THE SLURRY PHASE FISCHER-TROPSCH REACTOR

A comparison of slurry versus fixed-bed reactor designs for Fischer-Tropsch distillate production.

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Introduction

Bechtel recently concluded a study for the DOE which compared fixed-bed versus slurry reactors for several applications, including coal based methanol, mixed-alcohols and Fischer-Tropsch (F-T) synthesis. This paper reports on the evaluations performed on F-T synthesis for distillate production. Process design bases were established and block flow diagrams prepared for two cases of indirect coal liquefaction, in which

(1) slurry reactors were operated on synthesis gas directly from Shell gasifiers, after purification, at slightly under 0.5 H_2/CO ratio, and
(2) fixed-bed reactors were operated on shifted gas with a 2.0 H_2/CO ratio.

Reactor design principles were established, process flow diagrams were prepared, equipment was sized and costs compared for those sections of an overall coal-to-oil operation affected by reactor selection.

Capacity in each case was roughly 20,000 BPD of distillates, representing the output from three Shell gasifiers, each with a capacity of gasifying 2500 TPD of coal. The only distinction in product selectivity between the two cases was that the product at 0.5 ratio was more olefinic. Yields were based on Mobil's pilot plant data obtained for the DOE under Contract DE-AC22-83PC60019 (October 1985). The F-T reactor is operated at low temperature where a broad molecular weight range, waxy product is obtained. The wax fraction is then hydrocracked to maximize distillate production.

Figure 1 is a plot showing typical Schulz-Flory carbon number distributions for values of the Fischer-Tropsch chain growth probability factor, P, of 0.4, 0.65 and 0.9. The distribution at $P = 0.65$ is typical of an operation aimed at maximizing gasoline production. Gasoline and other products produced in this type of operation require extensive upgrading before they are marketable.

The distribution at P = 0.9 is typical of operation aimed at maximizing distillate production. There is a great deal of interest in this type of operation today because yields of light gases and oxygenates are reduced and because an easily upgraded wax fraction and high quality distillates are produced.

Types of Fischer-Tropsch Reactors

The challenge in Fischer-Tropsch reactor design is to remove the large heat of reaction, some 55,000 to 60,000 kJ/kgmol (24,000 to 26000 Btu/lbmol) of synthesis gas reacted. For Fischer-Tropsch operations directed at gasoline production, two types of reactor have been used:

- The entrained fluidized-bed Synthol reactor with riser coolers, used at Sasol.
- The fixed fluidized-bed with internal cooling coils used at the Carthage-Hydrocol plant at Brownsville, Texas.

The above reactor types are not satisfactory for waxy distillate production because the high molecular weight products and low temperatures would cause fluidization problems. While numerous reactor types have been proposed for distillate production, the selection narrows down to two main candidates:

- The low conversion per pass, fixed-bed, tubular reactor used at Sasol (the ARGE reactors).
- The higher conversion per pass, slurry bubble column reactor with internal cooling coils demonstrated by Rheinprüssen in the 1950's.

Mobil Research and Development chose the slurry reactor for the Fischer-Tropsch first step in its two step process for distillates production. It is also of interest to note that for natural-gas-based Fischer-Tropsch distillate designs, Shell has selected the tubular fixed-bed for their new plant in Malaysia, whereas Statoil has recently announced a slurry reactor design for the same type of application. Rentech and Ultrasystems have also announced that a slurry reactor design is being used for a 235 BPD landfill gas to distillate operation to be operational this year.

The Fixed-Bed Coal Based Fischer-Tropsch Design

A modern coal gasifier of the Texaco or Dow design produces a synthesis gas with a H_2/CO ratio of about 0.55, a Shell gasifier produces about 0.43

H_2/CO ratio. In order to use this gas in a fixed-bed Fischer-Tropsch reactor, the ratio must be adjusted by shift and the CO_2 removed prior to F-T synthesis. A H_2/CO ratio of 2 is stoichiometric for the reaction:

$$2H_2 + CO \rightarrow -CH_2- + H_2O \qquad (1)$$

In this case, the water gas shift reaction is not desired since it produces unwanted CO_2 and cobalt type catalysts have been developed which do not have this activity.

Figure 2 shows the Block Flow Diagram for the fixed-bed case. Key plants for which process flow diagrams and equipment lists were developed are shown as shaded blocks in the diagram.

The Shell gasifier package includes a waste heat boiler and a scrubber for carbon removal. A portion of the gasifier outlet is shifted with steam addition to produce CO_2 and H_2. The water condensed in the effluent cooler is sent to sour water stripping and the gas is compressed to synthesis pressure and sent to a selective Rectisol unit for CO_2 and H_2S removal. This process was chosen over Selexol since the latter would have required a double COS hydrolysis and CO_2 removal sequence to achieve adequate sulfur removal. A large amount of CO_2 (17,244 mph) is removed in the Rectisol unit by nitrogen stripping and vented to the atmosphere. A zinc guard bed is employed for sulfur polishing prior to F-T synthesis.

The fixed-bed converters operate at 37% CO conversion per pass and 95% ultimate conversion with a 2.34 recycle to fresh feed feed ratio. This high level of conversion is only possible because of the very low inerts level (0.4%) in the synthesis gas. Off gas from the F-T effluent separator (122,629 mph) is treated with inhibited 30% MEA for CO_2 removal and returned to the cryogenic F-T gas plant for recycle to the F-T converter. Enough hydrogen is recovered to treat the liquid product and a small purge is taken for inerts removal.

Product upgrading follows the sequence defined by MITRE (Gray, et. al., Sandia Report WP89W00144-1) and includes wax hydrocracking, distillate hydrotreating, catalytic polymerization of C3/C4's, heavy poly gasoline hydrotreating, isomerization of the C5/C6's, catalytic reforming of the naphtha from wax hydrocracking and middle distillate hydrotreating, and alkylation of cat poly olefins with isobutane from the cat reformer. MITRE also shows "alcohols recovery" from the small amount of product water. Actually, there are other oxygenates present than just alcohols.

The question of oxygenates recovery from the product water is not addressed in this study. It could be more of a problem in the fixed-bed than in the slurry reactor case because of the larger quantity of water to be handled. On the other hand, if a cobalt based catalyst is used, oxygenates production could be so low that only a biotreatment step is required on the product water before its reuse as a utility.

Low H_2/CO Ratio Fischer-Tropsch in the Slurry Reactor

A 0.6 H_2/CO ratio is approximately stoichiometric for a F-T reactor, without steam addition, where full water gas shift (WGS) equilibrium is achieved. The reactions involved are reaction number (1) plus:

$$H_2O + CO \leftrightarrow H_2 + CO_2 \qquad (2)$$

giving the overall reaction:

$$H_2 + 2CO \rightarrow -CH_2- + CO_2 \qquad (3)$$

Because equilibrium in reaction 2 heavily favors CO_2 production at F-T conditions, reaction 3 predominates over reaction 1. Catalysts such as precipitated iron promote the WGS reaction to the extent that usage ratio, U, equals the H_2/CO inlet ratio, I, at a value of I of about 0.7.

The BFD for the slurry reactor case, given in Figure 3, differs from Figure 2 only in the location of some steam additions, the replacement of the water gas shift step with a COS/HCN hydrolysis step and the use of Selexol instead of Rectisol for acid gas removal. Conversion per pass is 80% in the F-T reactor. Since the gas is below stoichiometric H_2/CO ratio, steam is added to the recycle gas to supplement the water produced by reaction 1, shifting additional CO to produce the required amount of hydrogen.

The recycle loop and product recovery are similar to that provided for the fixed-bed reactor case except that:

- Much more CO_2 (19,475 mph) is removed from the recycle gas,
- More hydrogen recovery is required to supply the treating units, and
- Considerably less product water must be handled.

