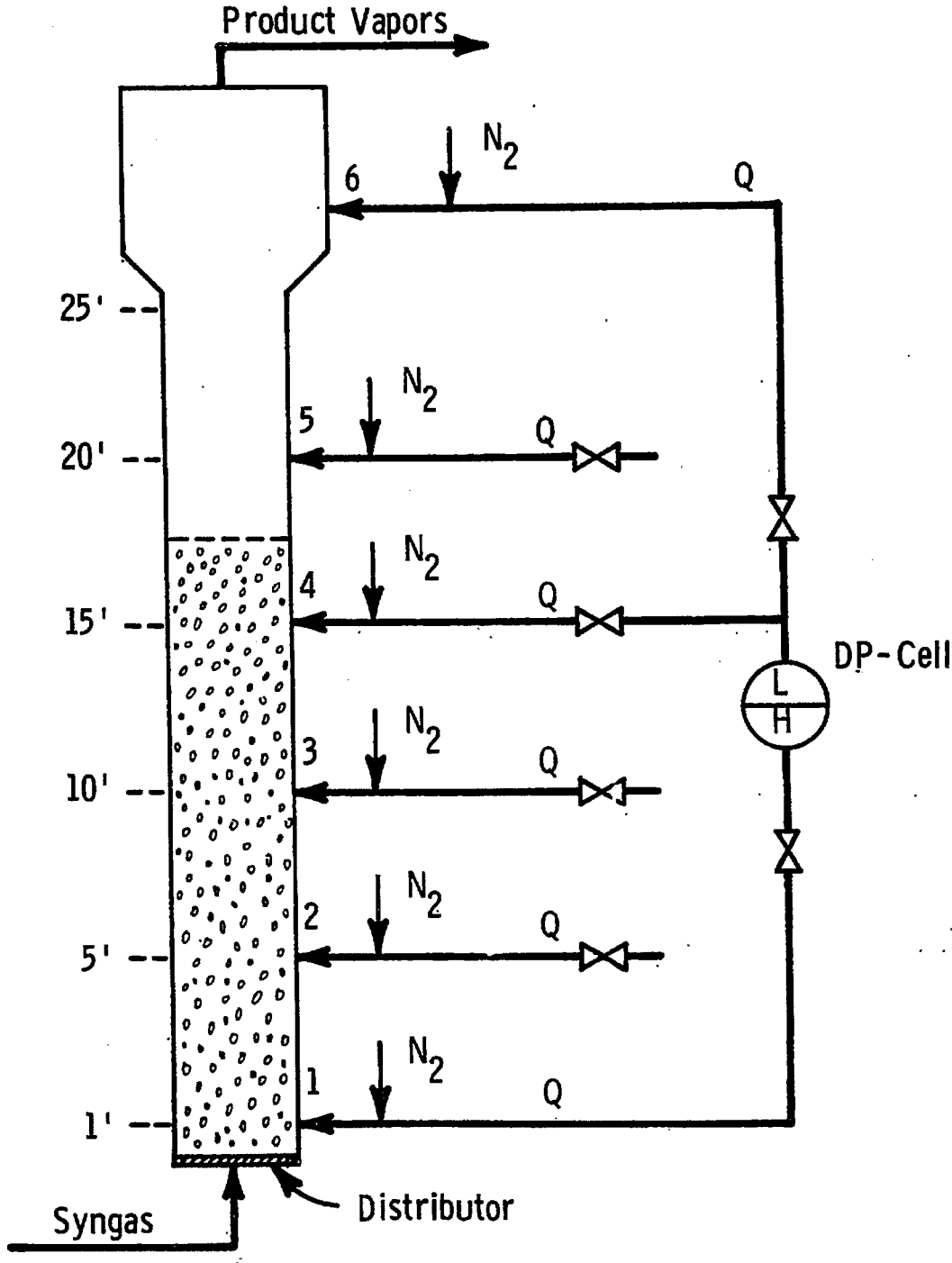


Figure 4

SCHEMATIC ARRANGEMENT OF DP-CELL FOR LIQUID LEVEL MEASUREMENT



Q: DP-Cell Lines

