OPPORTUNITIES FOR REDUCING PRODUCT COSTS IN INDIRECT LIQUEFACTION

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BACKGROUND:

MITRE has been developing integrated computerized simulation models of indirect coal liquefaction over the past several years(1). The objective of these models is to quantify potential reductions in product costs that could result from improvements in the technology brought about by continuing research and development. The models have shown that there are significant reductions in the cost of diesel and gasoline from advanced indirect liquefaction plants compared to the SASOL configuration. SASOL uses fixed-bed Lurgi coal gasifiers and Synthol circulating bed Fischer-Tropsch (F-T) reactors. This configuration produces excellent gasoline and diesel fuel, however, because of the large production of methane from both gasifier and F-T synthesis, the overall efficiency of the process to an all-liquid product is low and hence the relative cost of production is high. The advanced configuration, although at present only conceptual and not commercial as is SASOL, would use Shell entrained coal gasification and slurry-phase F-T synthesis. This combination produces very little methane, and allows the Shell synthesis gas to be directly utilized in the F-T process. Operating the F-T synthesis at high alpha to produce wax, further allows the selective production of diesel fuel.

The MITRE simulation model of this advanced indirect liquefaction process has estimated that cost reductions of about 27 percent compared to the SASOL configuration could be possible provided that the results obtained at bench and pilot plant scale can be duplicated at commercial size, and that certain operational problems can be adequately solved(2). The cost data for unit operations used in the MITRE model have been obtained from various sources and some of it is outdated. The United States Department of Energy has awarded a contract to Bechtel and Amoco to update this cost information and to develop a flexible process simulation model. This updated model could then be used to verify the cost estimates of the MITRE analyses and to investigate the impact of process performance on product cost. Preliminary results from the Bechtel/Amoco indirect baseline configuration show that the capital cost estimate for the advanced conceptual plant is almost identical to the MITRE estimate, and thus the resulting cost of gasoline and diesel fuel from this plant should be similar, depending on the economic parameters used.

OBJECTIVE:

The advanced indirect liquefaction plant mentioned above represents a significant improvement in the technology to produce clean transportation fuels from coal. However, the potential still exists to further reduce the costs of fuels from this process by continued research and development. The objective of this paper is to identify those areas in the indirect coal liquefaction baseline plant where further reductions in cost may be possible. In particular, this paper shows the results of a preliminary analysis to quantify possible reductions in product cost resulting from improvements in the Fischer-Tropsch area. This area includes the F-T reactors and associated equipment, and the unit operations in the F-T gas recycle loop. The F-T area of the plant is the area impacted by ongoing research and development activities in the DOE indirect liquefaction program. This analysis is preliminary in that the costs of unit operations from the new Bechtel/Amoco baseline design have not yet been incorporated into the MITRE model. Although this will not effect the trends shown in this analysis the absolute costs of liquid products that are determined may be different when the Bechtel/Amoco costs are incorporated into the MITRE model.

APPROACH:

The MITRE indirect liquefaction simulation model for the advanced configuration that includes Shell gasification and slurry-phase F-T synthesis was downsized to coincide with the Bechtel/Amoco conceptual plant with a nominal capacity of 50,000 barrels per stream day. Then the kinetic parameters used by Bechtel/Amoco in the slurry F-T model were substituted in the MITRE model. This resulted in the same per pass conversion and in the same number of reactors as estimated in the Bechtel basecase. The total capital cost for this plant was estimated to be $2982 million using the MITRE model. This agrees well with the preliminary Bechtel/Amoco capital cost of $2961 million for the same size plant(3). Once the MITRE simulation of the basecase plant was shown to be in agreement with the Bechtel/Amoco case, the analysis of further potential cost reductions beyond the basecase could be investigated.

This analysis only investigated the potential cost reductions that could result from improvements in the F-T area of the conceptual plant. This is the area that is impacted by the research and development underway in the indirect program. The cost impact of the following potential improvements were investigated using the MITRE simulation model:

- doubling the baseline catalyst activity
- doubling the catalyst loading
- doubling the superficial gas velocity.