While feed gas preparation has been simplified, it has been necessary to increase the number of downstream CO_2 removal units from two to three to accommodate increasing CO_2 removal from 2,281 to 19,475 mph. The total amount of CO_2 to be removed is comparable in the two cases, the difference lies in where it is removed.

The Slurry Reactor Process Flow Diagram (Figure 4)

A slurry reactor has been chosen for low H_2/CO ratio synthesis for the following reasons:

1. Low H_2/CO ratio can lead to carbon formation via the Boudouard reaction:

$$2CO \leftrightarrow CO_2 + C \downarrow \qquad (4)$$

A slurry reactor, however, raises the H_2/CO ratio that the catalyst actually sees owing to the combination of a higher mass transfer coefficient for hydrogen and a higher CO consumption by reaction 2..

2. Even if some carbon formation does occur, the slurry reactor offers the possibility for its removal.

3. Catalyst deactivation can be handled by means of periodic catalyst withdrawal and addition, whereas replacement of fixed-bed catalyst requires a shutdown.

4. Distillate production requires low reaction temperatures for good yield. Because of uniform temperatures and catalyst replacement, a slurry reactor can be run continuously at end of run temperature for the fixed-bed reactor and give equivalent yield distribution. This gives both high reactor productivity and the potential for higher pressure steam generation.

Weighed against these advantages are the need to provide for product removal and separation from the catalyst as well as facilities for continuous addition of preactivated catalyst.

Figure 4 shows the details of the slurry reactor process flow diagram. An internally cooled reactor is provided, with steam generation in the cooling tubes. A portion of the slurry is continuously removed through a hydroclone for liquid product recovery and returned to the bottom of the

reactor. A makeup catalyst pretreatment section is provided which is a separate flow diagram in itself. Spent catalyst is removed from the lower part of the reactor and hot-centrifuged for wax recovery. The combined wax fraction is hot-filtered to remove traces of fines and sent to storage prior to hydrocracking.

The reactor effluent vapors are cooled and separated with the gases going to CO_2 removal and the gas plant, liquid hydrocarbons going to the gas plant and product fractionation and the water phase, to alcohols recovery and/or biological treatment.

The Tubular Fixed-Bed Reactor Process Flow Diagram (Figure 5)

Figure 5 shows the relative simplicity of the fixed-bed reactor system as compared to the slurry reactor system. No catalyst makeup and removal facilities are provided but, as a result, periodic shutdowns for catalyst replacement are necessary. This is only feasible if an extended catalyst life can be assured. It is understood that the ARGE reactors are shut down frequently for catalyst replacement but it is believed that Shell expects reasonable catalyst life in their new Middle Distillate Process. As the catalyst deactivates, temperature can be raised about 25 °C to compensate before selectivity starts to degrade. If fresh catalyst activity is used for design, then start of run temperatures must also be used.

The catalyst in the tubular fixed-bed reactor is contained within tubes which are supported at each end by tubesheets. Steam is generated in the shell, outside of the tubes. A temperature rise of 25 °C is taken from inlet to outlet of the tubes.

A holding zone is provided at the bottom of the reactor for heavy liquids which are removed and sent to wax recovery. Handling of the various reactor effluents is similar to the slurry reactor case, except that catalyst fines removal is not necessary.

Reactor Sizing Considerations

At a given temperature, conversion in a Fischer-Tropsch reactor is correlated with space velocity per unit weight of catalyst. Satterfield, et. al. (1983) found that there was little difference between a slurry reactor and a fixed-bed reactor when expressed in these terms. Catalyst loading in kg/m3 of reaction volume is an important criteria in reactor design and here the fixed-bed reactor has a distinct advantage. This is compensated

for in Fischer-Tropsch synthesis by the higher average slurry reactor temperature described above.

In comparing reactor designs, the practical limit on shell diameter has been taken as 4.8 meters in both cases. This is the current commercial limit on Lurgi methanol reactors, which are tubular fixed-bed designs similar to the F-T reactors considered in this study.

Heat transfer rate, space velocity and pressure drop must be balanced in the tubular fixed-bed reactor, which is scaled-up by adding tubes of the same diameter as the pilot plant tubes. The slurry reactor is scaled-up by increasing the diameter of the vessel. Attention must then be paid to backmixing effects which will be much more pronounced as diameter increases.

Slurry reactor models have been proposed which incorporate axial dispersion coefficients for both the gas, liquid and solid phases. These models lead to boundary limit problems solvable by collocation methods. Since axial dispersion coefficients are generally not available for reactor configurations with many cooling tubes and since simplifying assumptions are still required in defining reaction kinetics and mass transfer, Bechtel elected to approach the design by means of limiting models. While approximate, these are much simpler to work with and give a better feel for what variables are important in design.

These limiting model are summarized in an addendum to this paper. They are as follows:

- Model 1 assumes plug flow of both gas and liquid
- Model 2 assumes plug flow of the gas with the liquid being fully backmixed
- Model 3 assumes both gas and liquid backmixed (i.e. a CSTR)

All three models assume that reaction rate is proportional to liquid phase hydrogen concentration, that there is a constant H_2/CO usage ratio, U and a constant contraction factor relative to conversion, α. They also assume isothermal behavior and that the solids are uniformly dispersed in the liquid.

Models 1 and 3 lead to the summation of mass transfer and reaction rate resistances in terms of an overall rate coefficient K, where

$$1/K = 1/K_M + 1/K_R \qquad (5)$$

Examination of Model 2 reveals that it reduces to Model 3 when reaction rate controls and to Model 1 when mass transfer controls. Intuitively this is correct since gas phase mixing should be important only when gas-liquid mass transfer is important.

The design assumption for this study is that the pilot plant slurry reactor will closely approach plug flow (Model 1) whereas the commercial reactor, because of its much lower L/D, follows Model 2. It is believed that this is a valid, though somewhat conservative, assumption.

Figure 6 shows the relationship between hydrogen conversion and the dimensionless Stanton Number (K/SV) for the three models, where K is the overall rate constant and SV is the space velocity, both in sec^{-1}. The slurry reactor design case, using Model 2, is shown as a circle in Figure 6 and falls close to the Model 3 line since mass transfer only contributes 10% of the total resistance. The design hydrogen conversion is 72.6% for an overall synthesis gas conversion of 80%. By limiting conversion per pass to 80%, backmixing effects are minimized without greatly increasing total flow.

Commercial Reactor Designs

The design principles used for slurry and fixed-bed reactor design are summarized in Tables 1 and 2. For further details the reader is referred to the final DOE report on this project. For present purposes it is only important to say that the rate constant for the slurry reactor design was obtained by backfitting Rheinprüssen laboratory data on precipitated iron catalyst (Kölbel, 1980) and the fixed-bed space velocity was obtained from published Shell Oil data for 2.5 mm diameter particles correlated according to Model 1 (Post, 1989). Other fixed-bed data showed a non-linear improvement in space velocity requirement with pressure which could be approximated as proportional to $P^{0.5}$ (Singleton, 1983). The same type of effect was assumed for the slurry case as well.

The resulting designs are compared in Table 3. Bed heights are comparable (11.7 m for the slurry reactor versus 12.7 m for the fixed-bed) but effective reaction volume is less for the fixed-bed because of the lower effective cross sectional area. It will be noted that, owing to the difference in design temperature, the design space velocities in Nm^3 of total feed gas/(h·kgCat) are quite comparable, even though the conversion drops from 80% to 37% between the slurry and fixed-bed cases. The catalyst loading is 250 kg/m^3 for the slurry reactor and 850 kg/m^3 for

the fixed-bed and this is reflected in the much higher allowable space velocity for the fixed-bed in $Nm^3/(h \cdot m^3)$.