Figure 5

EFFECT OF AXIAL LIQUID MIXING ON BSU F-T REACTOR PERFORMANCE

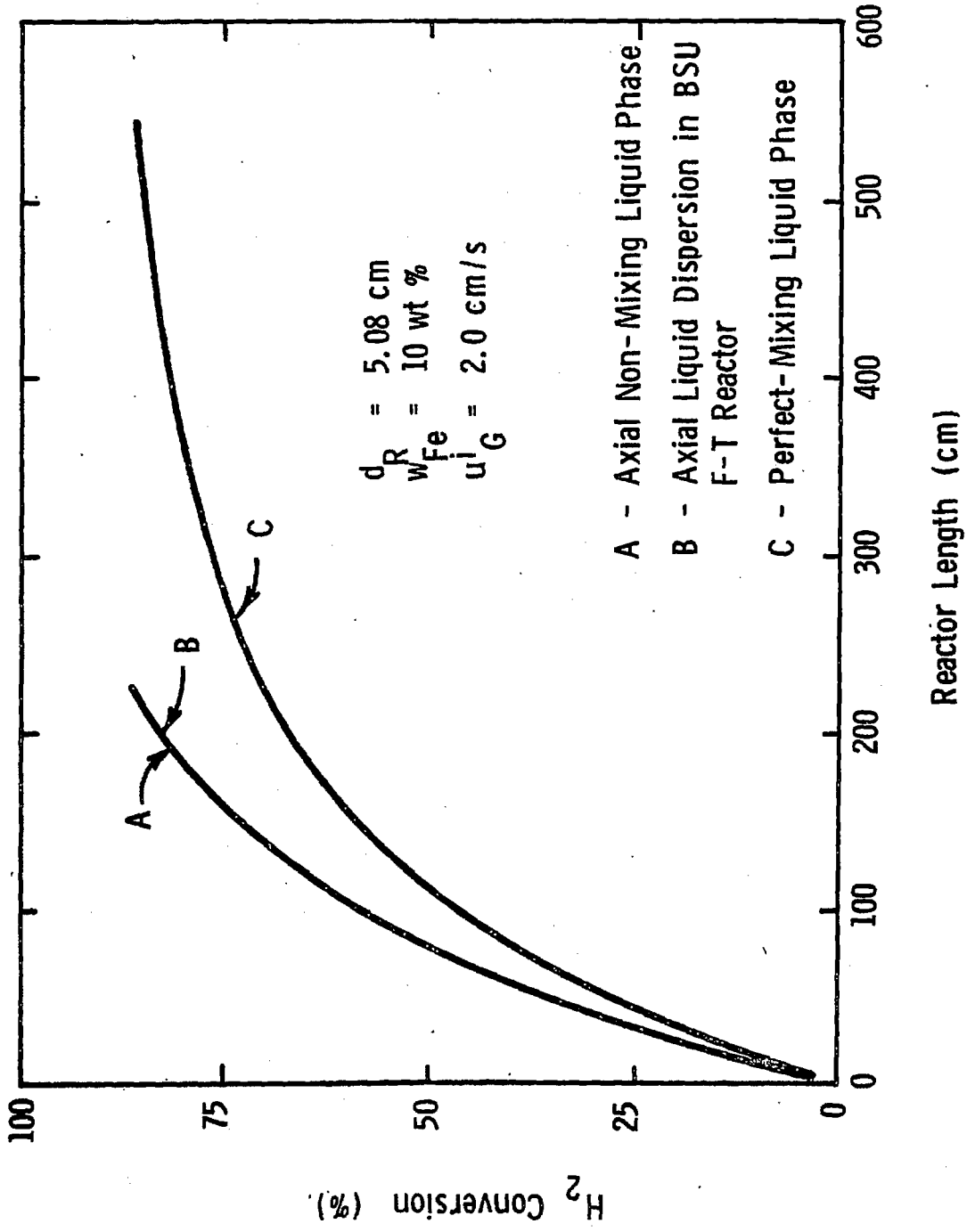


Figure 6

