<u>Design/Economics of a Once-Through Natural Gas Fischer-Tropsch Plant With Power Co-Production</u>

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INTRODUCTION

In 1994, Bechtel, along with Amoco as the main subcontractor, developed a Baseline Design (and an ASPEN Plus computer process simulation model) for indirect coal liquefaction using advanced Fischer-Tropsch (F-T) technology to produce high-quality, liquid transportation fuels. This was done under DOE Contract No. DE-AC22-91PC90027. In 1995, the original study was extended to develop a case in which natural gas, instead of coal, is used as the feedstock.[1] The results, presented at last year's Contractors' Conference, show that a natural gas F-T plant is less capital intensive, and as a consequence, attractive F-T economics may be attained with low cost remote gas.[2]

In 1996, the natural gas based F-T study was extended to develop a conceptual design for a once-through F-T plant with power co-production using remote gas as the feedstock. This once-through facility is designed to reduce the initial cost at the expense of a small sacrifice in overall thermal efficiency. In addition, the product upgrading area is simplified to produce only shipable F-T liquids. This paper describes the results of this study. It discusses the overall plant design, preliminary economics, and some site-specific situations under which additional capital cost savings can be realized.

DESIGN BASIS

The plant is designed to process 100 MMSCF/day of natural gas and produce roughly 8,820 BPD of F-T syncrude and 84 MW of export power. For design purposes, a typical natural gas composition containing about 95 mole% methane was assumed. The F-T liquefaction plant design is based on the Syncrude Technology Incorporated (STI) cobalt-based, F-T catalyst. In this design, the H_2/CO ratio and the partial pressures of H_2 and CO in the syngas are important for proper F-T synthesis reactor performance. An autothermal reformer using enriched air with partial CO_2 recycle is used for syngas preparation. This design was arrived at after studying several alternatives, including air-blown autothermal reforming operating at low steam/carbon feed ratio. After satisfying the internal fuel and heating requirements, the plant sends all of the byproduct steam and tail gas to a combined-cycle plant for air compression and power generation.

The plant is assumed to be located at a hypothetical U. S. Gulf Coast site, a common basis for such studies. For guidance as to how to relate such costs to remote sites, see the earlier natural gas based study[2]. In developing this case, individual plant designs and cost estimates were prorated, where applicable, from the previous coal and natural gas based deisgns[1, 2].

OVERALL PLANT CONFIGURATION

<u>Syngas Generation and F-T Synthesis</u>
Figure 1 is a simplified block flow diagram showing the process configuration of the synthesis section. This portion of the plant consists of two main processing areas; Area 100, synthesis gas preparation, and Area 200, once-through F-T synthesis and product fractionation. This portion of the plant was simulated using ASPEN Plus. In the F-T synthesis plant, the ASPEN Plus model was modified to duplicate the results of STI's reactor model.

<u>Combined Cycle Power Plant</u>
Figure 2 is a simplified block flow diagram showing the combined-cycle plant which uses a General Electric (GE) Frame 7 gas turbine. In addition to producing power, this gas turbine drives the air compressors for the air separation plant. The 635

psig steam from the ATR and the 150 psig steam from the F-T synthesis plant are superheated and imported directly into the steam turbines. Overall performance of the combined-cycle plant was calculated using the GateCycle power cycle simulation software (Enter Software, Menlo Park, CA).

Outside Battery Limits (OSBL) Plants

There are eighteen ancillary offsite plants, ranging from water treatment to interconnecting piping, which are similar to those which were developed for the coal-based Baseline Design and were modified, as required, for this natural gas case.[3]

SYNTHESIS GAS PREPARATION (AREA 100)

This area consists of three major plants; air compression and separation, autothermal reforming (ATR), and CO_2 removal and recycle. The objective of Area 100 is to produce a nitrogen-diluted syngas with a molar H_2/CO ratio of 2.0. Sulfur is removed from the natural gas before syngas generation by adsorption on ZnO. The H_2 and CO partial pressures entering the F-T slurry-bed reactors also were governing factors in the design of the syngas preparation area. Autothermal reforming calculations/trade-off studies were carried out to determine that the syngas requirements could best be met with an ATR using 40% enriched air operating with CO_2 recycle. The ATR reactor is designed to operate at a reasonable 1) O_2/C ratio of 0.7 to keep the maximum adiabatic flame temperature below 4000 ^oF, and 2) steam/C ratio of 0.6 to avoid potential soot formation. The final syngas preparation design was confirmed by Lurgi Corporation.[4]

Plant 101, the air compression and separation plant, supplies enriched air containing 40 mole% oxygen at 650 psig to the ATR. The 95 mole% O_2 product from cryogenic air separation is diluted with bypassed, compressed air to produce the required enriched air feed.