The syngas space time yield (STY) equals the syngas conversion times the total feed space velocity times the syngas concentration in the total feed gas. The -CH2- space time yield is 1/3 of that for syngas and is 90 kg/$(h \cdot m^3)$ for the slurry reactor versus 111, for the fixed-bed reactor. Multiplying STY by the effective reaction volume, the overall hydrocarbon production from 8 fixed-bed reactors is equivalent to that from 6 slurry reactors at about 2365 metric tons per day.

Reactor Configurations

The final reactor configurations arrived at are shown in Figures 7 and 8. Because of the requirement for 2481 cooling tubes in the slurry reactor case, a double tube sheet arrangement was devised with bayonet tubes circulating coolant from the top head to the space between the tubesheets. The slurry level is kept between the 14" gas outlet and the 4" slurry outlet leading to the hydroclone for liquid product separation. Feed gas is brought in through a ring type distributor and the space below the distributor has been assumed to be 15% effective for reaction.

The design of such reactor to accommodate a great many cooling tubes is a challenge and it may be desirable to revisit the idea of circulating the slurry through an external exchanger, especially if reactor productivity is further increased. Such a design had been studied at the LaPorte LPMeOH™ pilot plant but abandoned in favor of internal cooling coils (Studer, 1989).

The design of the fixed-bed reactor (Figure 8) is modelled after the ARGE reactors at Sasol and is very much like that of a tubular methanol reactor, except for the vapor-liquid separation zone at the bottom. The ARGE reactors also have a grid below the bottom tubesheet, with a grid-release device for catalyst removal. This is not shown in the drawing but would certainly be included in the design.

The weight of the slurry reactor is 639,000 pounds. The fixed-bed reactor, because of its greater number of tubes, weighs appreciably more at 961,000 pounds. This has a direct effect on the costs quoted for the two vessels.

Capital Cost Comparisons

The slurry reactor design shown in Figure 7 and Table 3 was based on a superficial gas velocity close to 15 cm/s and a slurry concentration of 35 wt%. These values are well above what has been demonstrated in F-T pilot plant operations to date but have been adequately demonstrated in the LaPorte pilot plant for the slurry methanol process (LPMeOH™). In fact, Air Products has stated that their current design practise is well beyond this.

Increasing superficial velocity gives higher reactor feed gas capacity and should improve performance provided that gas holdup remains reasonable. Recent hydrodynamic measurements (Bukur, 1987) lend additional confidence that superficial velocities of 15 cm/s or above are feasible. Increased slurry concentration will increase catalyst loading, which is important from a reaction rate standpoint, but will increase the slurry viscosity leading to poorer mass transfer. At some point, mass transfer will become controlling but the LaPorte data indicate that this level is somewhat above 35 wt% for methanol synthesis.

The design methods outlined in Table 1 take in to account the effects of both superficial velocity and slurry concentration and Bechtel has considerable confidence that the designs are feasible. Nevertheless, a more conservative design, using existing F-T pilot plant limitations as a basis, has been developed as an alternative. If superficial velocity is cut in half and slurry concentration is limited to 20 wt%, reactor sizing stays roughly the same but the number of reactors increases from 6 to 11.

Table 4 shows the effect of base case and conservative slurry reactor design conditions on the cost comparison with fixed-bed reactors. The base case cost of the 6 slurry reactors is half of that of the 8 fixed-bed reactors. Because of the auxiliary equipment required with the slurry reactor, the cost of the synthesis section in the slurry case is a somewhat higher percentage, $49 million versus $82 million. The cost advantage for the slurry reactor is almost lost, however, if the more conservative design basis is employed.

Table 5 shows the cost comparison for key process plants between the two cases. These plants are the shaded process blocks in Figures 2 and 3 and have been separated for comparison into gas handling units (gas preparation and downstream CO_2 removal) and Fischer-Tropsch synthesis units. It is of interest to note that essentially all of the savings are in the

synthesis units (F-T reaction section and the F-T gas plant). The F-T gas plant is considerably smaller in the slurry case because of the lower amount of recycle gas.

Surprisingly, the total cost of the gas handling units does not differ appreciably between cases, despite the difference in H_2/CO ratio to F-T synthesis. One way of rationalizing this result is that the same amount of CO_2 must be removed regardless of whether this is done before or after the F-T reaction. The cost of CO_2 removal in the slurry case centers around the regenerator and solution circulation and could possibly be improved with a higher capacity solvent. Nevertheless, these results suggest that further consideration should be given to a slurry reactor design at 2.0 H_2/CO ratio.

The cost differential shown in Table 5 is $54.6 million in favor of the slurry reactor case. Other cost savings have been identified in the upgrading plants and in the offsites. These are summarized in Table 6 and the total project cost differential comes out to be $91.4 million (Gulf Coast basis, first quarter 1990). From other studies, such as those of MITRE, it is estimated that these savings are about 8.5% of the total cost of a coal to liquids project.

Reaction Pressure Considerations

In both cases, a compression step was included to raise the pressure in the F-T reactors to 28.3 atmospheres. As noted previously the assumption was made in both cases that allowable SV increases as $P^{0.5}$. With this assumption, the following effects are noted:

- On halving the pressure, the number of slurry reactors increases from 6 to 11. At the same time, the shell thickness is halved and the bed length decreases as $P^{0.5}$. The net effect on reactor section cost is probably a wash.
- The capacity of a tubular fixed-bed reactor of given dimensions increases as $P^{0.5}$. There is no effect on shell thickness since this is set by steam side pressure, though the heads become thicker. On halving the pressure, the number of reactors increases from 8 to 11 while the cost per reactor decreases only slightly.

The conclusion is that the higher pressure level may be beneficial in either case but favors the fixed-bed more than the slurry reactor. Once the effect of pressure on reaction kinetics is established, further studies are recommended to optimize each design.

Operating Cost Differential

Very few differences were identified in operating costs between the two cases and these are summarized in Table 7. A somewhat higher gas consumption was identified for the fixed-bed case but this was roughly balanced by the expected lower catalyst life for the slurry reactor case because of the lower H_2/CO ratio feed gas. These values of catalyst life are not well substantiated, however, and the slurry reactor case would improve significantly if catalyst usage were lower.

Conclusions

- The slurry reactor can achieve 80% conversion per pass at the same space velocity (per unit weight of catalyst) as a fixed-bed reactor running at 37% conversion per pass - throughput is only 40% of that of the fixed-bed reactor
- This because the slurry reactor can run continuously at end of run temperature for the fixed-bed reactor
- This overcomes the lower catalyst loading in the slurry reactor (250 vs 850 kg/m^3 Cat) and 6 slurry reactors produce the same as 8 fixed-bed reactors
- However, this advantage is lost if conventional values of slurry concentration and superficial gas velocity are employed - 11 reactors are then required
- Pilot plant demonstration of high superficial velocity, high slurry concentration operation is strongly recommended
- Fewer cooling tubes are required and each slurry reactor weighs and costs 2/3rds that of the comparable fixed-bed reactor
- At the design conditions used the slurry reactor F-T sections (including F-T gas plants) are half the cost of the fixed-bed F-T sections
- Somewhat surprisingly, the remainder of the plant remains constant in cost regardless of the H_2/CO ratio of the feed gas to F-T Synthesis.
- The case of a slurry reactor running on adjusted feed gas ($H_2/CO = 2$) should be studied
- The effect of pressure also needs further definition - a compression step has been included and this helps the fixed-bed more than the slurry reactor case
- As superficial velocity, slurry concentration and pressure are increased the number of cooling tube increases and an external cooling loop should be considered

Addendum

Limiting Fischer-Tropsch Models

Model 1
Plug Flow Both Phases

$$(1 + \alpha^*)\cdot\ln(1 - X_H) + \alpha^*\cdot X_H = - K/SV$$

Model 2
Plug Flow Gas - Fully Mixed Liquid

$$((\alpha^*\cdot X_H + (1 + \alpha^*\cdot Y)\cdot\ln(1 - x_H/Y))/(1 + \alpha^*\cdot n) = K_M/SV$$

where

$$Y = (1 - n)/(1 + \alpha^*\cdot n)$$

$$n = X_H/(K_R/SV)$$

Model 3
Both Phases Completely Mixed

$$X_H\cdot(1 + \alpha^*\cdot X_H)/(1 - X_H) = K/SV$$

In all cases:

$$K_R = k_H\cdot\varepsilon_L/H_H, \qquad K_M = k_La/H_H$$

$$K = \frac{K_R.K_M}{K_R + K_M}$$

$$SV = \frac{GHSV}{3600} \times \frac{T}{273} \times \frac{101.3}{P}$$

Models 1 and 2 follow the derivations of Deckwer ((1981) and Bukur (1983) and all models assume the reaction is first order in hydrogen concentration.

