Detailed material balances for the reactor and product recovery sections are shown on Tables 44 and 45.

5. Product Quality

The estimated finished 10 RVP gasoline (including alkylate) properties are summarized below:

Research Octane, Clear	89
Motor Octane, Clear	83
RVP, Psia	10

Distillation, °F	ASTM	TBP
10 Vol. %	109	84
30 " %	139	141
50 " %	199	161
70 " %	240	197
90 " %	286	319

PONA (Vol. %)	67/13/4/16

6. Process Flow Scheme

All process flow diagrams are summarized in Appendix G. Figure G-1 (DWG. B-00242-60-0107) is a block flow diagram of the entire process plant. The process schemes for the reactor and product recovery sections are shown on Figures G-2 (DWG. B-00242-60-0104) and G-3 (DWG. 2B-00242-60-0106), respectively.

D. Plant Description

The plant consists of two main sections: (1) the reactor section and (2) the product recovery section.

1. Reactor Section

The reactor section encompasses the two stages, slurry F-T reactors followed by fixed-bed ZSM-5 reactors. In addition, special functions such as slurry filtration, catalyst activation and ZSM-5 catalyst regeneration are accomplished in smaller processing facilities that operate in a batch mode.

The clean sythesis gas is heated by the second-stage reactor effluent in heat exchanger E-3 and combined with steam before entering the SFT reactor (see Figure G-2). The synthesis gas is converted to hydrocarbons by the F-T catalyst suspended in the liquid phase. The heat of reaction in each slurry F-T reactor is removed by steam generation. The overhead reactor effluent stream includes carbon dioxide, hydrocarbon vapor products and the unconverted synthesis gas. Small amounts of high molecular weight waxy liquid hydrocarbon remain in the

Table 44
Slurry Fischer-Tropsch/ZSM-5 Process
Reactor Section Material Balance

Stream		1 Clean Syn-Gas	2 Steam	3 HC Vapor	4 HC Liquid	5 Process Water	6 Waxy Liquid
BPSD							2,386
Lb/Hr		2,111,070	121,351	2,081,460	104,340	16,314	30,260
MW		19.77	18.01	33.7	96.7	18.01	389.1
Component	MW						
CO_2	44.01	2360.0		34392.9	79.2		
C_2H_4	28.05	26.0		171.5	0.5		
CO	28.01	64319.0		5777.9	1.7		
H_2	2.01	31841.0		7791.3	1.1		
CH_4	16.04	7490.0		9293.3	7.9		
C_2H_6	30.07	341.9		727.5	3.3		
N_2	28.01	365.0		365.0	0.1		
H_2O	18.01		6738.0	19.7	0.1	905.5	
C_3H_6	42.08			354.0	4.7		
C_3H_8	44.09			432.8	6.5		
$i\text{-}C_4H_{10}$	58.12			491.4	17.0		
C_4H_8	56.10			287.7	12.4		
$n\text{-}C_4H_{10}$	58.12			269.1	13.5		
C_5^+	79.79			1296.6	930.7		
C_6^+							77.8
Total Lb-Mol/Hr		106,742.9	6,738.0	61,670.7	1078.7	905.5	77.8

Table 45
Slurry Fischer-Tropsch/ZSM-5 Process
Product and Recovery Section Material Balance