FISCHER-TROPSCH SYNTHESIS AND PRODUCTION FRACTIONATION (AREA 200)

This area consists of four plants; once-through F-T synthesis, product separation, hydrogen recovery and wax hydrocracking. The wax hydrocracking design is similar to that used in the previous F-T studies. Its objective is to ensure that the produced F-T product is pumpable at ambient conditions. A small hydrogen recovery (PSA) plant is used to recover enough hydrogen for hydrocracking. The PSA plant feed is a portion of the syngas leaving Area 100 rather than the unconverted syngas from the F-T synthesis plant, as was done in the previous recycle designs, because the hydrogen concentration in the F-T effluent is too dilute for effective hydrogen recovery. The effluent from the PSA plant goes to the F-T synthesis reactors.

A total of three, 6 meter diameter by 20 meter high, slurry-bed reactors are required to process the syngas from Area 100. These reactors are arranged so that two parallel first-stage reactors feed a single second-stage reactor. The unconverted syngas leaving the first-stage is cooled to 150 ^oF and flashed to condense and remove liquids before being reheated and fed to the second-stage. The CO conversion in each of the F-T reactors is about 61% with an overall CO conversion of about 85%. The first-stage F-T slurry reactors operate at about 430 ^oF and 490 psig, and the second-stage reactor operates at about 430 ^oF and 440 psig. Excess heat is removed by the generation of 150 psig steam from tubes within the reactors.

The design is based on STI's specified F-T kinetics and product distribution. The slurry-bed reactors are sized in accordance with STI's guidelines and include STI's proprietary catalyst/wax separation technology. Overall reactor weight and cost were determined using Bechtel's slurry-bed reactor design model.[5]

The hydrocarbon product recovery section consists of a conventional fractionation column with a side stripper to recover a naphtha stream, a distillate stream and a wax stream which goes to the wax hydrocracking plant. The column overhead stream goes to a LiBr absorption chiller, using waste heat, to recover additional hydrocarbons. Plant 204, the wax hydrocracking plant, is similar to that of the previous cases except that the fractionation system has been simplified to produce only butanes, naphtha and distillate products. The light gases also go to the LiBr cooled chiller to maximize hydrocarbon recovery.

PLANT SUMMARY

Process Design Basis

The plant processes 100 MMSCF/day of natural gas and produces about 8,820 BPD of liquid F-T products; namely a C_5-350 ^oF naphtha, a 350-850 ^oF distillate, and a small amount of butanes. Both the naphtha and distillate are essentially free of sulfur and nitrogen containing compounds. The naphtha is a raw product which either can be upgraded to produce a high-quality gasoline blending stock or used as a feedstock for steam cracking to produce light olefins. The distillate is a high-quality diesel blending stock which requires no further upgrading. The plant uses all the byproduct steam and fuel gas to supply its internal electric power and heating requirements. In addition, it produces about 84 MW of exported electric power. The only materials delivered to the plant are natural gas, raw water, catalysts and chemicals. The major feed and product streams entering and leaving this plant are shown in Table 1.

Following the philosophy of the indirect coal liquefaction study, the overall plant is designed to comply with all applicable environmental, safety and health regulations. Air cooling is maximized, wherever possible, in order to minimize cooling water usage.

Capital Cost Estimate

Total capital cost of this natural gas F-T plant is about $415 MM (mid 1996 dollars). This cost includes offsites and allowances for home office costs, service fees and contingency. Since the exported power is a major product, the combined-cycle plant is considered to be one of the ISBL plants. Table 2 shows a breakdown of the capital cost. Area 100, the syngas preparation area, is about half of the total ISBL cost with the remaining half being equally divided between Area 200, the F-T synthesis and product recovery area, and Plant 31, the combined-cycle power plant. The air compression and separation (Plant 101) and combined-cycle (Plant 31) plants are the two most expensive ISBL plants, and together, they constitute 58% of the ISBL cost. The estimated cost of this once-through F-T plant is about a third less than that of a F-T plant of the same size using gas recycle to maximize liquid production.