Notation

a	gas-liquid interfacial area, m^{-1}
C'Cat	catalyst concentration, kg/m^3
C_{HG}	hydrogen concentration in gas phase, kg mole/m^3
C^*_{HL}	hydrogen concentration,liquid, in equilibrium with gas, kg mole/m^3
C_{HL}	hydrogen concentration in the liquid phase, kg mole/m^3
D	I.D. of reactor, m
GHSV	Gas hourly space velocity, Nm^3 $(H_2+CO)/[h \cdot m^3$ reactor volume], (reactor volume is expanded slurry height times cross section area)
H	solubility coefficient of hydrogen = C_{HG}/C^*_{HL}
I	Inlet ratio of CO/H_2
k_L	liquid side mass transfer coefficient, m/s
k_H	effective reaction rate constant for hydrogen consumption, s^{-1} (note that to agree with space velocity in $Nm^3/[s \cdot kgCat]$, k_H = $k_H' \cdot C'Cat$ where k_H' is in $m^3/[kg \cdot s]$
L	Length of expanded slurry bed, m
P	pressure, kPa
r	rate of hydrogen consumption, $r = k_H \cdot C_{HL}$, kg moles/$[m^3 \cdot s]$
SV	Space velocity in actual m^3 inlet gas/$[s \cdot m^3]$
T	temperature, °K
U	Usage ratio of CO/H_2
X_H	hydrogen fractional conversion per pass (If U = I, $X_H = X_{CO}$)
α	contraction factor, $\alpha=[m^3/s(X_{H_2+CO=1})-m^3/s(inlet)]/[m^3/s(inlet)]$
α^*	contraction factor modified for H_2 conversion, $\alpha^* = \alpha \cdot (1+U)/(1+I)$
ε_L	fractional liquid hold-up

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Table 1

Commercial Slurry Reactor Design

- Bukur expression used for fractional gas holdup, εG
- Akita-Yoshida correlation used for k_La (k_La proportional to diffusivity and to $\varepsilon G^{1.1}$), with Akgerman correction for slurry concentration (new)
- Akgerman data used for hydrogen diffusivity
- Isothermal behavior assumed with coolant to reactor $\Delta T = 50$ °F and $h = 120$ for overall heat flux of 6000 Btu/(hr · ft^2)
- Properties evaluated at average gas velocity with no allowance for pressure drop
- Rate constant fit to Rheinprüssen lab data using precipitated iron catalyst
- Commercial design assumes Model 2 behavior - syngas conversion limited to 80%

Table 2

Commercial Fixed-Bed Tubular Reactor Design

- Model 1 applies (gas and liquid plug flow)
- Shell Oil correlations which use this model and allow for internal pore diffusion are used to define space velocity requirement (cobalt based catalyst)
- Properties are evaluated at average T, P, flow rate
- Tube diameter, length and number adjusted to give pressure drop and average film ΔT comparable to commercial tubular reactor designs
- Conversion per pass limited to 25 - 40% to balance heat transfer and pressure drop

Table 3

REACTOR DESIGN COMPARISON
4.8 Meter Diameter

	Slurry	Fixed-Bed
Number of Reactors	6	8
Height of Bed, m	11.69	12.65
No. of Tubes	2481	9602
Effective XSect Area, %	84	48
Eff. Reactor Volume, m^3	183	110
Temperature, °C	257	225 (outlet - start of run)
R/FF Ratio	0.264	2.34
Syngas Conversion, %	80.0	36.9
SV, $Nm^3/(h \cdot kgCat)$	2.38	2.26
Catalyst Loading, kg/m^3	250	850
SV, $Nm^3/(h \cdot m^3)$	595	1917
H2+CO in Total Feed, %	90.8	75.5
STY, kg $-CH_2-/(h \cdot m^3)$	90	111
$-CH_2-$ Production, MTD	2365	2365

Table 4

Capital Cost Comparison
Fischer-Tropsch Synthesis Plant
Effect of Slurry Reactor Design Conditions

	Slurry Reactor		Fixed-Bed Reactor
	Base Case Design	Conservative Design	Base Case Design
Gas Velocity, cm/s	14	7	43
Slurry Conc., Wt%	35	20	NA
No. of Reactors	6	11	8
Cost in $Millions - Gulf Coast Basis - 1st Q 1990			
Reactors Only	17.1	31.4	34.4
Total Major Equipment	26.8	42.1	44.8
F-T Synthesis Section	49.0	77.0	82.1

Table 5

Capital Cost Comparison
Fischer-Tropsch Cases - Selected Units
20,000 BPSD of Distillates
$Millions - Gulf Coast Basis - 1st Q 1990

Key Process Plants:	Slurry Reactor		Fixed-Bed Reactor
Gas Handling			
Shift Conversion	NA		22.4
COS Hydrolysis	15.3		NA
Acid Gas Removal	43.3		65.7
SynGas Compression	7.2		11.6
CO2 Removal	47.6		14.3
Total Gas Handling	**113.4**		**114.0**
Fischer-Tropsch			
F-T Synthesis	49.0		82.1
F-T Gas Plant	11.3		32.4
Total Fischer-Tropsch	**60.3**		**114.5**
Total Key Plants	**173.8**		**228.4**
Cost Differential		**54.6**	

Table 6

Capital Cost Differential
Fixed-Bed Over Slurry Reactor
20,000 BPSD of Distillates
$Millions - Gulf Coast Basis - 1st Q 1990

	Fixed-Bed Reactor less Slurry Reactor
Key Process Plants	**54.6**
Other Process Plants	3.0
Offsites	6.5
Total Direct Cost	**64.1**
Contractor's Indirects	9.0
Eng'ng + Cont.	18.3
Project Cost Differential	**91.4**

Table 7

Fischer-Tropsch Operating Costs
Selected Cost Items
90% On-Stream Factor

	Slurry Reactor	Fixed-Bed Reactor
Variable Costs	$MM/yr	$MM/yr
Fuel Gas @ $2.50/MMBtu	12.44	17.58
Raw Water @ $0.08/MGal	0.25	0.30
Catalysts	10.99 (60 day F-T life)	4.61 (1 year F-T life)
Chemicals	3.05	3.59
Total Variable Costs	**26.73**	**26.08**
Fixed Costs @3% of Invest./yr	**10.66**	**13.27**
Total Operating Costs	**37.39**	**39.35**