	7 HC Vapor From CO_2 Removal Unit	8 CO_2	9 Fuel Gas	10 Alkyl. Feed	11 Lt. Gasoline	12 Hvy. Gasoline	13 Dist. Blend Stock	14 Purchased i-C_4	15 Total Propanes	16 Total Butanes	17 Alkylate
BPSD					15,739	3,224	357	1,229	3,087	1,950	6,839
Lb/Hr	585,976	1,495,526	392,634	97,253	156,435	40,073	3,926	10,671	22,528	16,613	68,783
MW	20.9	44.0	15.9	51.5	83.4	125.8	167.7	57.4	43.4	58.6	103.3
Component											
CO_2	411.4	33,981.5	490.6								
C_2H_4	171.5		172.1								
CO	5,777.9		5,779.7								
H_2	7,791.3		7,792.3								
CH_4	9,293.4		9,301.2								
C_2H_6	727.5		702.7	28.1					28.1		
N_2	365.0		365.1								
H_2O			19.1								
C_3H_6	354.1		22.7	336.1							
C_3H_8	432.8		13.2	426.2				9.3	485.6		
i-C_4H_{10}	491.4			499.9	8.5			159.8	4.5	13.5	8.0
C_4H_8	287.7		296.1	296.1	4.0						
n-C_4H_{10}	269.1			278.3	4.3			16.7	0.2	261.2	33.6
i-C_5H_{12}				17.1						8.4	
C_5H_{10}				4.4						0.6	
n-C_5H_{12}				0.4							
C_5+	1,296.6				1,863.4						624.0
C_8+						318.6	23.4				
Total Lb-Mol/Hr	27,669.5	33,981.5	24,658.7	1,886.5	1,880.2	318.6	23.4	185.8	518.4	283.7	665.6

slurry reactor at the reaction condition. This excess reactor-wax is continuously removed from the reactor, separated from residual catalyst and stored.

Effluent from the slurry reactor is heated against the second-stage reactor effluent in heat exchanger E-1 and then enters the second-stage reactor where an exit temperature of 767°F is attained. After preheating the second-stage reactor feed, the second-stage reactor effluent is further cooled by generating 450 psig steam, then preheating the SFT reactor feed and by cooling water before entering the product separator. Three-phase separation of the hydrocarbon liquid, water and vapor occurs in this vessel. The water phase is sent to a waste water treatment plant outside the plant battery limit.

a. SFT Reactor

For the purpose of this study, the SFT reactors adopted are fourteen feet in diameter by thirty-five feet in length. To remove the heat generated by the reaction, the internal configuration of the reactor is similar to the Rheinpreussen-Koppers demonstration reactor. The selection of fourteen feet diameter for the SFT reactor is based on mechanial considerations foreseen in the construction of larger vessels, particularly in the design of the internal heat transfer components.

At the design feed gas rate of 2,111,070 lb/hr and a gas linear velocity of 0.3 ft/s, a total of forty SFT reactors are required.

The heat of reaction is removed by steam generation at approximately 2.5-3.0 million pounds of 450 psig steam per hour. The heat exchange elements of each SFT reactor are connected to steam drums.

To minimize the use of valves and piping, the reactors are grouped in 'clusters' of five reactors which function as a single reactor unit during normal operation. Each cluster can be taken off-line as a unit for catalyst replacement. Furthermore, the SFT reactors are arranged into two parallel trains of twenty reactors each. Figure G-4 in Appendix G shows the equipment and arrangement in the reactor section.

b. Reactor-Wax Withdrawal

A small fraction of the products in the first-stage reactor consist of high molecular weight compounds (reactor-wax) which remain in the slurry reactor. As a result, there is a continuous increase in the slurry inventory in the reactors which must be controlled without losing much of the dispersed catalyst. This is accomplished by withdrawing slurry from the reactor and

circulating it through a catalyst/reactor-wax separation unit where the slurry feed is separated into two streams. The stream with high catalyst concentration is returned to the reactor. The other stream, containing 0.5-1.0 wt % of solids, is removed from the reactor for further catalyst removal in the filtration system. The five reactors in a cluster share a common catalyst/reactor-wax separation system which consists of a circulation pump and the catalyst/reactor-wax separation unit. Figure G-5 in Appendix G is a schematic diagram of the SFT reactor clusters.