ECONOMIC SENSITIVITY STUDIES (PRELIMINARY)

A discounted-cash-flow analysis on the production cost of the F-T products was carried out 1) to compare this once-through, F-T power co-production design with that of the previous natural gas design which uses gas recycle for maximum liquid production, and 2) to examine the economics of this once-through case for remote gas using the same methodology and financial assumptions employed in the F-T baseline study.[2] As in the previous studies, the results are expressed in terms of a crude oil equivalent (COE) price which is defined as the hypothetical break-even crude oil price where the F-T liquefaction products are competitive with products made from crude oil.

The primary liquefaction products consist of about 1/3 raw naphtha and 2/3 distillate. The distlllate is a high-quality diesel blending stock which requires no further upgrading. A past linear programming study of a typical PADD II refinery indicated the F-T distillate can command a premium of 7.19 $/bbl above and beyond the value of conventional crude oil.[6] For the following analysis, it was assumed that the raw naphtha has the same value as that of crude oil.

The comparison between this once-through power co-production design with the previous gas-recycle case is carried out using 0.50 $/MMBtu gas, estimated 1993 costs and 1996 EIA escalation factors. Figure 3 shows the results in terms of COE vs. the electricity selling price. The original recycle plant design processes about 412 MMSCF/day of gas feed and has a COE price of 19.1 $/bbl.[1,2] The COE has only a slight dependence on the electricity selling price since the plant exports only a small amount of power. When the plant cost is capacity-factored to process 100 MMSCF/day of gas feed, the COE increases to about 23.0 $/bbl - approximately a 20% increase just due to the loss in the economy of scale.

The once-through F-T power co-production plant produces 84 MW of excess power for sale, and Figure 3 shows that its economics are very sensitive to the selling price of the co-produced power. Typical base load power plants sell electricity between 4 to 6 cents per kWh. Even at a conservative power selling price of 3 cents per kWh, the plant has a COE price of 18.8 $/bbl. This COE is 22% less than that of the scaled down recycle gas case which maximizes liquids production.

Figure 4 shows the portion of the calculated COE price attributable to various components for this once-through plant with 0.50 $/MMBtu gas. Capital servicing costs account for over one-third of the COE price. Even with 0.50 $/MMBtu gas, the feedstock cost still constitutes about one-quarter of the total COE price. The 34% federal income tax rate constitutes about one-fifth of the total COE price.

CONCLUSIONS AND RECOMMENDATIONS

A conceptual plant design and cost estimates have been developed for a once-through, Fischer-Tropsch liquefaction plant with power co-production producing about 8,820 BPD of F-T liquids and 84 MW of power from 100 MMSCF/day of natural gas. The plant is estimated to cost about $415 MM mid-1996 dollars. The economics are very sensitive to both the purchased gas price as well as the selling price of the co-produced power. With 0.50 $/MMBtu gas and a reasonable electricity selling price of 3 cents per kWh, the above plant will produce liquid transportation fuels which will be competitive with those produced from crude oil priced at about 18.8 $/bbl.

Capital servicing costs still play a dominant role in driving the overall process economics. Optimization and site-specific studies are needed to explore various opportunities to simplify the overall process and reduce cost. Examples of such optimization include:

- The arbitrarily chosen design basis of 100 MMSCF/day of gas feed should be adjusted for site-specific situations. Studies should be made to select an appropriate plant size which takes advantage of the economies of scale to optimize either the air separation and compression plant and/or the combined-cycle plant, the two most expensive sections of the plant.
- The use of sub-quality gas containing CO_2 should be considered as a prime feedstock for this purpose (e.g., Alaska Prudhoe Bay associated gas has 10 to 15 mole % CO_2). The CO_2 content of such gas might allow production of a syngas with a H_2/CO ratio of 2.0 without requiring the separation and recycle of CO_2 around the ATR. Furthermore, this type of gas also might allow the use of an air-blown ATR which will eliminate the costly air separation plant at the expense of somewhat higher pressure slurry-bed F-T reactors and a more elaborate product recovery scheme to compensate for the higher inert concentration in the unconverted syngas.
- A minor simplification of the recovery scheme could be achieved if the plant were designed to produce a stabilized, pumpable syncrude instead of separate naphtha, distillate and butane streams. The latter are separated in this design mainly to provide a better basis for evaluating the value of the product.
- A small once-through, F-T plant might be ideal at a mine mouth location for processing coal-bed methane. The co-produced power could be used at the mine for processing the coal, and the hydrocarbon products could be shipped in the same manner as the coal, either by train or barge. Other examples of low-cost gas utilization can also be envisioned.