TECHNICAL ACCOMPLISHMENTS AND RESULTS:

Table 1 shows the elements of cost that make up the total capital for the Bechtel/Amoco baseline plant. The field costs for the inside battery limits (ISBL), outside battery limits (OSBL), home office fee, and contingency shown in table 1 are preliminary costs from the Bechtel/Amoco baseline design. The coal feedrate for this plant is 18,592 tons per stream day of moisture free (MF) Illinois #6 coal. Coal receiving, storage, drying and grinding is $143.3 million, equivalent to 6 percent of the total field cost. Gasification includes nine Shell gasifiers (8 operating and 1 spare), gas treating and cooling, sour water stripping, acid gas removal, and sulfur polishing. This is 32 percent of the total field cost. Total clean synthesis gas production would include gasification, oxygen plant, coal handling, and by-product recovery, to give a total cost of $1307 million or 55 percent of the total field cost. If the proportion of OSBL related to synthesis gas production is included, then this cost would rise to about $1520 million or 64 percent of the total field cost. Clearly to obtain a significant reduction in capital cost for indirect liquefaction, reducing the cost of clean synthesis gas production will be necessary. If coal gasifier costs can be reduced as the technology for their production is improved, and membrane separation for oxygen production can be perfected, then the potential exists for these costs to be significantly reduced. For example, if gasification costs can be reduced by 25 percent, and membrane separation can reduce oxygen plant costs by 25 percent, then the resulting total plant capital could be reduced to $2620 million, resulting in a product cost reduction of about 10 percent.

The F-T and synthesis gas recycle loop costs amount to only about 20 percent of the total field cost. This area includes the unit operations shown in figure 1. It is this area that catalyst and F-T reactor research and development can influence. If improved catalysts and reactors result in increased space velocities and overall synthesis gas conversions, the resulting capital cost reduction will reduce the cost of the liquid products. However, because the F-T synthesis related area of the plant only constitutes about 20 percent of the plant capital, product cost reductions resulting from improvements in this area will be in the order of 5 to 10 percent. Although this may not appear to be a large reduction, nevertheless, a reduction in required selling price (RSP) of liquid transportation fuels from indirect liquefaction of 8 percent would be very significant. For example, if improvements in the F-T related area resulted in a 50 percent reduction in the capital related costs of this area, the plant capital cost would be reduced from $2961 million to $2677 million. This would result in a required selling price of about $41.70 per barrel, an 8 percent decrease from the baseline RSP. This baseline RSP of $45.52 per barrel for gasoline and diesel is calculated by MITRE using the consistent set of economic parameters that have been used in the past to estimate product costs. The RSP that Bechtel/Amoco calculates will be different from the MITRE RSP since they will be using a different methodology to estimate operating and maintenance costs. However, the intent of this paper is to show a differential cost relative to this baseline and not to predict the actual cost of liquids from this technology.

Figure 2 shows the resulting change in per pass synthesis gas conversion with increasing superficial gas velocity for several assumptions of relative catalyst activity and catalyst loading. The curves in figure 2 were obtained from the MITRE indirect liquefaction slurry F-T simulation model. The Bechtel design point shown on the figure corresponds to the 81.7 percent per pass conversion assumed in the Bechtel/Amoco baseline design. The difference in the MITRE activity estimate is the result of differences in the interpretation of the results from the Mobil slurry phase experiments(4). Doubling the Bechtel baseline activity would increase the per pass conversion to over 90 percent at the same space velocity. A doubling of catalyst loading to 45 weight percent would further increase conversion to over 95 percent.

Figure 3 shows the resulting RSP of gasoline and diesel from a conceptual plant that achieves the improved performance assumed in figure 2. The Bechtel baseline RSP of $45.50 per barrel, calculated from the MITRE methodology, is reduced to about $43 per barrel at the same space velocity if both the catalyst activity and the catalyst loading can be doubled. If the reactor can, in addition, be operated in a hydrodynamic regime allowing a superficial gas velocity of 20 cm per second, the RSP can be reduced to about $41.70, an 8 percent decrease.

Figure 4 shows the same curves as in the previous figure, but instead of the per pass synthesis gas conversion, the number of slurry F-T reactors are shown.

CONCLUSIONS AND FUTURE PLANS:

If continuing research and development can adequately solve the operational problems of integrating and running advanced gasifiers and slurry F-T reactors, and can improve catalyst activity and slurry F-T performance, then reductions in the RSP of gasoline and diesel of between 5 and 10 percent compared to the current conceptual baseline design should be possible. Further cost reductions must come from capital cost reductions in other plant areas, especially in gasification and oxygen production.

As a continuation of this project, MITRE will use the unit operations costs developed by Bechtel/Amoco in their baseline design and reanalyze these sensitivities to catalyst activity and reactor performance. This will allow the product RSPs to be determined using the same cost basis as the Bechtel design, and will allow the potential future cost reductions as a result of performance improvements to be substantiated.

