of the Wentworth Brothers' process (using available published information) with other processes did not show any inherent economic advantage for the WBI process.[13]

A. ICI Low-Pressure Methanol Synthesis

Status: The ICI low pressure (50-100 atm) methanol synthesis process is commercially proven worldwide.[6,13,22,23,24,33]

Process: Feed gas consisting of an approximately 2 to 1 ratio of hydrogen to carbon monoxide is fed into the synthesis loop. The methanol conversion is highest when the hydrogen to carbon monoxide ratio is 2 to 1 and the carbon monoxide to carbon dioxide ratio is as high as possible.

The first part of the synthesis consists of desulfurizing the feed gas when necessary to prevent the highly sensitive copper-based catalyst from being poisoned. This is accomplished by passing the gas through sulfur guard beds, which are typically made of zinc oxide (or, less commonly, activated carbon) to achieve sulfur levels below 1 ppmv.

The feed gas is then compressed to the recycle loop pressure, mixed with the recycle gas and then compressed to reactor pressure as it enters the methanol converter. The converter is a pressure vessel containing a bed of catalyst.

The temperature of the bed is controlled by the extent of the exothermic reaction and the queching of the reaction by cold feed gas. The pressure range is 50 to 100 atm while the temperature must be kept below 300°C (210 to 300°C) since the catalyst becomes deactivated at higher temperatures. The exit gas is passed through heat recovery units for initial cooling and then sent to the methanol separation unit where a crude methanol product is produced (95 percent methanol by weight). Conversion of CO to methanol per pass is about 5 percent.[13]

Catalyst life at pressures of 50 to 60 atm is 4 years while maximum catalyst life at 100 atm is 2.5 years (average is 1-2 years).

Advantages/Disadvantages: Compared with high-pressure processes, the ICI process is more adaptable to both large and small plants (55 to 2750 TPD) whereas high pressure processes are limited to large plants (1,000 to 2,500 TPD). Compression costs are lower because of reduced pressure.

The disadvantages are that the high-pressure processes allow a higher throughput of gas for the same size reactor and that the catalyst cost for low pressure technology is five times as expensive (50 cents/ton vs. 10 cents/ton).[13]

B. Lurgi Low-Pressure Methanol Synthesis

Status: The Lurgi low-pressure synthesis process (30-50 atm) is commercially proven.[13,25,23,22,33]

Process: The Lurgi methanol synthesis process uses a shell-and-tube reactor. The copper-zinc catalyst is packed in vertical tubes contained within a reactor shell which is filled with boiling water. The exothermic heat of reaction is removed by the generation of steam, thereby controlling the temperature of the reactor.

The hydrogen to carbon monoxide ratio of the feed gas is normally between 2 and 3, whereas the ratio of (hydrogen minus carbon dioxide) to (carbon monoxide plus carbon dioxide) is held around 2.2. After desulfurization the feed gas is compressed, combined with recycle gas and preheated before being fed into the reactor at one specific location. The Lurgi reactor has an operating range of 30 to 100 atm and 200 to 300°C but is typically operated at 70 atm and 260 to 270°C. The exit gas contains about 4-6 percent methanol and is sent to condensors to recover the crude methanol product which is generally sent on for purification.[13]

Advantages/Disadvantages: Like other low-pressure processes the Lurgi process has an economic advantage over high-pressure processes due to decreased compression costs at lower pressure. The reactor design also permits direct recovery of the exothermic heat of reaction by steam generation rather than a partial quench of the reaction to control heat build up.[13]

According to Lurgi, a natural gas to methanol plant using their synthesis technology is more efficient[22] and consumes 3-5 percent less natural gas per ton of pure methanol than competing technology. They estimate the annual savings for a 2,500 stpd plant is $4.2 - 5.0 million (U.S. dollars).[25]

When Badger compared the low pressure (1500 and 1400 psig) methanol synthesis processes employing quench type converters and licensed by Imperial Chemical Industries and Mitsubishi Gas Chemicals Corporation with that of Lurgi's tubular type low pressure (750 and 1200 psig) process they selected Lurgi's process for two reasons:

- Lower investment and operating costs for syn-gas compression
- Maximizes medium pressure steam production; thus minimizing overall utility costs and pay out time.