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Table 1

Overall Plant Major Input and Output Flows

Feed		
Natural Gas	100 MMSCF/day	(17,800 MMBtu/hr)
Raw Water Make-up	5.3 MMgal/day	
Primary Products		
F-T Naphtha	30.3 Mlbs/hr	(2,933 bbl/day)
F-T Distillate	64.6 Mlbs/hr	(5736 bbl/day)
Butanes	1.2 Mlbs/hr	(146 bbl/day)
Electric power	2018 MW-hr/day	(84.1 MW)

Table 2

Cost Breakdown of the Once-Through F-T Liquefaction Plant

Plant	Description	Cost (MM$)	% ISBL
101	Air Compression & Separation	70.4	32.7
102	Autothermal Reforming	22.8	10.6
103	CO_2 Removal and Recycle	13.4	6.2
201	Fischer-Tropsch Synthesis	35.8	16.6
202	Hydrogen Recovery	3.6	1.7
203	Product Fractionation	3.2	1.5
204	Wax Hydrocracking	11.8	5.5
31	Combined Cycle Plant	54.5	25.3
	Total ISBL	215.5	
	Offsite	120.3	
	Subtotal:	335.8	
	HO Service/Fees & Contingency	79.4	
	Total Cost:	415.2	

Figure 1
Once-Through Fischer-Tropsch (F-T) Design With Power Co-Production
(Overall Process Configuration)

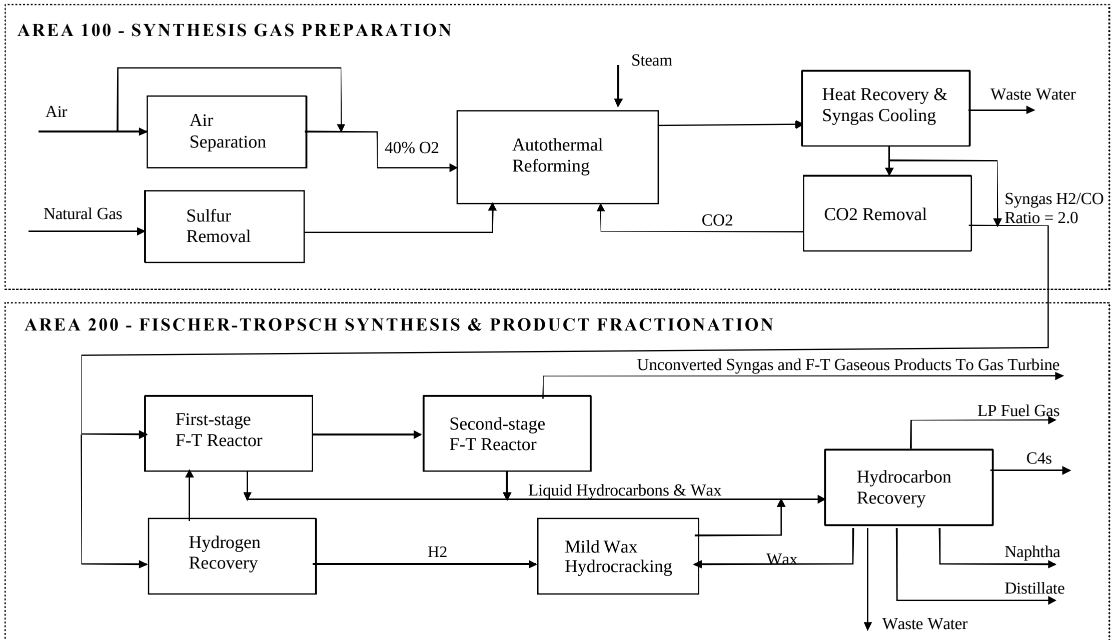

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Figure 2
Block Flow Diagram - Combined Cycle Plant

