5.0 AREAS NEEDING FURTHER DEVELOPMENT

5.1 Backmixing Effects

One of the key issues left only partially defined in this study is the exact extent of backmixing effects on scale-up. The effect has been minimized by the choice of 80% rather than 90% as the design conversion per pass. There may be cases where higher conversions are desired and further study of backmixing effects is recommended.

Several more detailed slurry reactor models have been developed, and are discussed in Appendices A and B, which provide solutions to backmixing effects by incorporating axial dispersion coefficients. In order to use these models for scale-up, it is necessary to obtain axial dispersion data in a system which is physically and geometrically similar to the proposed design. This means that pilot plant data are required over a range of reactor diameters at superficial velocities and catalyst concentrations equal to those proposed for design. It is also important that cooling tubes be incorporated into the reactor design in the same fashion and with the same surface to volume ratio proposed for the commercial reactor.

It has not been possible to use published models directly for scale-up because of the way they handle gas holdup and other factors. All of the models proposed to date use an overly simplified expression in average gas velocity to estimate gas holdup. Most assume a constant contraction factor. All use a simplified expression for reaction rate which is first order in hydrogen concentration. These approaches may well be adequate for design purposes, but pilot plant confirmation is needed. In addition, none of the previous experimental work has been at the design superficial velocity and catalyst concentration proposed in this study.

The La Porte reactor offers the possibility of obtaining useful design information for model development if converted to Fischer-Tropsch operation. If backmixing effects are indeed significant, some consideration might be given to installing baffles or trays in the reactor to reduce backmixing. The presence of suspended catalyst is a potential problem, but if effective baffling can be provided in a fluidized-bed reactor (as in Mobil's MTG process) then its use in a slurry reactor may also be feasible.

5.2 Pressure Effect

As discussed in Section 2, Bechtel was unwilling to assume a linear pressure effect on the GHSV requirement for a given conversion level as predicted by the Fischer-Tropsch slurry reactor models. Gulf data on fixed-bed cobalt catalyst indicated that the "catalyst activity" is not linear with pressure but flattens off at pressures above 200 psia. For design purposes, a square root decrease in the rate constant with pressure level above 1100 kPa (160 psi) was assumed. Further data would be useful and some may be available in the literature (see Appendix B - part 3) but further measurements of the pressure effect at reactor design conditions are recommended.

It would be of interest in future studies to examine the effect of pressure on the reactor cost comparison. As described in Section 4, a compression step has been added to roughly double the pressure out of the Shell gasifier before F-T synthesis. It should be possible to gain a rough idea of the effect of pressure on cost by prorating from this study. The assumption of a square root effect of pressure on reactor size could then be compared with the linear assumption. This would set a reasonable goal for the proposed experimental studies.

5.3 Heat Removal

By increasing reactor pressure and catalyst concentration, heat removal requirements per unit reactor volume have been increased to the point where the reactor becomes quite packed with

cooling tubes. A double tube sheet design with bayonet tubes has been adopted for this study, but at some point it may be worth again considering an external cooling loop. External cooling loops have been provided in bubble columns in which rapid circulation is provided by the difference in density between the aerated reactor and the exchanger. No pump is required. As far as is known, such a design has yet to be applied when a slurry is present, but the concept still seems applicable.

5.4 Improved Catalyst Activity

Allowable space velocity in a methanol reactor is roughly four times that in a F-T reactor, indicating that there may be room for improvement in F-T catalyst activity. If activity is improved, the mass transfer resistance will become more limiting. Some guidance could be provided by estimating the capacity of the reactor if the mass transfer resistance were completely controlling. Under these conditions, heat removal would become a problem and an external circulation loop might be a necessity.

5.5 Use of Steam at Low H2/CO Ratio

The slurry F-T reactor used in this study operates below 0.67 H_2/CO inlet ratio so that the inlet ratio is less than the expected consumption ratio. This has been compensated for by steam addition. This concept appears reasonable but it would be useful to have actual data under these conditions. If it is not feasible, the solution is an extra water gas shift reaction step, ahead of F-T synthesis, as provided by MITRE.

5.6 Catalyst Activity Maintenance

A primary consideration in choosing a slurry reactor is the expected life of the catalyst. If only a few months life is expected, there is considerable incentive to go to a system which can handle continuous catalyst replacement. This is primarily an operating problem and the relative economics can be defined by a sensitivity analysis.