C. Haldor Topsoe Methanol Synthesis

Status: The Haldor Topsoe Methanol Synthesis process is commercially proven.[13,22,33]

Process: The Haldor Topsoe process is similar to other low- and intermediate-pressure methanol synthesis processes. The synthesis utilizes a copper-zinc-chromium catalyst in two or three radial flow converters operated in series. After being desulfurized (20 ppbv sulfur level) with zinc oxide guard beds, the feed gas is mixed with recycle gas and passed through the reactors flowing radially outward through each catalyst bed. Operating pressure and temperature ranges are respectively 48 to 144 atm and 220 to 350°C. The exothermic heat from each reactor is recovered by heating boiler feedwater with the hot exit gases. The gases are then condensed and sent to a separator where crude methanol is separated from uncondensed gases and later sent to product upgrading.[13]

Advantages/Disadvantages: The Haldor Topsoe process can operate at intermediate pressures for higher throughputs. It can also operate at higher temperatures which increases the activity of the catalyst provided it can retain its active sites and structural integrity.

D. Mitsubishi Gas Chemical Methanol Synthesis

Status: The Mitsubishi Gas Chemical (MGC) methanol synthesis process is commercially proven.[13,33]

Process: The MGC process appears to be very similar to ICI's intermediate-pressure process since both designs use a quench converter with a ternary copper-based catalyst operated at low temperature and intermediate pressure (240-310°C and 50-130 atm). The feed gas is split into a feed stream which is heated and fed into the converter, and a quench stream which is injected at several bed levels to control the buildup of the exothermic heat of reaction. After being used to preheat the feed gas the exit gas is condensed and sent on to distillation for a product purity in excess of 99 percent methanol by weight. Part of the recycle gas is used for fuel. The catalyst has an expected life of just over 1 year since it is very sensitive to sulfur.

Advantages/Disadvantages: The MGC intermediate pressure process has the advantage of accomodating moderately higher throughputs than lower pressure processes while keeping compression costs down as compared to high pressure processes.

The ICI catalyst appears to have a longer catalyst life (2 years for ICI vs. greater than 1 year for MGC). The MGC process typically uses a higher hydrogen to carbon oxides ratio in the feed gas than other processes (3.1 compared with 2.2 for other processes) but this is because it has only been used when natural gas is the feedstock.[13]

E. Vulcan-Cincinnati High Pressure Methanol Synthesis

Status: The Vulcan-Cincinnati high-pressure process is commercially proven.[13,33,22] However, the company had to stop operation in 1973 when the Middle East war forced the cancellation of a very large methanol plant in Saudi Arabi in which Vulcan had heavily invested (see Wentworth Brothers' process below).

Process: The feed gas ratio for $H_2/(CO + 1.5\ CO_2)$ should be adjusted to a value of 2 for optimum conversion after desulfurization. The feed gas is then compressed and fed to the converter which is usually operated in the range of 340 to 400°C and 200 - 300 atm. The convertor operates adiabatically with considerable temperature rise due to the exothermic heat of reaction, which is controlled by quenching the reaction with cold feed gas at several levels. After conversion the crude methanol product is condensed for removal yielding a product containing up to 97 wt percent methanol (depending on feed gas composition). There is also an option of producing up to 20 wt. percent of higher alcohols by changing operating conditions which would be helpful if used for blending with gasoline.

The catalyst, which is poisoned by H_2S levels greater than 3 to 5 ppm, has a typical life of about 4 years and can be regenerated. Conversion of CO to methanol per pass is approximately 5 percent.

Advantages/Disadvantages: The high pressure process is well suited for large methanol synthesis trains due to the high throughputs occuring. Catalyst costs are also less than the low-pressure copper-based catalysts, are not as sensitive to sulfur and can be regenerated. The process can produce a wider range of fuel products (3 - 20 percent higher alcohols).

Some disadvantages are: 1) that the high-pressure process has greater compression costs, 2) that the catalyst requires higher temperatures and pressures because it is not as active as the copper-based catalysts, and 3) that it may not be suited for small methanol plants.[13]

F. Wentworth Brothers Methyl Fuel Process

Status: The term methyl fuel, copyrighted by Vulcan-Cincinnati, represents the product of a methanol synthesis process which is focused on producing methanol for fuel rather than chemical uses. After Vulcan-Cincinnati stopped operation in 1973, the Wentworth Brothers and other engineers from Vulcan formed a new corporation in May 1975. Based on Vulcan experience and technology and relying on catalyst improvements and a reactor design adapted from proven petroleum technology, Wentworth Brothers, Inc. (WBI), is now marketing what they believe to be a much improved process for the production of large quantities of fuel-grade methanol. No commercial plants are in operation, but short-term tests of the catalyst at 300 TPD in a commercial methanol train have reportedly verified the basic operating parameters for the WBI methyl fuel process.[13]

Process: Details of the process operation, catalyst formulation, and reactor configuration are considered proprietary and are not available.