Figure 1

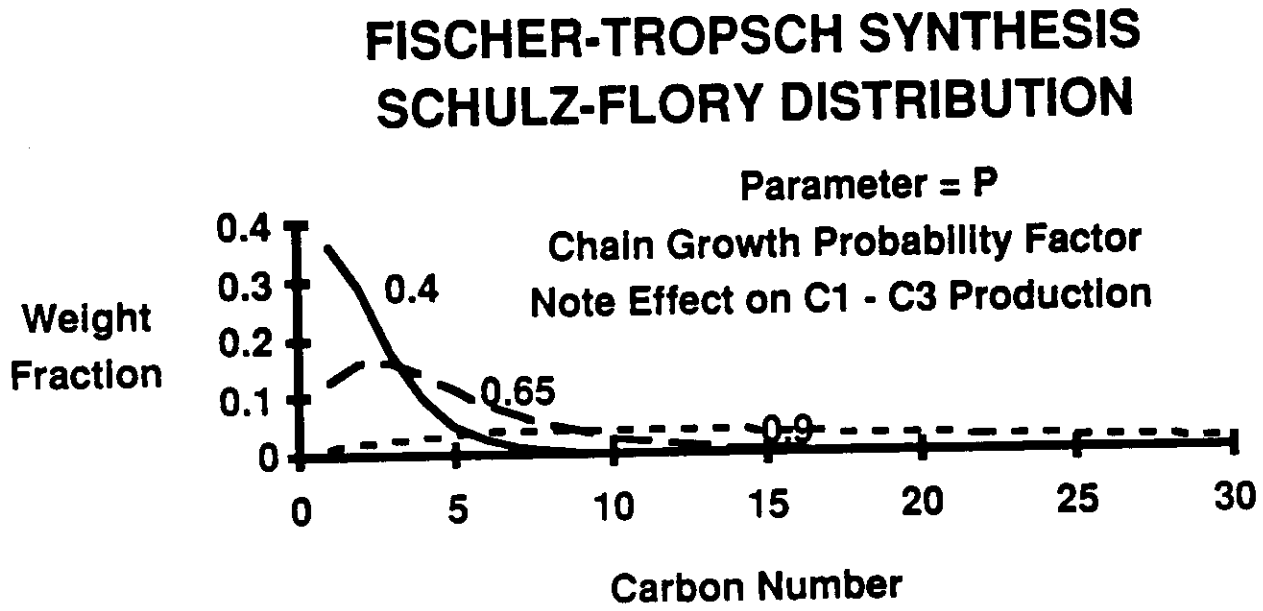


Figure 2

FISCHER TROPSCH SYNTHESIS
FIXED-BED REACTOR CASE
BLOCK FLOW DIAGRAM

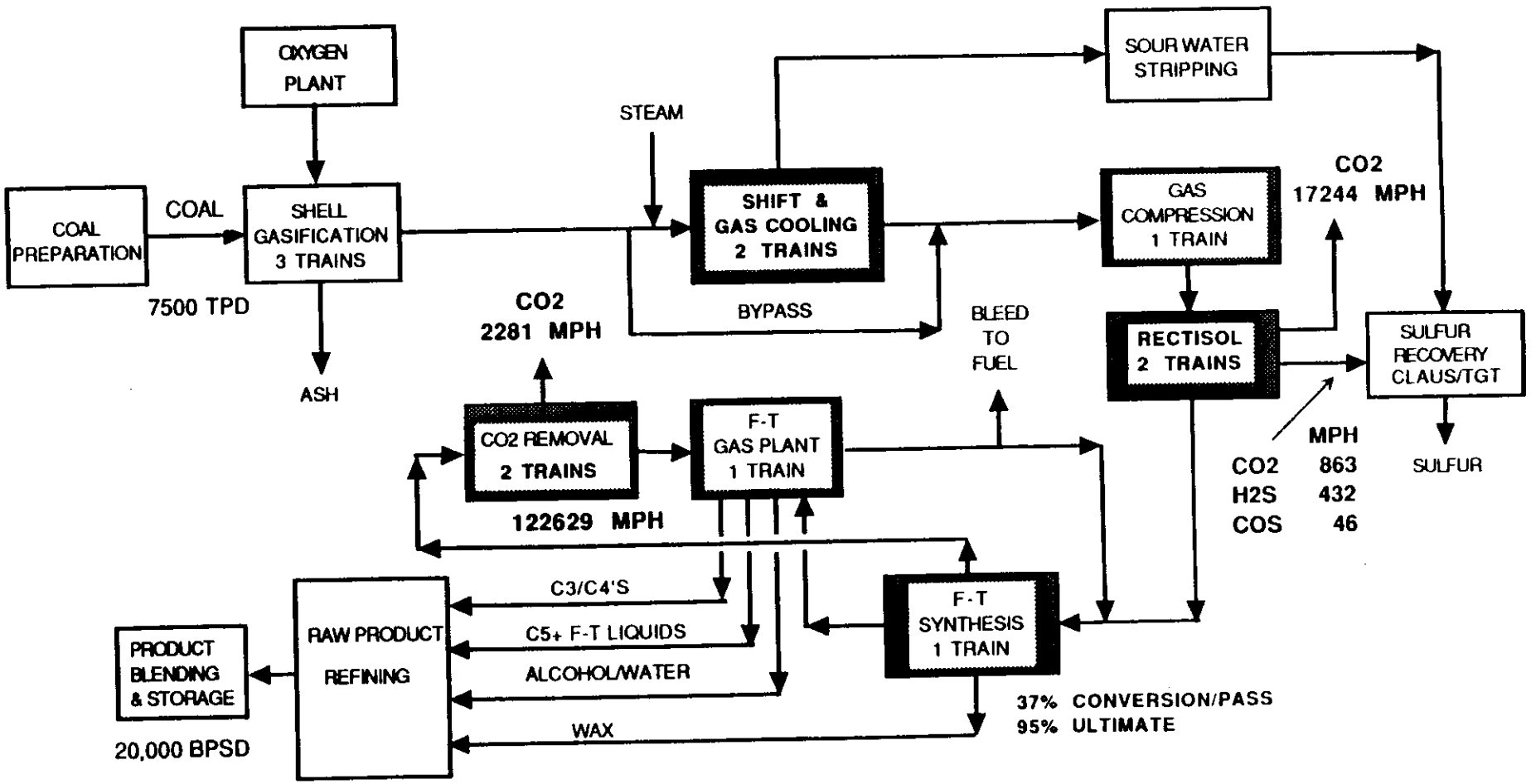


Figure 3

FISCHER TROPSCH SYNTHESIS
SLURRY REACTOR CASE