The reactor-wax filtration system consists of a holding vessel to provide surge for the catalyst-containing reactor-wax from the catalyst/reactor-wax separation unit. This vessel is equipped with an agitator to prevent settling and agglomeration of the catalyst, heating coils to maintain adequate temperature control and a overhead vapor condenser and an accumulator to recover lighter hydrocarbons flashing from the reactor-wax. The reactor-wax is then pumped through a vertical leaf type filter to remove the suspended catalyst and the clear product is sent to storage. Auxiliary equipment for filter precoating and filter cake handling are also part of the filtration system. The plant is equipped with two filtration systems, one for each train of the SFT reactors.

c. F-T Slurry Preparation and Activation

Deactivation of the F-T catalyst requiring periodic replacement of the slurry. At the end of the catalyst life, which is assumed to be 60-70 days, the slurry from each cluster of SFT reactors is transferred to a surge tank for later filtration and reactor-wax recovery. The reactors are then loaded with new slurry which has been preactivated in the slurry preparation and activation system. This system consists of an agitated and heated vessel to prepare the mix of catalyst powder with reactor-wax from storage. The slurry is then transferred to the activation vessel which is a SFT reactor equipped with a heating and cooling system using Mobiltherm-600 fluid. Once in the activation vessel, the slurry is heated to 540°F while maintaining hot nitrogen injection through the bottom to keep the catalyst in suspension and to improve the heat transfer. The hot nitrogen is supplied by a closed circuit nitrogen circulation system. On reaching 540°F the nitrogen flow is replaced by synthesis gas from the feed header to initiate the activation step. The activation is characterized by gradual increase in synthesis gas conversion. The removal of the heat of reaction is accomplished using the Mobiltherm-600 system in the cooling mode. During the activation, the effluent gas is combined with the effluent from the SFT reactors. This step lasts approximately 15-20 hours and once completed, the activated slurry is transferred to the SFT reactor cluster to start production at the normal conditions. Figure G-5 in Appendix G also shows a

schematic diagram of the reactor-wax liquid filtration, slurry activation and nitrogen circulation systems.

The effluent gas from the two trains of SFT reactors merge in a common header that provides the feed to the second-stage ZSM-5 reactors.

d. Fixed-Bed ZSM-5 Reactors

The ZSM-5 reactors are fixed-bed, downflow, adiabatic type reactors. There are four reactors on stream and sized for a WHSV of 1.65 1/hr (based on hydrocarbons) and a mass flow rate of 2500-3500 lbs/hr-ft^2 for a good flow distribution and low pressure drop. A fifth reactor permits the periodic removal of one reactor for catalyst regeneration.

The feed to the ZSM-5 reactor is preheated by exchange with its effluent since the adiabatic temperature rise provides a difference larger than 200°F between the SFT and the ZSM-5 reactors effluents. Further heat is removed from the effluent in a series of heat exchangers and coolers described previously.

e. ZSM-5 Catalyst Regeneration

Due to coke deposition on the catalyst, the ZSM-5 catalyst undergoes deactivation which requires a progressively higher reactor feed temperature to maintain the yield structure. When the inlet temperature has reached the design limit, the reactor is taken out of line for catalyst regeneration and another reactor with regenerated catalyst replaces it.

For the purpose of this conceptual design, the catalyst cycle between regeneration is thirty days and the regeneration time allowed is three days. In preparation for regeneration, the reactor is purged with nitrogen to minimize the hydrocarbons content in the bed and then is heated to combustion temperature (approximately 700-900°F) using hot nitrogen in a closed circuit. Air added to the hot recycle regenerating gas so that the combined stream contains less than 1 vol % oxygen. This limits the temperature rise in the catalyst.

The regeneration system consists of a gas circulator, regeneration gas heater, heat exchangers and compressor suction vessel.

2. Product Recovery Section

The product recovery seciton consists of a conventional distillation train to produce the gasoline product. Included in this section also are the carbon dioxide removal and the alkylation units. The vapor phase from the product separator is composed mostly of carbon dioxide, hydrocarbons, and unconverted

carbon monoxide and hydrogen. To increase the efficiency of the hydrocarbon recovery system, carbon dioxide is removed and discharged to astmosphere. A hot potassium carbonate absorption system is used for carbon dioxide removal. The hydrocarbon vapor is then cooled and compressed to combine with the deethanizer tower overhead stream.

A lean oil absorption type gas plant with a sponge absorption tower is required to maximize the proplyene recovery.