ACKNOWLEDGMENT:

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REFERENCES:

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2) Gray, David, Glen Tomlinson, and Abdel ElSawy, Quantification of Progress in Indirect Coal Liquefaction. MITRE Working Paper WP-92W000045, February 1992.
3) Personal Communication from Sam Tam, Bechtel, August 24, 1993.
4) Kuo, J.C.W., Slurry Fischer-Tropsch/Mobil Two Stage Process of Converting Syngas to High Octane Gasoline. DOE/PC/30022-10, June 1983.

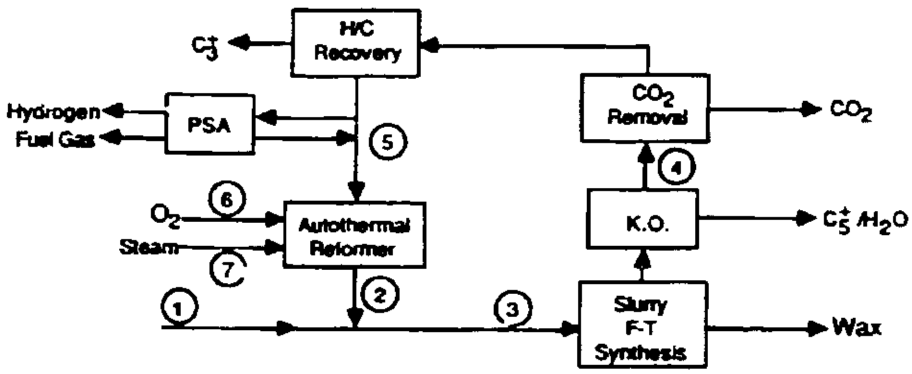
Table 1. Elements of Cost—Bechtel Design Basis

	MM$	%
Coal Handling	143.3	6
Gasification	770.3	32
Oxygen Plant	326.8	14
Byproduct Recovery	67.1	3
Fischer-Tropsch Synthesis	173.7	7
Syngas Recycle Loop	279.2	12
Product Refining	152.2	6
ISBL Field Cost	1912.5	
OSBL Field Cost	483.9	20
Total Field Cost	2396.4	100

Total field cost	2396.4
Home office and fee	178.5
Contingency	386.2
Total capital cost	2961.2

	$/Bbl	%
Capital	29.80	65
Coal	9.70	21
Catalyst	1.02	2
Other O and M	5.00	12
	45.52	100

Plant size: 50,000 BPSD
18,592 TPD (MF) coal
Illinois #6



				MMPH			
Stream #	1	2	3	4	5	6	7
Description	FF	Reformer Output	Input to F-T	F-T Output	Reformer Input	O_2	Steam
CH_4	0.3	0.6	0.9	2.4	2.2		
H_2O	1.0	2.5	3.5	0.1	0.1		2.83
H_2	72.4	17.1	89.5	17.8	10.9		
CO	125.2	8.4	133.6	6.3	5.0		
CO_2	0.3	0.7	1.0	62.8	0.5		
N_2	1.4	5.4	6.8	6.9	5.4		
O_2						1.74	
C_2				0.8	1.0		
C_3/C_4				1.3			
Totals	200.6	34.7	235.3	98.4	25.1	1.74	2.83