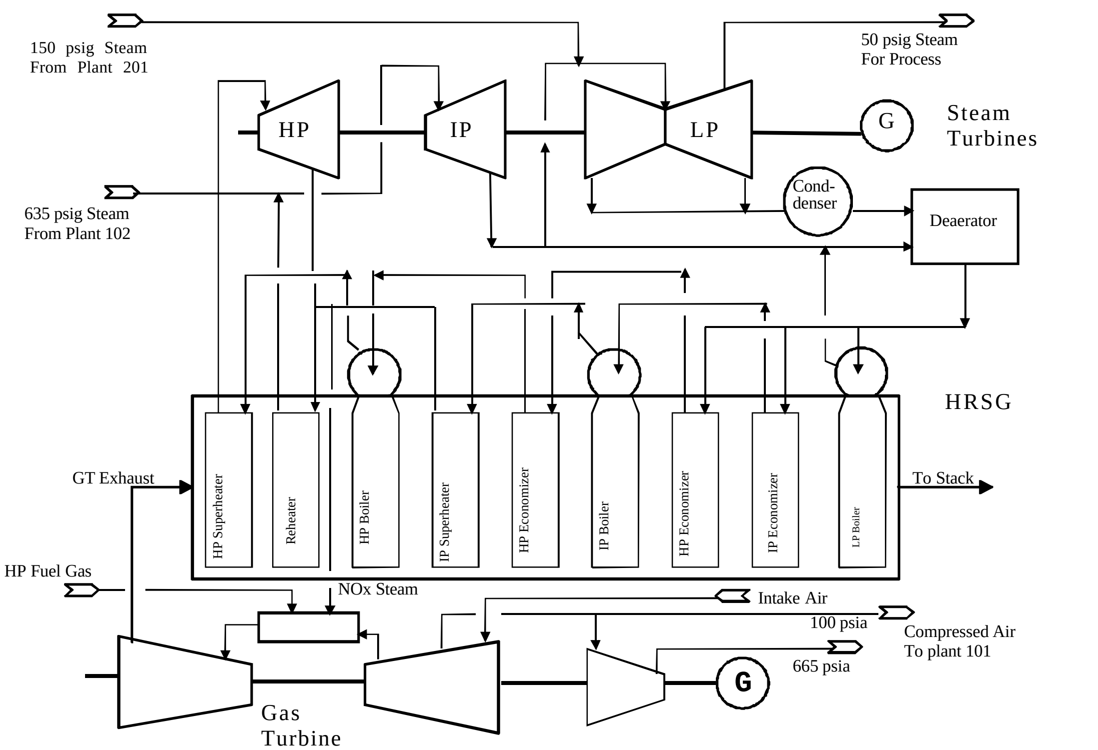

Figure 3
COE as a Function of Electricity Selling Price

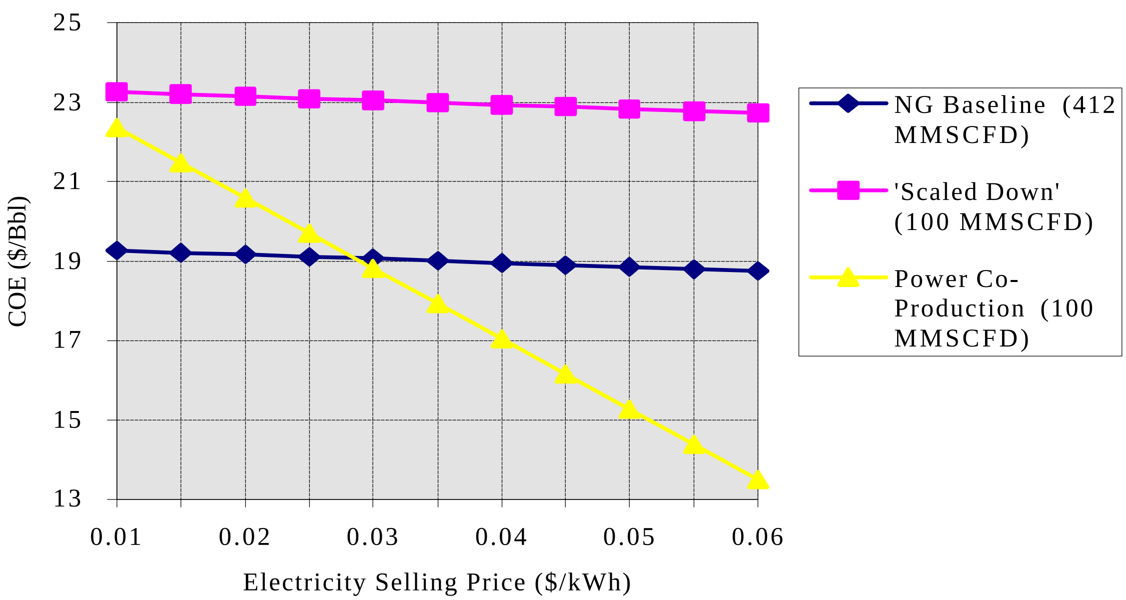

Figure 4
COE Cost Distribution With 0.5 $/MMBtu Gas @ 0.03 $/kWh Power Selling Price