BLOCK FLOW DIAGRAM

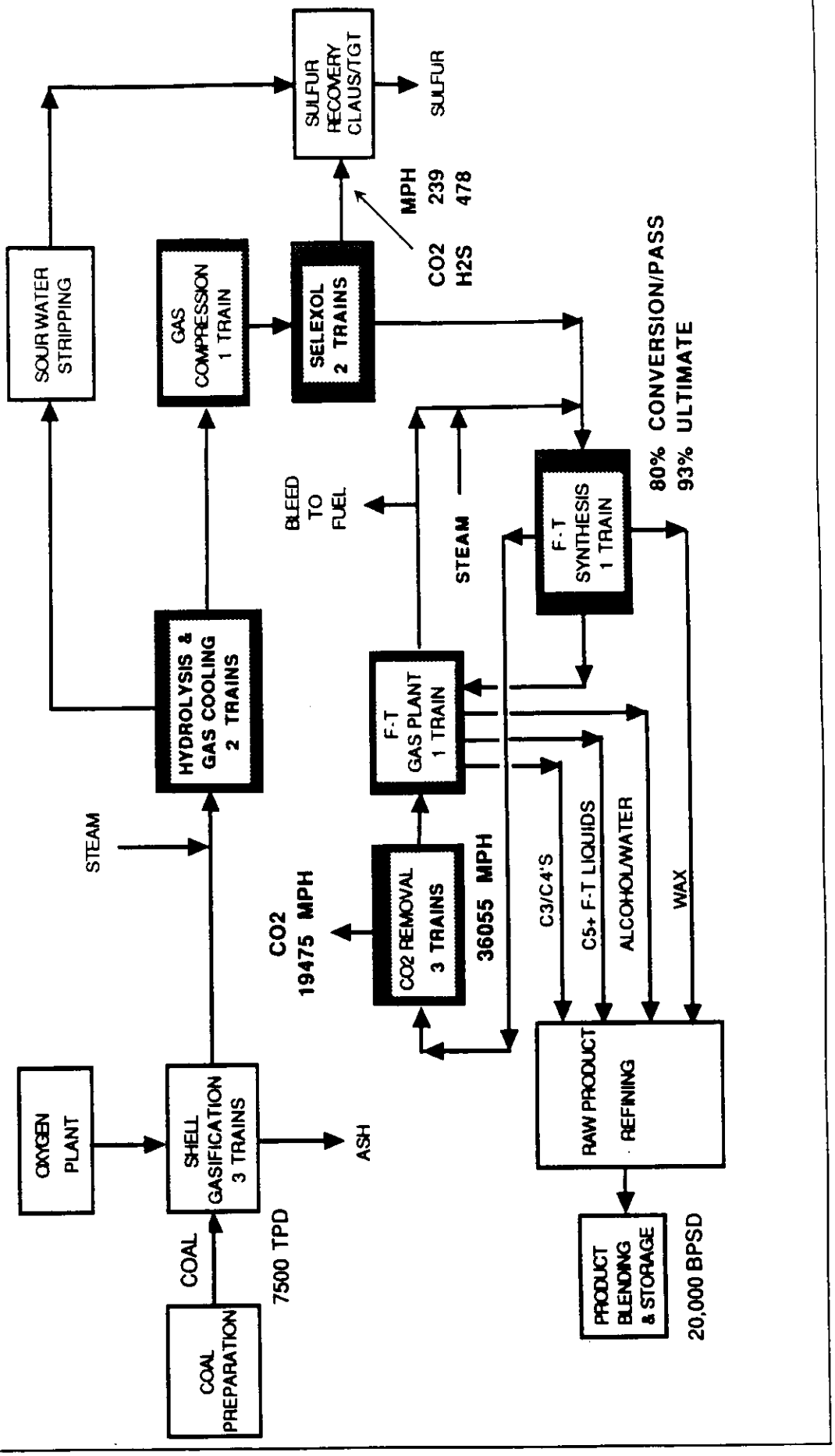


Figure 4

FISCHER-TROPSCH SYNTHESIS
SLURRY REACTOR CASE
PROCESS FLOW DIAGRAM

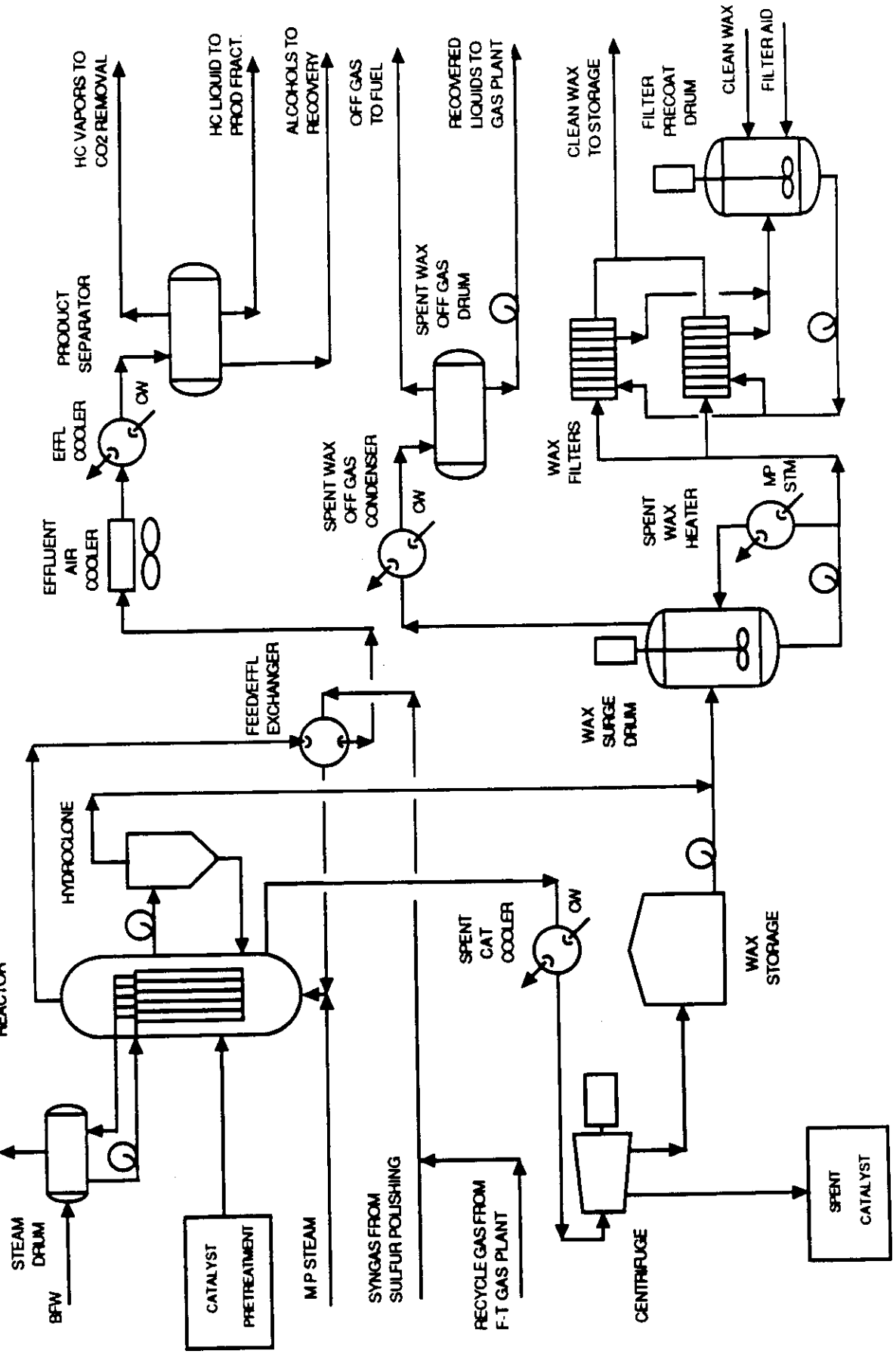


Figure 5

FISCHER-TROPSCH SYNTHESIS
FIXED-BED REACTOR CASE