Advantages/Disadvantages: What is known is that the new catalyst is reportedly more active and durable than conventional low pressure catalysts although at the expense of the selectivity for methanol. The more active catalyst allows operation of the process at increased space velocities (throughput per reactor volume), and at higher temperatures and pressures which maximize fuel production per reactor train. The range of operating conditions for the methyl fuel process includes pressures up to 4,000 psi (270 atm) and temperatures from 200 to 400°C. The catalyst is claimed to be effective at CO_2 concentrations ranging from 20 percent to essentially zero versus conventional copper-based catalysts which require some CO_2. Oak Ridge National Laboratory states that from available published information it is not possible to ascertain whether the WBI Methyl Fuel Process has any inherent economic advantage over conventional copper-based methanol synthesis processes.[13]

G. Chem Systems Synthesis

Status: As of August 1980, the Chem Systems process was ready to move to the pilot plant stage.[21]

Process: The major difference between the Chem Systems synthesis and the other synthesis processes is that an inert hydrocarbon liquid is used as the medium for the catalyst instead of a gaseous phase. This liquid phase allows high conversions of carbon monoxide and hydrogen to methanol in addition to maximum recovery of reaction heat.[21]

In the process synthesis gas containing carbon monoxide, carbon dioxide and hydrogen is passed upward into the reactor concurrent with the inert hydrocarbon liquid, which is recovered in the separation plant and recycled back to reactor with the unconverted synthesis gas.[1,21] The fuel grade methanol product is 95-96 percent methanol by wt.[13]

Advantages/Disadvantages: Chem Systems claims that their conversion to methanol per pass is about 4 times as great as other processes (20 percent vs. 5 percent). However, Parsons believes that this process only has slight cost advantages over the existing processes.[1][13]

Ralph M. Parsons states that a new catalyst formulation with superior mechanical strength still needs to be developed to make the liquid phase methanol synthesis viable.[1] Breakdown of catalyst, inhibition of catalyst by fluid and insufficient solubility of the synthesis gas in the fluid are other possible problem areas with this design.[22]

H. Mobil Methanol-To-Gasoline (MTG) Process

Status: Mobil has conducted developmental studies of this process in fixed- and fluid-bed bench-scale units with two reactors being used in the fixed bed unit. The fixed bed reactor achieved over 200 days of successful operation. The single reactor fluid-bed unit has undergone two months of testing. Since the fluid bed reactor had a number of advantages over the fixed bed reactor, a 4-BPD fluid-bed pilot plant was designed, built and operated under a follow-on DOE contract in 1976-78. Startup and operation of this fluid-bed unit were reported to be very successful. Plans are currently under way for a 100 BPD fluid bed pilot plant sponsored by DOE, the Federal Republic of Germany, German industrial participants, and Mobil.[13,18] Of interest also is the reported news that, since November 1979, the government of New Zealand has been pursuing the Mobil methanol-to-gasoline process, with negotiations proceeding for a 13,000 BPD fixed-bed unit for installation almost immediately.[18,19]

Process: The conversion of methanol to hydrocarbons and water is a very exothermic reaction giving off 740 Btu/lb of methanol. Heat removal is therefore the principal problem in designing a reactor system.[7] For the fixed bed reactor the problem is minimized by dividing the reaction into two steps and using two reactors in series. In the first reactor crude methanol is partially dehydrated to an equilibrium mixture of methanol, dimethyl ether and water over a dehydration catalyst[14,19] releasing about 20 percent of the reaction heat.[7] In the second reactor the new shape-selective zeolite catalyst is used to convert both methanol and dimethyl ether to a liquid hydrocarbon product. This hydrocarbon liquid product is then sent to a fractionation unit where a deethanizer sends the ethane rich overhead product to the SNG train and the bottoms are sent to a stabilizer. The overhead product of the stabilizer is composed of isobutane and butene/propene which are sent to an alkylator to produce more gasoline and commercial grade propane and butane. The bottoms product of the stabilizer is a stabilized gasoline which is mixed with the gasoline product from the alkylation unit and sent to the gasoline blending unit to yield a high octane (93 research octane) gasoline.[7]

The inlet temperature of the second reactor is about 625°F. The adiabatic fixed bed process operates at essentially 100 percent conversion of methanol to hydrocarbons and water until the catalyst deactivates by carbon formation to an activity level where only partial conversion of methanol is achieved.[19] The zeolite catalyst must be regenerated once every 14 days.[7]

In the fluid bed process one reactor is used, operating at 750°F and 40 psig. The hydrocarbon product is generally treated in the same manner as the fixed bed product with the exception of a few changes. A deethanizer absorber is used in place of the high pressure deethanizer tower to provide a recycle stream to the reaction for increased propane-plus yield. A rich oil tower is also required.[7]