EFFECT OF AXIAL CATALYST DISTRIBUTION ON BSU F-T REACTOR PERFORMANCE

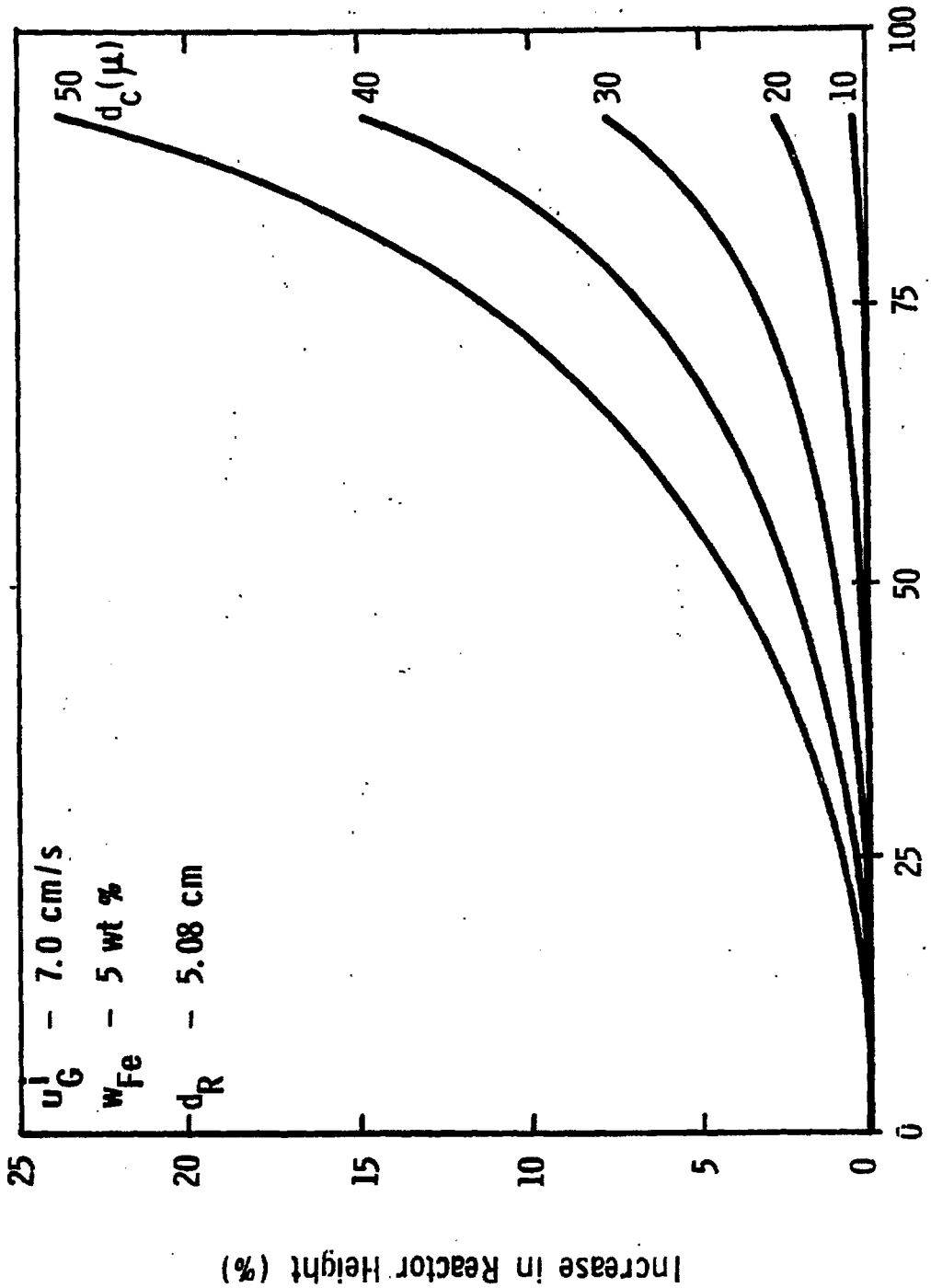
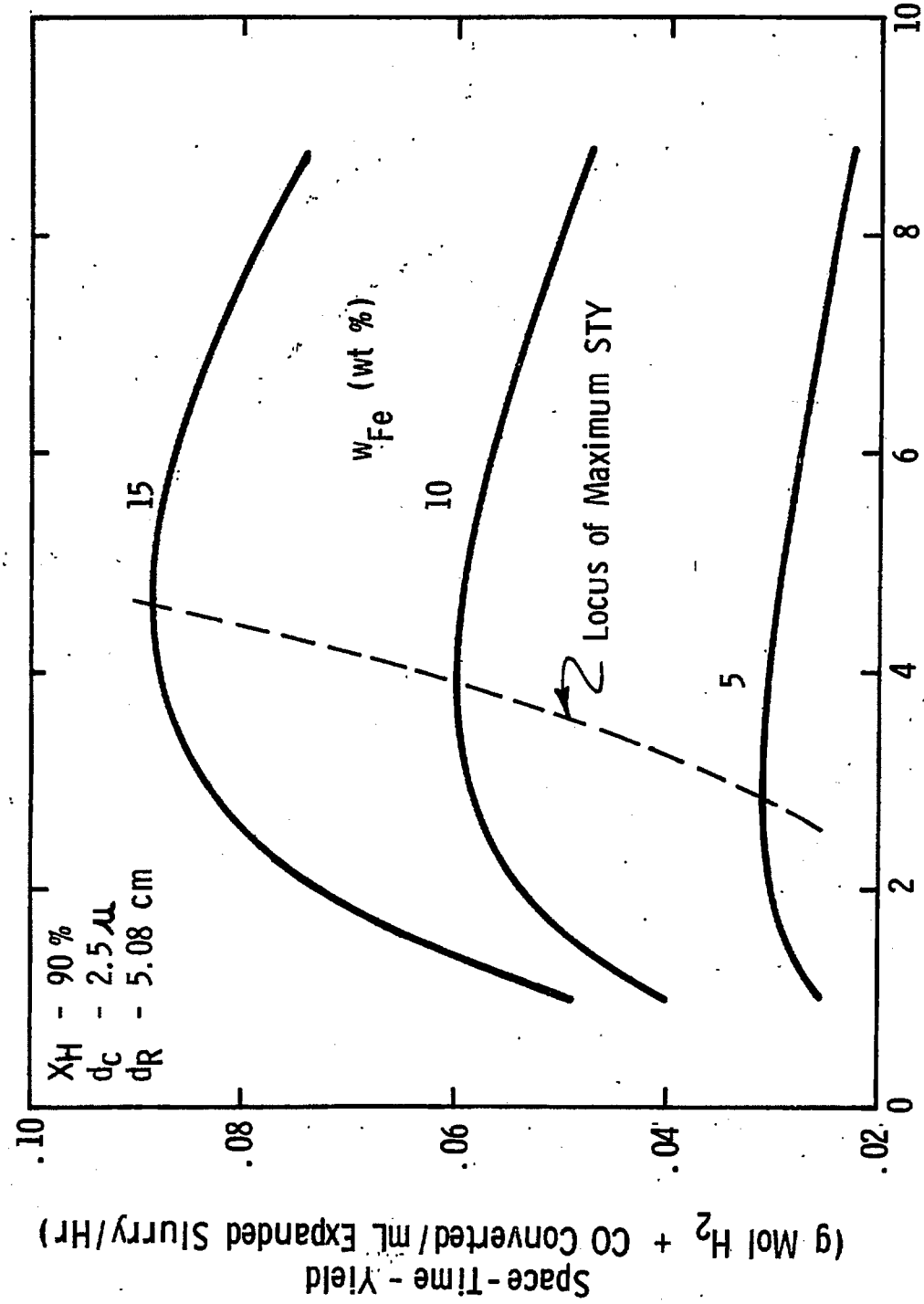


Figure 7

EFFECT OF CATALYST LOADING AND GAS VELOCITY ON BSU F-T REACTOR PERFORMANCE



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