Figure 1. F-T Recycle Loop for Baseline Plant

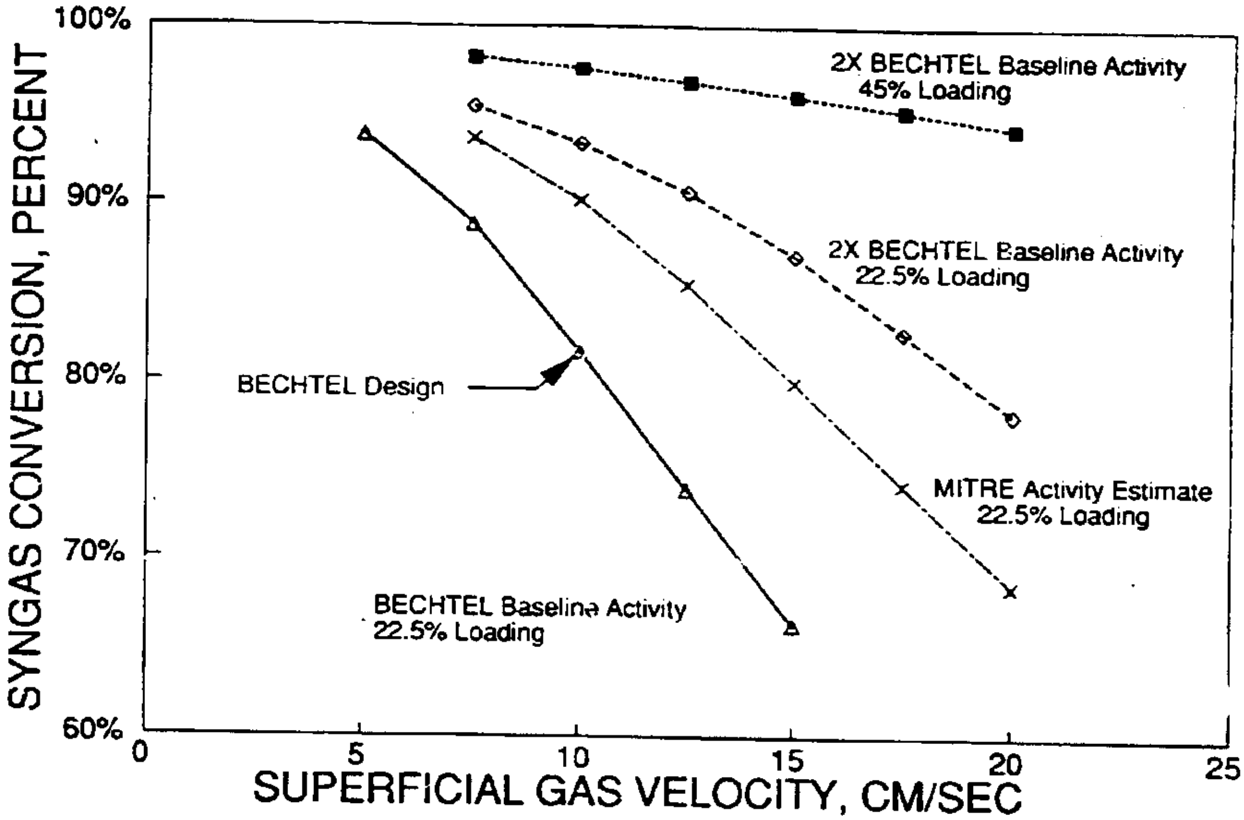


Figure 2. Syngas Conversion vs Superficial Gas Velocity for Four Catalyst Assumptions