Advantages/Disadvantages: For the fluid-bed reactor the methanol conversion is greater than 95 percent, producing about 44 percent hydrocarbons and 56 percent water. The pentane-plus gasoline fraction of the hydrocarbons is about 60 percent. The propene, butene, and isobutane produced are approximately the right proportions for alkylation, bringing the total yield of 9 lb Reid Vapor Pressure gasoline (96 unleaded RON) up to 88 percent of the total hydrocarbon yield. The thermal efficiency for the methanol conversion is quoted at 95 percent.[13,27]

One potential problem with the gasoline produced from both of these processes is the presence of durene which boils in the gasoline range but has a freezing point of 175°F. This could cause engine problems since the durene could crystallize out in an engine's carburetor. Durene is present in conventional gasoline in very small amounts but could be present in relatively large amounts (3-6 percent) in Mobil MTG-gasoline. Durene levels of 5 percent in gasoline did cause some unsatisfactory engine operations during tests but at 4 percent levels effects were minimal. Since durene levels can be maintained to acceptable levels by proper process controls and it could always be mixed with conventional gasoline[28] the presence of durene may not pose too much of a problem.

I. Fischer-Tropsch

Status: The Fischer-Tropsch process is proven technology which has been producing liquid hydrocarbons at SASOL in Sasolberg, South Africa since 1955.[7] This is the only commercial scale coal liquefaction plant operating in the world today. The current SASOL plant uses two reactor schemes, a fixed-bed and a fluid bed. The current SASOL expansion to 50,000 BPD is based on a fluid-bed design.

Process: With the fluid bed reactor, purified synthesis gas is compressed and charged into the reactor. After mixing with the circulating hot iron catalyst, the reaction takes place as the mixture flows up the reactor through tube bundles in which oil is pumped for heat removal. At the top of the reactor, the mixture enters a large vessel in which cyclones separate the iron and vapor. The hot oil is circulated to a steam generator where 200 psig steam is produced. The overhead vapor is condensed and the vapor split into a recycle and purge stream, the latter being sent to hydrocarbon recovery. The condensed liquid is split into a cold recycle liquid and a light oil product. Based on a 2.13 molar feed ratio of H_2:CO to the F-T synthesis units the final product yield consists of 24 percent SNG, 54 percent 10 RVP gasoline, 9.8 percent diesel fuel, 5.5 percent alcohols, 3.8 percent LPG and 2.8 percent heavy fuel oil (percentages on a HHV BTU basis).[7]

The commercial catalysts include cobalt, for the fixed-bed reactor, and iron for both the fixed- and fluid-bed reactors. Operating conditions range from 200-325°C and from atmospheric pressure to 25 atm depending on the desired products. Because all of the reactions in the F-T process are exothermic, heat removal is important. For the fixed bed reactor, designs include a heat exchanger being cooled by boiling water or circulating oil. Fluid bed reactors use internal tube bundles for reaction heat removal.[7]

Advantages/Disadvantages: According to a study done by Mobil[7] which compares the F-T process to the Mobil MTG process, the MTG process has a number of advantages over F-T. For example, the Mobil technology gives a higher liquid product/SNG ratio (energy basis), 47/53 vs. 35/65. The F-T route has 18 processing steps compared with nine for MTG. The MTG gasoline had a higher RON (93 vs. 91) and a lower olefin content (11 vs. 20 percent). The economics show that the MTG gasoline cost is moderately cheaper than F-T gasoline ($.60 - $1.00 per gal. vs. $.70 - $1.35 per gal.). The overall efficiency for MTG was 62 perscent vs. 58 percent for F-T.[7]

A recent article in Hydrocarbon Processing advocating F-T stated that the 40-45 percent efficiency reported for the operating F-T process is really depressed because the SNG produced is reformed to make additional H_2 and CO instead of being sold as a product (SNG is not marketable in South Africa). If the process was brought to the U.S. the author believes that SNG would be a viable product as a pipeline quality methane gas, thus increasing the F-T efficiency to 60 percent (close to 58 percent which was previously stated by Mobil). In addition the article points out that coal consumption for the F-T process is high only because they use a low quality coal with 30 percent ash.[26]

VII. Comparison of Indirect Liquefaction Design Studies

As discussed earlier, there have been a number of studies evaluating the indirect liquefaction of coal. This chapter is divided into four main sections, the production of methanol from: 1) bituminous coals, 2), subbituminous coals, and 3) lignite and 4) the production of gasoline from methanol using the Mobil MTG technology; however, this fourth section also includes the production of gasoline and other products via Fischer-Tropsch technology.