The hydrocarbon liquid from the product separator is pumped to the deethanizer tower. A light lean oil stream is combined with the deethanizer overhead vapor stream cooled by water and separated. The liquid from the accumulator is refluxed to the tower, while the vapor stream is combined with the hydrocarbon vapor from the carbon dioxide removal unit. This stream is further cooled by process streams and refrigeration and is then flashed. The liquid effluent is heated by a heat exchange and enters the deethanizer tower. The vapor stream from the flash drum enters the sponge absorption tower.

A heavy sponge oil from the gasoline splitter tower bottom is cooled and enters on the top tray of the sponge absorption tower. The vapors from the sponge tower is the fuel gas stream composed mostly of carbon monoxide, hydrogen and methane. The sponge tower bottom stream is heated and enters the deethanizer tower with the hydrocarbon liquids from the reactor section separator.

The bottom stream from the deethanizer tower is the stabilizer feed. The overhead liquid stream is the feed to the Alkylation unit. The stabilizer bottom stream is split to provide a lean oil to the deethanizer and a gasoline splitter feed.

Overhead liquid stream from the gasoline splitter is sent to gasoline blending. The bottom stream is split to provide the lean oil to the sponge tower and the gasoline fractionator tower feed. The gasoline fractionator is used to eliminate a small fraction of high boiling range hydrocarbons that will otherwise interfere with the gasoline boiling range specifications.

Stabilizer overhead liquid hydrocarbon stream enters the Alkylation process. A small amount of i-butanes is imported to supplement a requirement for alkylation. Alkylation yields used reflect typical commercial experience.

3. List of Major Equipment

Lists of major equipment for the reactor section and the product recovery section are given in Tables 46 and 47, respectively.

E. Operating Requirements

1. Utilities

Unit	Reactor Section	CO_2 Removal Unit	Compression	Product Sep.	Alkylation	Total
Steam Prod. (Consump.) Mlb/Hr						
Sat.450 Psig	2,812					2,218
	(221)		(217)(1)	(225)	(146)	(809)
Sat.50 Psig		(1,100)				(1,100)
BFW Lb/Hr	2,953					2,953
Cooling Water GPM	27,000	43,000	18,000	3,500	12,200	103,700
Power KW	252	8,950		1,450	400	11,052
Fired HTR Fuel MMBtu/Hr	327			260		587
Demin. Water Lb/Hr	300					300

(1) Assumed available as superheated at 690° in the complex.

Table 46

List of Major Equipment - Reaction Section

Service	No.	Description
SFT Reactors	41	14'-0" ø x 35'-0" T-T
ZSM-5 Reactors	5	15'-0" ø x 16'-6" T-T
Product Separator	1	17'-0" ø x 20'-0" T-T
Steam Drum (Second Stage)	1	8'-6" ø x 20'0" T-T
Steam Drum (First Stage)	2	12'-0" ø x 54'-0" T-T
ZSM-5 Rx Fired/Effluent Exch.	4	42.0 MM BTU/HR each
ZSM-5 Rx Effluent/Steam Gen.	4	14.0 " " "
" " " /Syn.Gas Feed	4	73.0 " " "
" " " /Cooler Condensers	4	8.0 " " "
Slurry Filtration System	2	System includes: Surge Vessel, Filtration Equipment and Filter Cake Handling Facilities
Slurry Activation System	1	System includes: Slurry Preparation Vessel, Activation Reactor (SFT), Mobiltherm System
Nitrogen Circulation System	1	System includes: Nitrogen Circulator, Fired Heater Exchangers and Separator Vessel
ZSM-5 Catalyst Regeneration	1	System Includes: Air Compressor, Regen. Gas Circulator, Fired Heater, Exchangers and Separator