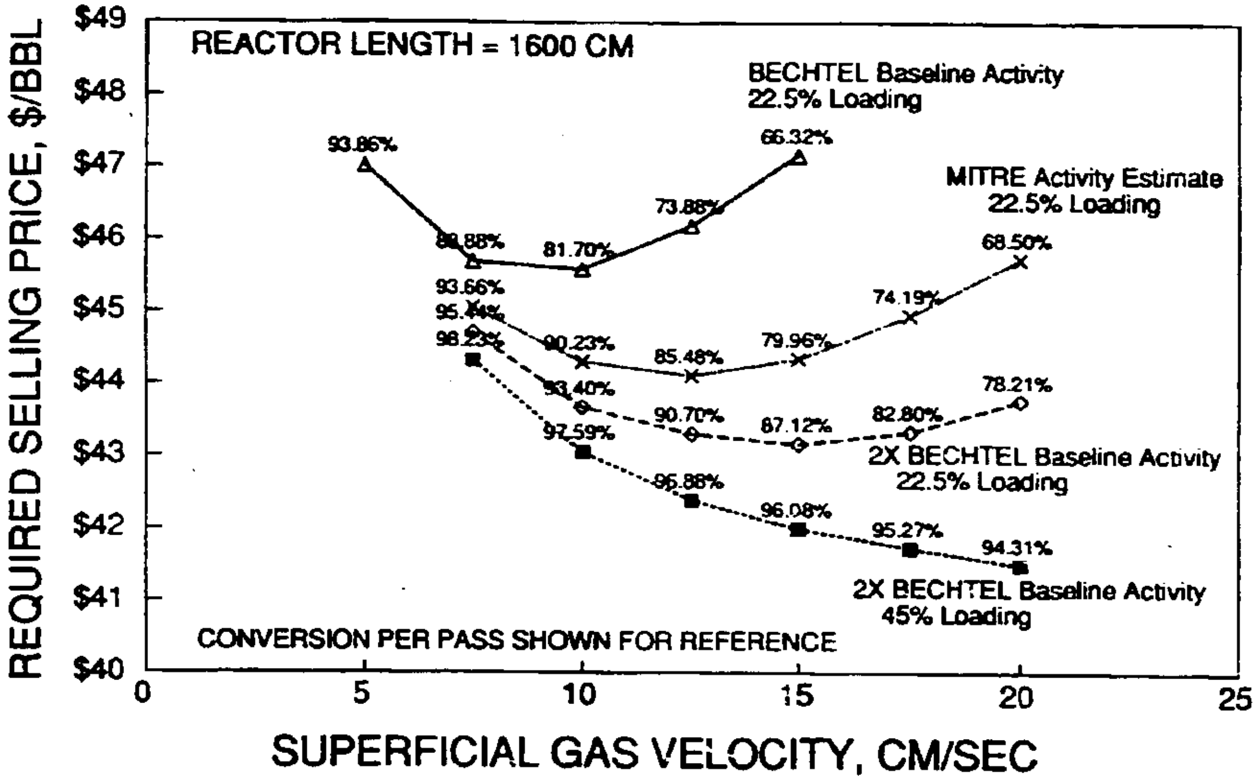


Figure 3. F-T Product RSP vs Superficial Gas Velocity for Four Catalyst Assumptions

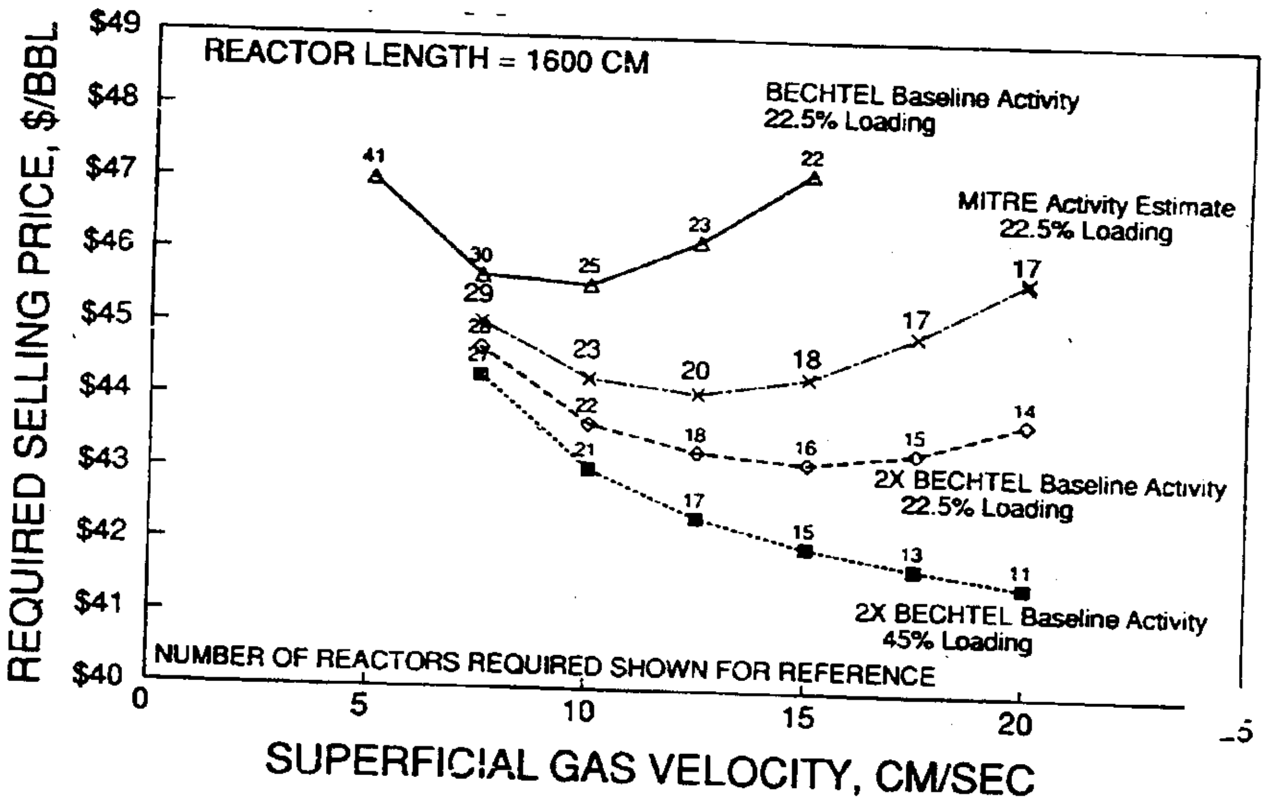


Figure 4. Product RSP vs Superficial Gas Velocity for Four Catalyst Assumptions