PROCESS FLOW DIAGRAM

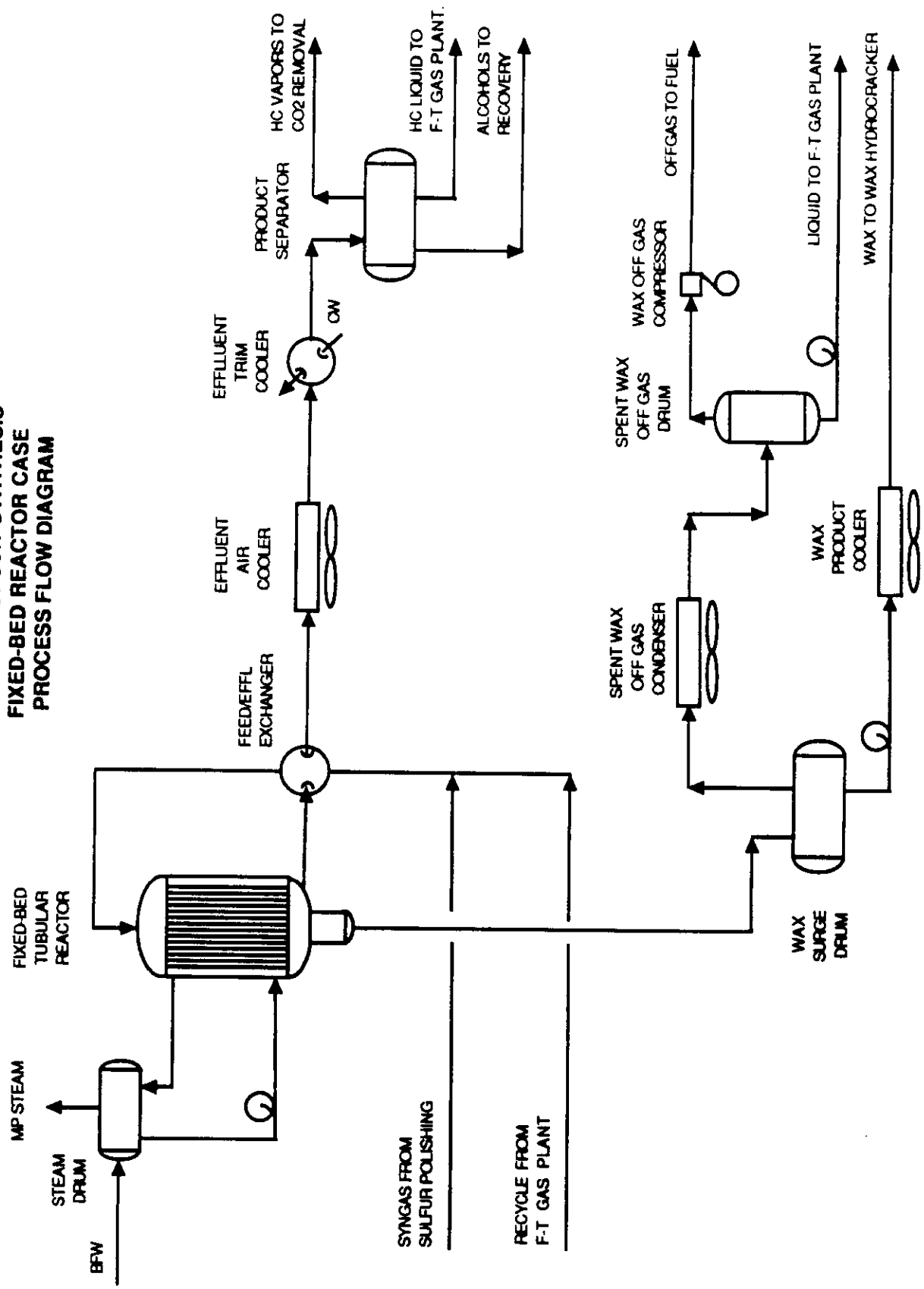


Figure 6

COMPARISON OF MODELS
HYDROGEN CONVERSION VS STANTON NUMBER

$\alpha^* = -0.5$

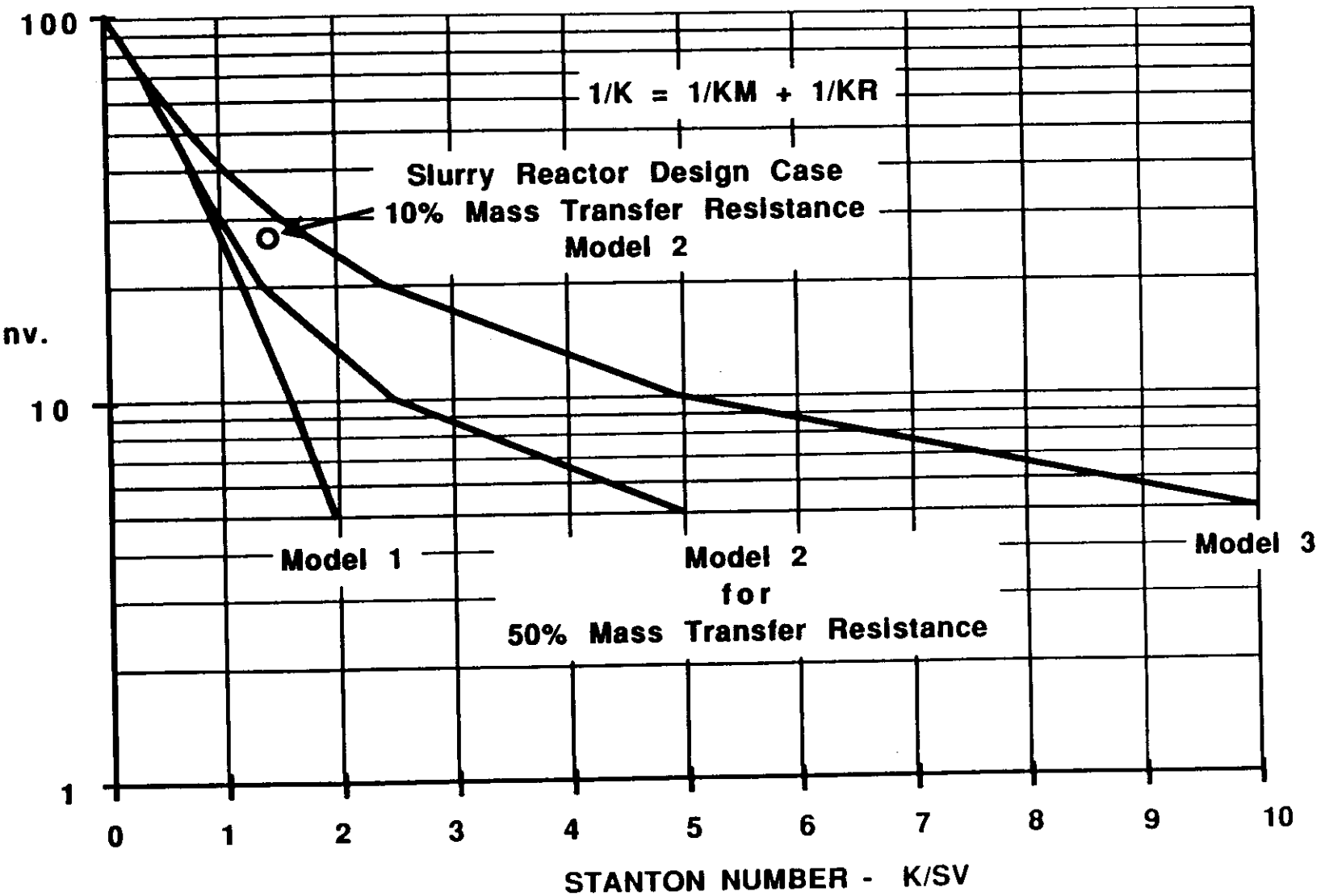
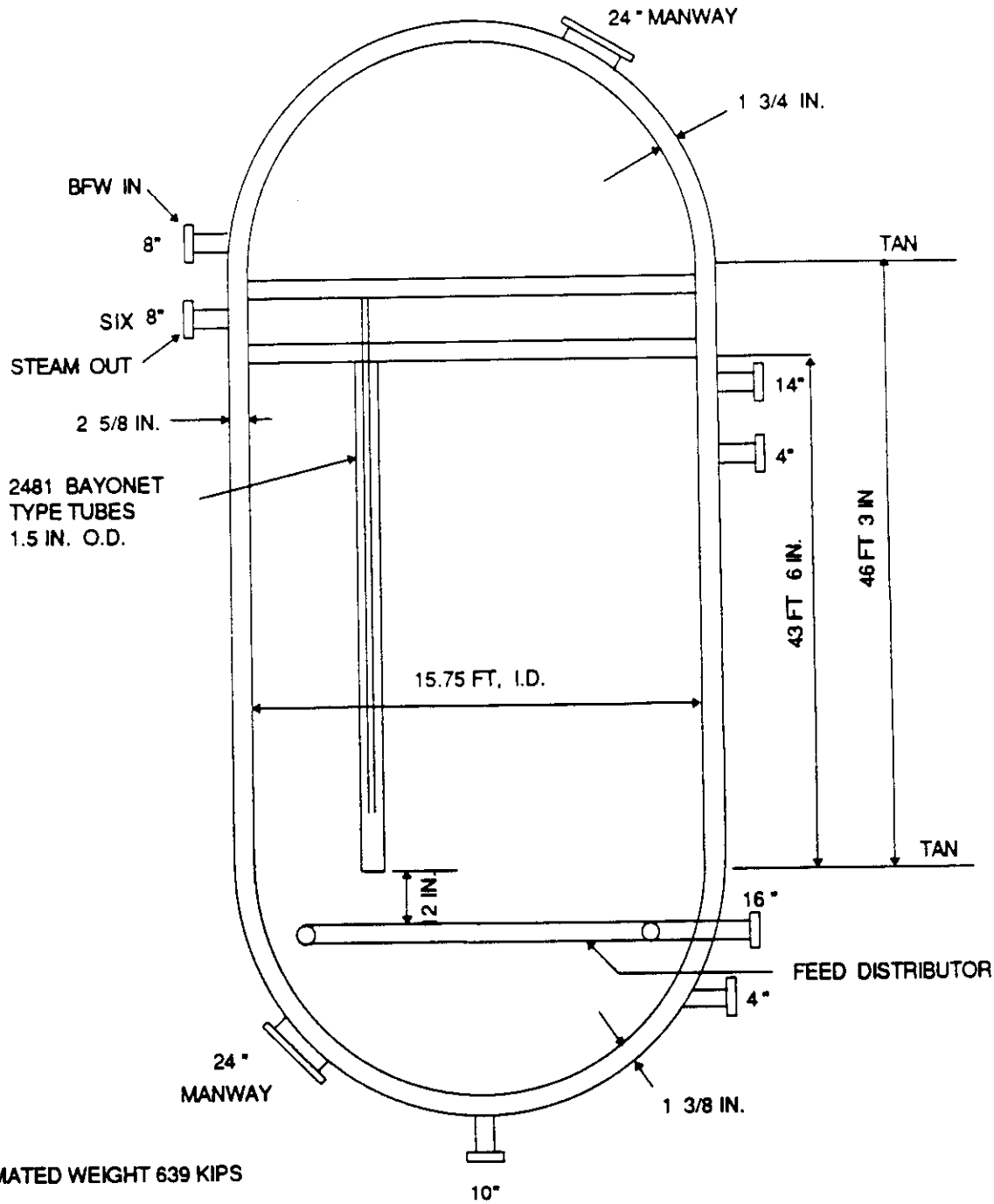