Table 47

List of Major Equipment - Product Recovery Section

Service	No.	Description
Carbon Dioxide Removal	1	Four parallel Trains of Potassium Carbonate Absorption Towers with Associated Regeneration
Compressor	1	Centrifugal compressor 16,000 HP
Deethanizer Absorber	1	7'-6" ø x 16'0" ø x 120' T-T With Fired Reboiler and Overhead Condenser
Sponge Absorber	1	9'-6" ø x 60' T-T
Stabilizer Tower	1	12'-6" ø x 105' T-T With Steam Reboiler and Water Cooled Overhead Condenser
Gasoline Splitter Tower	1	14'-6" ø x 65' T-T With Steam Reboiler and Water Cooled Overhead Condenser
Gasoline Fractionator Tower	1	7'-6" ø x 60' T-T With Fired Reboiler and Air Cooled Overhead Condenser
Alkylation Plant	1	6900 BPSD Alkylate Unit

2. Initial Catalyst and Chemicals Requirements

F-T Catalyst, Lbs.	1,000,000
ZSM-5 Catalyst, Lbs.	400,000
Potassium Carbonate Solution, Gal.	500,000

3. Operating Manpower

	Total
SFT/ZSM-5 Reactor Section	32
Carbon Dioxide Removal, Gas Plant and Alkyl. Unit	19
TOTAL	51

F. Scoping Cost Estimate

The cost of the battery limits facilities is estimated to be 700 million dollars based on July 1983 and Wyoming location. This estimate does not include coal gasification and gas cleanup facilities, utilities and offsites, SFT catalyst manufacture facilities, and catalyst fills and royalties.

The facilities included are the following:

- SFT and ZSM-5 reactor section
- Carbon dioxide removal unit
- Alkylation unit
- Product recovery section
- SFT catalyst slurry filtration system
- SFT catalyst slurry activation system
- ZSM-5 catalyst regeneration and nitrogen circulation systems

The cost contribution of each plant section as percent of the total investment is approximately as follows:

	Percent
Reactor section	70
Carbon dioxide removal	20
Alkylation	5
Product recovery	5

Included in the 700 million dollars are equipment, bulk materials, labor, field indirects, contractor engineering and fees, owners engineering and project management costs and capitalized spares. Because of the Wyoming location, an allowance for a construction workers' camp is also included. The investment estimate has been developed using the Rand Corporation guidelines (Merrow, et al., 1980) for calculating project cost growth factors. The cost growth factor is estimated at 0.59. This is equivalent to a 69% cost increase as a contingency for a first-of-a-kind plant.

The effects of design conditions and yields structure changes on the investment were investigated to develop a qualitative understanding of their impact. The results are summarized below:

- A decrease of approximately twenty percent in the yield of the methane and ethane in favor of either higher reactor-wax or higher gasoline range products will not result in appreciable changes in plant investment.

- Elimination of the F-T slurry catalyst activator and its related equipment will result in a reduction of less than ten percent of the total investments.

- Larger SFT reactors (i.e., 20-25 ft ID) will decrease significantly the number of reactors required. However, the total investment for reactors could be higher if they have to be field-fabricated. In addition, the designs and fabrication methods for the internal heat transfer elements are not well studied.

- Higher operating pressure (i.e., 350 psig) in the SFT reactor section is expected to cause an increase of less than ten percent of the total investment as a result of increased material cost for the section. The number of reactors required is reduced proportionally to the pressure increase; however, the height must be increased to accommodate larger heat transfer area requirements. The possibility of increasing the heat transfer area per reactor volume was not investigated.

XII. Recommendation for Further Study

Based on studies carried out in this Contract, areas of major importance have been identified. Many of those areas are recommended for further study to improve the process economics, the product flexibility, and the scaleup of the slurry F-T reactor. For convenience, thses areas of interest are summarized below into four groups:

Slurry F-T Operation

- Low methane + ethane operation.
- Scaleup factors of slurry F-T reactor, including type, maximum size, internals, and hydrodynamics.
- Steam co-feeding to slurry F-T reactor to allow use of a synthesis gas of H_2/CO ratio lower than the usage ratio.
- Higher pressure operation and its effect on process performance.
- Slurry F-T catalyst activation.
- Effect of varying catalyst loading in SFT reactor on the reactor performance.
- Continuous removal and replacement of F-T catalyst to maintain a constant activity in the F-T reactor.