Each section begins with a brief introduction and a comparison of three aspects of the various studies: level of engineering design, feedstock analysis, and material balances and efficiencies. Then capital, operating and product costs for each study will be presented on a consistent economic basis and then compared to reconcile as many of the differences as possible.

A. Methanol from Bituminous Coal

There were five original studies available which investigated the technical feasibility of producing methanol from bituminous coals:

1. R.M. Parsons, Co., for EPRI, "Screening Evaluations: Synthetic Liquid Fuel Manufacture,"[1]

2. C.F. Braun for EPRI, "Coal to Methanol Via New Processes Under Development: An Engineering and Economic Evaluation,"[2]

3. Dupont Co. for U.S. ERDA, "Economic Feasibility Study, Fuel Grade Methanol from Coal for Office of Commercialization of ERDA,"[3]

4. Badger Plants, Inc., "Conceptual Design of a Coal-to-Methanol Commercial Plant,"[4] and,

5. Exxon Research and Engineering Co., "Production Economics for Hydrogen, Ammonia, and Methanol During the 1980-2000 Period."[5]

The studies differ in depth of design and use different assumptions with respect to key economic parameters; also plant sizes vary widely. Since methanol-from-coal technology is well developed and much of it is common to all of the studies, the large cost differences between the studies should be reconcilable by placing them on a consistent economic basis.

Depth of Design: The level of engineering detail of the five studies varies. The most detailed studies where found to be those performed by Badger and EPRI/C. F. Braun. These studies include complete material balance information for a detailed flowsheet and extensive design details, including drawings. The studies with the next highest level of engineering design are the EPRI/Ralph M. Parsons and Du Pont studies. These studies are screening evaluations and their level of detail is not sufficient to allow comparison with more detailed studies. The EPRI/Ralph M. Parsons study includes evaluations of four gasification processes in combination with Chem Systems Methanol synthesis; these four processes are denoted as Cases 1, 2, 3 and 4 of the Parsons study. The study found to be based on the least detail is the Exxon/Chem Systems study which consists of summary economic information for two gasification processes, but provides no details for material and energy balances. These two processes are denoted as Cases 1 and 2 of the Exxon study.

Ultimate Analysis for Bituminous Coals: Ultimate analyses of the bituminous coals used in the various studies are presented in Table 8. Also listed in this table are the heating values of the coals. The Exxon/Chem Systems study did not report an analysis for their coal; however, they used an Illinois high sulfur coal which is probably similar to the Illinois No. 6 bituminous coals reported in the two EPRI studies. The Du Pont coal is also similar to the coals presented in the EPRI studies. However, the coal considered in the Badger report is a low sulfur coal which would meet the sulfur dioxide standard for large power plants.[12] It is unlikely that a coal of this quality would be used to produce methanol.

Material Balance and Efficiencies: Feedstock and product rates for each study are presented in Table 9. Methanol and coal are presented on both a short ton per calendar day (tpd) basis and on an energy basis. Other products include fuel gas in Cases 1 and 2 of the Parsons study; the Badger study includes chemical grade methanol along with the fuel grade. All rates are based on 50,000 FOEB/CD of total products.

Table 9 shows that the process efficiencies for the studies vary from 49.3 percent for the Koppers-Totzek/ICI case prepared by Exxon/Chem Systems to 58.2 percent for the Texaco/Chem System case prepared by EPRI/Parsons. An investigation of these efficiencies shows that the two Koppers-Totzek cases are amongst those with the lowest efficiencies. This is not surprising since the Koppers-Totzek gasifier is the only first generation gasifier listed in this table, and since the gasifier operates at near atmospheric pressure so that the product synthesis gas must be compressed, thus resulting in an efficiency penalty. When comparing the three Texaco gasifier studies it can be seen that there is a substantial difference (7.5 percent) between the highest and lowest efficiency. The C.F. Braun study reports an efficiency of 55.7 percent whereas the DuPont study reports an efficiency of 50.6 percent. A key variable in the Texaco process which affects the efficiency is the coal/water slurry concentration. The greater the coal concentration in the slurry, the higher the efficiency of the Texaco process.[9] The C.F. Braun case utilized a 59 percent coal concentration in the slurry whereas the DuPont study utilized a 54 percent coal concentration. This difference will account for part of the efficiency discrepancy. The Parsons/EPRI study did not report any processing information for the Texaco gasifier and therefore the high efficiency reported for this process could not be investigated. The remainder of the processes reported in Table 9 utilize other advanced technology gasifiers, and therefore are expected to have higher efficiencies than first generation gasifiers, e.g., Koppers-Totzek.