Figure 7

SLURRY REACTOR DESIGN STUDIES
FISCHER - TROPSCH PLANT
SLURRY REACTOR



NOTES:

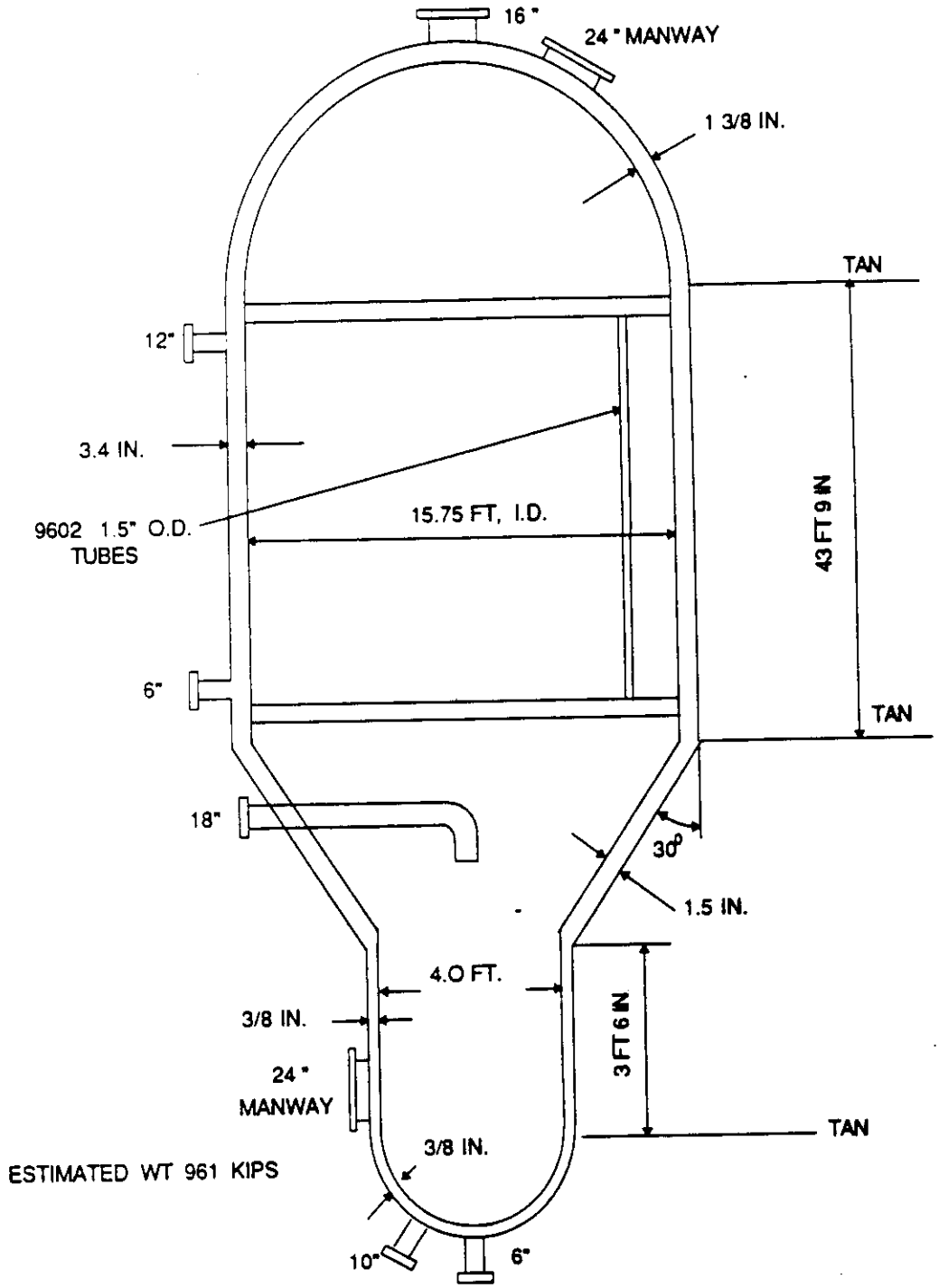
(1)DESIGN CONDITIONS- TUBESIDE 580 PSIG,550 F ; SHELLSIDE 460 PSIG 550 F

(2) METALLURGY: SHELL - SA516 GR 70 WITH 1/8" C.A. ; TUBES C.S. CHROMIZED;TUBESHEET A516 GR70 WITH 1/8" C. A.

(3) ALL NOZZLES ARE 400 LB CLASS

Figure 8

SLURRY REACTOR DESIGN STUDIES
FISCHER-TROPSCH PLANT
FIXED BED FISCHER-TROPSCH REACTOR



NOTES:
(1)DESIGN CONDITIONS- TUBESIDE 580 PSIG,550 F ; SHELLSIDE 600 PSIG 550 F
(2) METALLURGY: SHELL - SA516 GR 70 WITH 1/8" C.A. ; TUBES C.S. CHROMIZED; TUBESHEET A516 GR 70 WITH 1/8" CA
(3) ALL NOZZLES ARE 300LB CLASS