F-T Product Upgrading

- F-T catalyst/reactor-wax separation.
- Means to upgrade F-T reactor-wax into high quality distillate and gasoline.
- Effect of low methane + ethane mode operation on the performance of the second-stage ZSM-5 reactor.

Process Optimization

- Carbon dioxide removal schemes and their utility integration with other part of the plant.
- Schemes for further conversion of unconverted H_2+CO.

- Schemes for recycle of C_1+C_2 hydrocarbons to the slurry F-T reactor via a steam reformer or a partial oxidation unit.

- Schemes for further conversion of unconverted light olefins.

- Alternate distillation schemes for more efficient separation of heavier boiling hydrocarbons in the gasoline.

- Examining benefits or penalties of lower single-pass H_2+CO conversion with recycle to the slurry F-T reactor or with the use of multi-staged F-T reactors.

Process Economics

- Performing conceptual process design and scoping cost estimate of maximum distillate + gasoline mode operation.

- Detailed economic comparison of this technology against the best alternative.

Further discussions of some of these areas are given below.

Low methane + ethane mode operation is aimed at maximizing the liquid fuel yield and is therefore worthy of further investigation. The high F-T reactor-wax yield from this operation provides a possibly new route for distillate production. Such a route was proposed by M. E. Dry of SASOL (Dry, 1982). He reported that a high yield (80 wt %) of high quality diesel (65 cetane number) was obtained by mild hydrocracking of a F-T wax obtained from SASOL's fixed-bed tubular F-T reactor (Arge Process). The light hydrocarbon (C_1 to C_4) yield was only 5 wt %. Because of the expected high reactor-wax yield, the separation of the reactor-wax from the F-T catalyst is apparently an important problem.

Another area of major importance is the commercial scaleup of the slurry F-T reactor. The factors that need to be evaluated include the type and size of the reactor, its internals (buffers and/or heat transfer tubes), feed-gas distributor; gas bubble size, gas holdup, and the liquid- and gas-phase back-mixing. Although a high synthesis gas conversion was demonstrated by the well-known Rheinpreussen-Koppers demonstration plant (1.55 m ID x 8.6 m height with internal steam-generation tubes), simulation of such a performance by examining the important hydrodynamic factors in a large-diameter hot-flow reactor model is highly desirable.

Various process optimization schemes need to be examined. In any coal-to-hydrocarbon plant, the amount of carbon dioxide to be removed is directly proportional to the inefficiency of the plant. The amount of carbon dioxide to be removed in the two-stage slurry F-T/ZSM-5 plant is expected to be relatively small because of its high thermal efficiency. However, the investment associated with the carbon dioxide removal is still quite substantial. Examination of other removal schemes and their utility integration with other parts of the plant is warranted. Other possible optimization schemes are further conversion of unconverted H_2+CO and light olefins, and the recycle of methane + ethane to the slurry F-T reactor via a steam reformer or a partial oxidation unit. All these schemes will contribute to higher liquid hydrocarbon yield.

With maximum distillate + gasoline mode operation, an additional investment will be required for the upgrading of the reactor-wax. However, the final product value is also expected to be higher. Therefore, the conceptual process design and scoping cost estimate for this operation must be updated. Finally, to determine the priority on the development of various routes of coal-to-liquid fuel projects, a detailed economic comparison of this technology against the best alternate route of making similar products should be conducted.