5.7 Mixed Alcohols

Design data on the Octamix process in a slurry reactor are lacking. For one thing, the proper slurry liquid for mixed alcohol synthesis must be determined. Higher oxygenates will undoubtedly show some solubility in the hydrocarbon liquid used for the slurry methanol process. If the higher oxygenates form a stable liquid phase, then a portion of the product could be used for slurry liquid as in the Fischer-Tropsch design. Similar facilities would be required to recover product from catalyst.

The assumption that GHSV requirement is the same as the fixed-bed may be conservative. Since the equilibrium limitation is not as severe as when methanol alone is being produced, it may be possible to take advantage of a somewhat higher average temperature in the slurry reactor to reduce the GHSV requirement. In this case the height shown for the slurry reactor can be reduced. *Unless the design pressure can be reduced, however, further test work is not recommended.*

5.8 Fixed-Bed Modelling

The gas phase fixed-bed reactor can be accurately modelled using stepwise integration procedures and providing an indication of temperature profiles. The difficult part will be to simulate accurately the two-phase behavior in the portion of the reactor where condensation is occurring. This is known to occur in F-T synthesis and, apparently, can also occur in high conversion methanol synthesis with a stoichiometric feed gas, enhancing the conversion. These phenomena may require experimental verification before an acceptable model can be developed.

6.0 CAPITAL AND OPERATING COST COMPARISONS

Capital and operating costs differentials have been determined for the slurry and fixed-bed reactor Fischer-Tropsch processing systems described in Section 4.3. The cost estimates are for those specific units which are affected by the choice of reactors. Backup in terms of process flow diagrams, equipment lists, material balances, overall steam and water balance diagrams and utility summaries are given in Appendix F. Capital costs have also been compared for the slurry and fixed-bed methanol synthesis sections described in Section 4.1. Backup is provided in Appendix E . Appendix E also contains Lurgi material on their Octamix™ process.

For those plant sections where detailed information is provided, costs were estimated for each item of major equipment and an overall direct cost was built up by using Bechtel historical factors for installation labor, bulks and subcontracts. Cost of pertinent Fischer-Tropsch upgrading units and utility plants were read off of cost-capacity curves. To these direct costs were added the contractor's indirect costs (distributable field costs which are not identified with any particular process or utility unit) to give the total field cost. An allowance of 25% was then made for contractor's home office engineering, fee and contingency to give the total plant investment. Owner's costs, working capital, startup costs and initial catalyst and chemicals are not included. Import duties on equipment which might be purchased overseas are also not included. Costs are for mid-1990 and represent a typical U. S. Gulf Coast location, with labor at $16/manhour. These estimates should be accurate to within ± 25%.

The alternative cases described in Subsections 4.5, 4.6 and 4.7 have not been costed, nor have the mixed alcohol designs covered in Section 4.2. A preliminary estimate is provided of the cost savings effected by doubling superficial velocity and slurry concentration for the slurry reactor over what has been demonstrated experimentally for the Fischer-Tropsch process.

Because the Fischer-Tropsch cases are of the greatest interest, these will be discussed first.

6.1 Fischer-Tropsch Comparison

As described in Section 4.3, there are key processing differences between the slurry reactor case and the fixed-bed reactor case. The slurry reactors are operated on low H_2/CO ratio gas as produced in the Shell gasifiers without shifting whereas the fixed-bed reactors are operated on 2 to 1 ratio gas which requires shifting and CO_2 removal. The slurry reactor thus starts with a distinct advantage in terms of gas preparation. This is partially balanced by a large downstream CO_2 removal requirement. Nevertheless, it would be expected that, if the reactors are competitive in cost, the overall slurry reactor processing scheme would show a cost advantage.

In the following subsections, the design of the F-T reactors is reviewed and reactor drawings and costs are provided for both cases. The slurry reactor benefits from operating at a higher temperature level and a higher conversion level. Overall reactor dimensions are similar but only 6 reactors are required as compared to 8 fixed-bed reactors. Differential capital investments and operating costs for the two processing systems are then provided. Finally, some implications of changes in the reactor design parameters are discussed.

It is emphasized that these capital costs are for those selected process units which differ depending on which reactor is used, so only the differential costs are truly meaningful. Reference should be made to the block flow diagrams given in Figures 4.2 and 4.3 to see which units are covered and not covered by the estimate. Using previous Bechtel and MITRE studies, however, it is possible to put these differentials in perspective relative to the overall cost of a coal-to-liquids processing scheme.