Table 8

Ultimate Analysis of Bituminous Coal Feedstocks

Study:	Parsons Cases 1,2,3,4[1]	C. F. Braun[2]	DuPont[3]	Badger[4]	Exxon[5] Case 1 and 2
Coal Type	Bituminous Ill. No. 6	Bituminous Ill. No. 6	Bituminous	S. Appal.	Ill. High Sulfur
HHV (Btu/lb.)	12,235 (wet)	12,150 (dry)	12,113 (dry)	12,840 (dry)	11,390 (wet)
LHV (Btu/lb.)	11,709	-	10,874	-	-
Ultimate Analysis, Wt % Dry Coal					
C	69.5	68.25	66.89	73.8	
H	5.3	5.00	4.47	4.8	
O	10.0	11.23	8.41	6.4	
N	1.3	0.81	1.28	1.6	
S	3.9	3.88	4.47	1.1	
Ash	10.0	10.83	14.48	12.3	
Total	100.0	100.0	100.0	100.0	
Wt% Moisture (As Recieved)	4.2	10	6.38	2.4	

Table 9

Methanol from Bituminous Coal: Feedstock and Product Rates (Normalized to 50,000 FOEB/CD Product)

	Parsons[1]							Exxon[5]	
Study:	Case 1	Case 2	Case 3	Case 4	C.F. Braun[2]	DuPont[3]	Badger[4]	Case 1	Case 2
Mass Basis									
Feedstocks									
Coal, tpd	21,722	21,150	23,006	20,714	21,795	25,685	20,048	24,447	26,188
Products									
Methanol, tpd	14,500	15,132	15,349	15,349	15,172	16,223	14,570	15,227	15,227
HHV Btu/lb	9,610	9,610	9,610	9,610	9,722	9,092	9,407	9,687	9,687
Fuel Gas, mscf/CD	20.39	4.89	-	-	-	-	-	-	-
Chemical Grade Methanol, tpd	-	-	-	-	-	-	1075		-
Energy Basis (HHV), mBtu/CD									
Feedstocks									
Coal	531,531	517,544	562,977	506,873	529,623	582,543	514,834	556,902	596,565
Electricity (energy equivalent)	-	-	-	-	-	-	-	1810	1812
Products									
Methanol (Fuel Grade)	278,682	290,845	295,000	295,000	295,000	295,000	274,121	295,000	295,000
Fuel Gas	16,317	4,155	-	-	-	-	-	-	-
Methanol (Chemical Grade)	-	-	-	-	-	-	20,879	-	-
Thermal Eff'cy, %	55.5	57	52.4*	58.2	55.7	50.64	57.3	52.8	49.3

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* 95% conversion assumed for gasifier as opposed to 100 percent for the other cases.

Economics: In this sub-section the capital and operating costs for each study will be presented in order to obtain the desired product costs. These costs have been placed on a consistent economic basis as reported in a previous report.[10a]

Table 10 presents all of the investment costs (normalized to 50,000 FOEB/CD) broken down into individual process unit costs. This table shows that the total instantaneous investments range from $1.62 billion for the Shell-Koppers/ICI case of the Exxon study to $2.56 billion for the Koppers-Totzek case of the Parsons study, which represents a $940 million instantaneous investment difference.

Operating costs are presented in Table 11. Some of the studies did not itemize the operating costs which makes it difficult to compare these costs between each study. The net annual operating costs range from about $315 - 480 million. Most of the operating cost estimates lie in the $340 - 440 million per year range.

Tables 12 and 13 present economic summaries of methanol costs for capital charge rates of 11.5 and 30 percent. For the lower capital charge rates product costs vary from $5.30 for the Parsons/BGC Lurgi study to $7.23/mBtu for the Parsons/Kopper-Totzek study. For the higher capital charge rate product costs vary from $8.74 to $12.42/mBtu.

Now some of the capital and operating cost differences will be reconciled. First, one would expect the Koppers-Totzek cases to have a high capital cost because it is a low pressure process; this results in higher compression and gasification costs than for high pressure gasification technologies. Also, since the Koppers-Totzek gasifier is a first generation gasifier, its capital cost is expected to be higher than the more advanced technology cases. The instantaneous plant investment for the Parsons/Koppers-Totzek study is $2.56 billion while that of the Exxon study is $1.9 billion, which represents about a $650 billion investment difference. The investment estimate of the Exxon study is similar to those of the advanced technology gasification cases which seems unreasonable. The only difference between the two Koppers-Totzek cases is the methanol synthesis technology used. Since there is very little difference expected in the capital costs of the Chem Systems and ICI synthesis units, it is surprising to see such a large capital cost difference between these two cases, and the difference may be attributed to lack of design detail for the Exxon study which was discussed earlier. Therefore, it is believed that the Parsons study is more representative for the Kopper-Totzek case.