XIII. Nomenclature

a_c	Catalyst particle external surface area per slurry volume, $6C_c(1-v_c)/\rho_c d_c$, (cm^2 solid-liquid area/cm^3 slurry)
a_g	Gas bubble interfacial area, $6\epsilon_g/d_B$, (cm^2 gas-liquid area/cm^3 expanded slurry)
a_i	Coefficients, i=1,2,--, 2(N+2), given in Equation (27)
B_1	A function of Pe_c and B_2, defined as Equation (20)
B_2	R_k/R_d
C	Concentration, (mol/cm^3 liquid or gas)
C*	Concentration at gas-liquid interface, (mol/cm^3)
C_c	Catalyst loading, ($gCat/cm^3$ liquid)
$\bar{C}_c$	Dimensionless catalyst loading, C_c/C_{ca}
C_{ca}	Average catalyst loading, ($gCat/cm^3$ slurry)
C_{Fe}	Iron loading, (gFe/cm^3 liquid)
$\bar{C}_g$	C_g/C_{gl}^i
C_{Hs}	Liquid phase H_2 concentration inside catalyst, (mol/cm^3 liquid)
$\bar{C}_L$	$C_L K/C_{gl}^i$
d_B	Bubble diameter, (cm)
d_c	Catalyst particle diameter, (cm)
d_R	Reactor diameter, (cm)
E	Axial dispersion coefficient, (cm^2/s)
f	Molar H_2/CO ratio at reactor inlet
f_{Fe}	Weight fraction of Fe in catalyst
g	Gravitational constant, 981, (cm/s^2)
I	Carbon number

K	Solubility coefficient C_g^*/C_L^*, (cm^3 liquid/cm^3 gas)
k_1, k_2	Intrinsic kinetic rate constants for F-T and water-gas shift reactions, respectively, used in Equations (29) and (31), (cm^3 liquid/s-gFe)
$\bar{k}_2$	k_2K_1/K_4, (cm^3 liquid/gFe-s)
k_3, k_4	Constants used in the rate expressions (29) and (31)
$\bar{k}_3$	k_3K_2/K_4
$\bar{k}_4$	$k_4K_1K_3/K_2K_4$
k_c	Liquid-particle mass transfer coefficient, (cm/s)
k_g	Gas side mass transfer coefficient, (cm/s)
k_H''	Intrinsic kinetic rate constant for H_2 conversion, $r_H/(1-\epsilon_g)(1-v_c)C_{HL}C_{Fe}$, ($cm^3$ liquid/s-gFe)
k_L	Liquid side mass transfer coefficient, (cm^3 liquid/ s-(cm^2 gas-liquid area))
L	Bubble-column height, (cm)
M_I	Weight fraction of the I carbon-number hydrocarbon
m	Average H/C atomic ratio of F-T products
N	Number of interior collocation points
P	Pressure, (Pa)
$P_j(\bar{z})$	Jacobi polynomials, j=1,2--, defined as Equation (24)
R_c	H_2 transport resistance, from bulk liquid phase to liquid-solid interface, $K_H/k_ca_c(1-\epsilon_g)$
R_d	H_2 transport resistance from gas-liquid interface to bulk liquid phase, K_H/k_La_g, (s-cm^3 expanded slurry/ cm^3 gas)
R_G	Gas law constant, 8.2, (MPa-cm^3)/(mol-°K)
R_g	H_2 transport resistance, from bulk gas phase to gas-liquid interface, $(k_ga_g)^{-1}$, (s)
R_k	Kinetic resistance, $K_H/k_H''C_{Fe}(1-\epsilon_g)(1-v_c)$ for single component model, $K_1/\bar{k}_jC_{Fe}(1-\epsilon_g)(1-v_c)\bar{C}_c$ (j=1,3) for multi-component model, (s-cm^3 expanded slurry/cm^3 gas)