Three studies investigated the use of Texaco gasification technology: Parsons, C.F. Braun, and DuPont. The respective plant investment costs for these studies are $2.05, $1.75 and

Table 10

Methanol From Bituminous Coals: Capital Costs Summary*
(Millions of First Quarter 1981 Dollars)

Study:	Parsons[1]				C.F. Braun[2]	DuPont[3]	Badger[4]	Exxon[5]	
	Case 1	Case 2	Case 3	Case 4				Case 1	Case 2
Technology: Gasification/ Synthesis	Foster Wheeler/ Chem. Systems	BGC Lurgi/ Chem. Systems	Koppers Totzek/ Chem. Systems	Texaco/ Chem. Systems	Texaco/ Chem. Systems	Texaxo/ ICI	Slag Bath/ Lurgi	Shell-Koppers/ ICI	Kopper Totzek/ ICI
Coal Preparation	83	40	87	73	110	112	37	-	-
Tar and Phenol Recovery	-	99	-	-	-	-	33	-	-
Gasification	186	145	566	385	167	205	187	-	-
Shift Conversion	42	56	51	51	97	75	66	-	-
Acid Gas Removal	281	241	266	283	187	175	193	-	-
Sulfur Removal	14	14	15	15	-	-	16	-	-
Synthesis Gas Compression	71	76	125	-	-	-	33	-	-
Methanol Synthesis	170	175	177	177	164	190	197	-	-
Cryogenic Recovery	29	30	-	-	-	-	16	-	-
Sulfur Recovery/ Methanol Drying	28	29	29	29	26	-	24	-	-
Oxygen Production	348	263	374	430	300	373	313	-	-
Steam and Power Generation	68	80	197	106	123	302	90	-	-
Subtotal	1320	1248	1887	1549	1174	1432	1205	775	951
General Facilities and Offsites	198	187	283	232	306	209	453	597	666
Contingency	228	215	325	217	222	246	249	206	243
Contractor's Fee	44	42	63	52	43	48	42	40	47
Total Instantaneous Plant Investment	1790	1692	2558	2050	1745	1935	1949	1618	1907

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* Investment costs are based on 50,000 FOEB/CD of product.

Table 11

Methanol From Bituminous Coals: Operating Cost Summary (Millions of First Quarter 1981 Dollars Per Year)

Study:	Parsons[1]				C.F. Braun[2]	DuPont[3]	Badger[4]	Exxon[5]	
	Case 1	Case 2	Case 3	Case 4				Case 1	Case 2
Technology: Gasification/ Synthesis	Foster Wheeler/ Chem. Systems	BGC Lurgi/ Chem. Systems	Koppers Totzek/ Chem. Systems	Texaco/ Chem. Systems	Texaco/ Chem. Systems	Texaxo/ ICI	Slag Bath/ Lurgi	Shell-Koppers/ ICI	Kopper Totzek/ ICI
Raw Materials:									
Ill. No. 6 Coal	218	212	231	208	219	258	199	245	263
Catalysts and Chemicals	10.6	10.6	12.1	10.6	20.7	15	3.4	4.1	4.1
Utilities:									
Power	-	-	-	-	-	-	23	6.8	6.8
Process Water	-	-	-	-	-	-	-	19.4	20.6
Stack Gas Clean-up	-	-	-	-	-	-	-	8.7	12.1
Labor and Related:									
Labor	-	-	-	-	36.3	12	29.9	1.3	1.3
Supervision	-	-	-	-	10.9	-	1.5	4.1	4.1
Plant Overhead	-	-	-	-	-	-	2.9	1.7	1.7
Capital Related:									
Maintenance	-	-	-	-	40.1	97	-	63.4	74.5
General Plant Overhead	-	-	-	-	23	-	-	41.2	48.4
Insurance and Property Tax	-	-	-	-	47.4	41	-	42.1	47.5
Interest on Working Capital	7.5	7.1	10.7	8.8	7.3	8.1	7.2	6.8	8.0
Other Operating Costs	124	120	181	149	1.2	17	52	-	-
Gross Annual Operating Cost	360	350	435	387	406	458	319	445	492
By-Product Credit	(14.7)	(13.9)	(12)	(13.1)	(13.2)	(19.5)	(3.8)	(15.0)	(16.0)
Net Annual Operating Cost	345	336	423	374	393	429	315	430	476

Table 12

Economic Summary of Methanol from Bituminous Coal, CCR = 11.5%
(Millions of First Quarter 1981 Dollars)

Study:	Parsons[1]				C.F. Braun[2]	DuPont[3]	Badger[4]	Exxon[5]	
	Case 1	Case 2	Case 3	Case 4				Case 1	Case 2
Total Instantaneous Investment	1790	1692	2558	2050	1745	1935	1949	1618	1907
Total Adjusted Capital Investment	2030	1919	2901	2325	1979	2194	2210	1835	2163
Start-up Costs	126	119	179	147	122	136	137	114	134
Pre-paid Royalties	8	9	13	10	10	10	10	10	10
Total Capital Investment	2164	2047	3093	2482	2111	2340	2357	1959	2207
Working Capital	125	118	179	147	122	135	137	113	133
Total Capital Requirement	2289	2165	3106	2629	2233	2475	2494	2072	2440
Annual Capital Charge	249	235	356	302	243	269	287	225	265
Annual Operating Cost	345	336	423	374	393	429	315	430	476
Total Annual Charge	594	571	779	676	636	698	602	655	741
Product Cost									
$/FOEB of Product	32.55	31.29	42.68	37.00	34.85	38.25	32.99	35.89	40.60
$/mBtu of Product	5.52	5.30	7.23	6.28	5.90	6.48	5.59	6.08	6.88

Table 13

Economic Summary of Methanol from Bituminous Coal, CCR = 30%
(Millions of First Quarter 1981 Dollars)

	Parsons[1]							Exxon[5]	
Study:	Case 1	Case 2	Case 3	Case 4	C.F. Braun[2]	DuPont[3]	Badger[4]	Case 1	Case 2
Total Instantaneous Investment	1790	1692	2558	2050	1745	1935	1949	1618	1907
Total Adjusted Capital Investment	1998	1888	2855	2287	1947	2159	2176	1806	2128
Start-up Costs	126	119	179	147	122	136	137	114	134
Pre-paid Royalties	8	9	13	10	10	10	10	10	10
Total Capital Investment	2132	2016	3047	2444	2079	2305	2323	1930	2272
Working Capital	125	118	179	147	122	135	135	113	133
Total Capital Requirement	2257	2134	3226	2591	2201	2440	2460	2043	2405
Annual Capital Charge	640	605	914	777	624	692	738	579	682
Annual Operating Cost	345	336	423	374	393	429	315	430	476
Total Annual Charge	985	941	1337	1151	1017	1121	1053	1009	1158
Product Cost									
$/FOEB of Product*	53.97	51.56	73.26	63.07	55.73	61.42	57.70	55.29	63.29
$/mBtu of Product	9.15	8.74	13.42	10.69	9.44	10.41	9.78	9.37	10.73

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* One FOEB = 5.9 mBtu

$1.94 billion. The total gasification costs for the C.F. Braun and DuPont studies are almost identical (about $565 million) whereas the total gasification cost from the Parsons study is about $240 million higher. The main difference in the C.F. Braun and DuPont studies lie in differences in the cost of the non-gasification equipment (oxygen production, offsites, etc.). It is difficult to determine which of the studies is most representative; based on level of design, the C.F. Braun study should be considered most representative. However, to be somewhat conservative in comparing methanol costs to direct liquefaction cost both the C.F. Braun and the DuPont studies will be chosen and a range of costs used. The product costs for these Texaco studies lie in about the same range as those for the other advanced technology cases, which seems reasonable.

In an effort to determine the representativeness of the Badger costs, DOE commissioned Oak Ridge National Laboratory in June, 1978 to make an independent assessment of the Badger report.[35] ORNL reported that Badger's design is based on equipment sizes well beyond the present state-of-the-art in order to take advantage of the projected economies of scale. Therefore, ORNL believes that the Badger design is more representative of an Nth plant design rather than a first plant design. For a first plant design the Badger capital and operating costs appeared to be unreasonable to ORNL. The operating cost for the Badger study listed in Table 11 are lower than any of the others listed in the table and does appear to be low. However, the Badger study is based on an advanced technology "slag bath" gasifier, and the capital cost based on this design is a bit higher than those of the other studies based on advanced technology gasifiers listed in Table 10. It is expected that these designs would have a lower capital cost than those based on technology commercially available or near commercially available. The degree to which these plants may represent Nth plant designs is uncertain.

Even though the Texaco gasifiation process is advanced, its costs will be presented separately from the other advanced gasification costs. Based on the above dicussion the DuPont and C.F. Braun studies will be used to represent the range of product costs for methanol from Texaco gasification technology, and the Parsons study for Kopper-Totzek technology. The remainder of the studies represent methanol costs for the other advanced gasification technologies. These costs are as follows:

	$/mBtu	
	Capital Charge Rate	
	11.5%	30%
Koppers-Totzek	7.23	12.42
Texaco	5.90-6.48	9.44-10.41
Advanced Technology	5.30-6.08	8.74-9.78