r_H	H_2 conversion rate, (mol/s-gFe)
r_1	Kinetic rate of F-T reaction, given as Equation (29), (mol/s-gFe)
r_2	Kinetic rate of water-gas shift reaction, given as Equation (31), (mol/s-gFe)
$\bar{r}_1$	$\bar{C}_{L1}\bar{C}_{L2}/(\bar{C}_{L2} + \bar{k}_3\bar{C}_{L4})$
r_2	$(\bar{C}_{L2}\bar{C}_{L4}-\bar{C}_{L1}\bar{C}_{L3}/\bar{k}_4)(\bar{C}_{L2} + \bar{k}_3\bar{C}_{L4})$
S_{ij}	Elements of stoichiometric matrix, i = 1,...,4; and j = 1,2
T	Temperature, (°C)
U	Molar 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)
$\bar{u}_g$	$u_g/u_g{}^i$
v_c	Volumetric fraction of catalysts in slurry, $\rho_L w_c/\rho_s + w_c(\rho_L-\rho_s)$), ($cm^3$ catalyst/cm^3 slurry)
$v_c{}^*$	v_c as C_c=0.1 gCat/cm^3 slurry, (cm^3 catalyst/cm^3 slurry)
w_c	Weight fraction of catalyst in slurry, (gCat/g slurry)
w_{Fe}	Weight fraction of Fe in slurry, (gFe/g slurry)
X_H	H_2 conversion
X_{H2+CO}	Molar H_2+CO conversion
$\bar{x}$	Dimensionless liquid-phase H_2 concentration, $K_H C_{HL}/C_{Hg}{}^i$
Y	$(1-\bar{x})/(1 + \alpha^*\bar{x})$
y	Gas-phase H_2 mole fraction
$\bar{y}$	Dimensionless gas-phase H_2 mole fraction, y/y^i
Z	$X_H{}^e/St_k$
z	Axial reactor distance, (cm)
$\bar{z}$	Dimensionless axial reactor distance, z/L

Greek Letters

α	Contraction factor, molar contraction per mole of H_2+CO converted, defined in Equation (13)
α'	Probability of the chain-growth
α^*	$\alpha f(1+U)/U(1+f)$
γ_{ij}	Coefficients for Jacobi polynomials, i,j = 1,2, , defined by Equation (25)
ϵ_g	Gas holdup, (cm^3 gas/cm^3 expanded slurry)
ρ	Density, (g/cm^3)
ρ_c	Catalyst particle density, (gCat/cm^3 catalyst particle)
ρ_s	Catalyst solid density, (gCat/cm^3 catalyst solid)
μ	viscosity, (g/s-cm)

Dimensionless Numbers

Ar	Archimedes number, $\rho_L(\rho_c-\rho_L)gd_c/\mu_L^2$
Fr	Froude number, $u_{gm}/(gd_R)^{0.5}$
Pe_c	Axial Peclet number (catalyst particle), $u_{cs}L/E_c$
Pe_L	Axial Peclet number (liquid), $u_g^iL/E_L(1-\epsilon_g)(1-v_c)$
Re_c	Reynolds number (catalyst particle), $u_{ct}d_c\rho_L/\mu_L$
Sc	Schmidt number (liquid), $\mu_L/\rho_L D_L$
Sh	Sherwood number (liquid), $k_c d_c/D_L$
St_d	Stanton number (diffusion resistance), $L/u_g^iR_d$
St_k	Stanton number (kinetic resistance), $L/u_g^iR_k$

Acronyms

AD	Axially Dispersed liquid phase model
BPSD	Barrels Per Stream Day
BSU	Bench-Scale Unit
DOS	Days on Stream
DP	Differential Pressure
F-T	Fischer-Tropsch
GC	Gas Chromatography
GHSV	Gas Hourly Space Velocity
FS-GC	Fused Silica capillary Gas Chromatography
HOS	Hours on Stream
LC	Liquid Chromatography
MS	Mass Spectrometry
MTG	Methanol-To-Gasoline
NM	axially Non-Mixed liquid phase model
PM	axially Perfectly-Mixed liquid phase model
SASOL	South African Coal, Oil and Gas Corporation, Ltd.
SCFD	Standard Cubic Feet per Day
SFT	Slurry Fischer-Tropsch
STP	Standard Temperature and Pressure
STY	Space-Time-Yield, (gMol H_2+CO converted/hr-cm^3 expanded slurry)
SV	Space Velocity, (NL/gFe-hr)
WHSV	Weight Hourly Space Velocity, (1/hr)

Superscripts

i	At reactor inlet
e	At reactor exit

Subscripts

c	Catalyst
esl	Expanded slurry
g	Gas
H	Hydrogen
i	Components, i = 1,2,3,4 for H_2, CO, CO_2, H_2O, respectively
L	Liquid
m	Arithmetic mean value of that at reactor entrance and that at reactor exit
sl	Slurry