6.1.1 Reactor Costs

The reactor design bases provided in Tables 4.6 and 4.8 were reviewed by Bechtel process, mechanical and materials experts who provided the process sizing criteria, material selection and vessel design basis. The resulting designs are shown in Figure 6.1, for the slurry reactors, and Figure 6.2, for the tubular-fixed-bed reactors.

These designs are preliminary. There will be specific site-related mechanical design criteria, for example, that need to be considered. There are also cost optimization factors to consider such as whether to design to Section VIII Division 1 or Division 2 of the ASME code. The former, used for this design, is more conservative but the latter requires a greater number of inspections, increasing the cost of manufacture.

Reactor costs were estimated by Bechtel based on cost quotations from related jobs and studies such as the California Fuel Methanol Cost Study. Confirmation was sought by obtaining quotations from Deggendorfer Werft und Eisenbau GmbH through their U. S. representative, the Ferrostaal Corporation. Agreement was good when all factors were taken into consideration.There is considerable variation, however, depending on the tightness of the market for equipment and on currency exchange rates.

The delivered cost of the reactors shown in Figures 6.1 and 6.2 is estimated at $2.85 MM and $4.3 MM, respectively. This cost includes shipping charges at $1000/ton but does not include any import duty since alternative U.S. sources are available. The lower cost for the slurry reactor reflects its lower overall weight which results from the simplification of the bottom head and the reduction in the weight of heat exchange tubes. Six slurry reactors are required as compared to eight fixed-bed reactors so the total cost of reactors is $17.1 MM versus $34.4 MM. The slurry reactor requires a number of auxiliary pieces of equipment such as cyclones, hydroclones, filters, centrifuges and a catalyst makeup and pretreatment system. Some of these can be common to a large number of reactors. When all such equipment is taken into consideration the delivered cost of the reactor systems rises to $23.3 MM and $35.4 MM, respectively.

6.1.2 Capital Investment

As shown in Table 6.1, a cost savings of $91.4 MM is estimated for a 20,000 BPSD coal-based Fischer-Tropsch plant using slurry reactors versus a comparable one using fixed-bed reactors. Some 85% of this savings is identified with the process plants and 15% with the utility plants. Judging from MITRE's study reported in WP89W00144-1 (February 1990), a complete facility of this size starting from coal and producing finished products would cost in the neighborhood of $1.08 billion (this estimate factors MITRE's total plant investment of $3.6 billion for an 80,000 BPSD plant by a 0.9 capacity exponent and adds 4% for escalation). Thus the projected cost savings are on the order of 8.5% of the total plant investment.

6.1.3 Operating Costs

Comparative operating costs are shown in Table 6.2 and are summarized below:

Variable Costs	Slurry Reactor $MM/yr	Fixed-Bed Reactor $MM/yr
Fuel Gas	12.44	17.58
Raw Water	0.25	0.30
Catalysts and Chemicals	14.04	8.20
Total Variable Costs	**26.73**	**26.08**
Fixed Costs[7]		
Maintenance @3% of Investment/yr.	**10.66**	**13.27**
Total Selected Operating Costs	**37.39**	**39.35**
Operating Cost Differential	**1.96**	

The largest single operating cost items are the fuel gas costs, the F-T catalyst replacement costs and maintenance. The fuel gas requirements are due to an imbalance in power and heating demands as compared to what could be supplied by heat recovery, including that from the gasifier. It was difficult to find a good use for the low pressure steam generated in the F-T reactors in the fixed-bed case. These initial results indicate that it is not necessary to achieve as high a synthesis gas utilization to liquid products as was done in this study, if this would result in a cost reduction.

In calculating the F-T catalyst makeup requirement the slurry reactor case assumes a 60 day catalyst life, following MITRE's lead. For the fixed-bed reactors a one year life is assumed since anything less than this would be impractical and inconsistent with a 90% on-stream factor. It is understood that Sasol dumps the ARGE catalyst more frequently than this but indications are that Shell expects a reasonable catalyst life in their Middle Distillate Process. If a one year equivalent life could also be demonstrated for the slurry reactor the operating cost differential would rise to $9.06 MM/year.

In the slurry reactor case, 16.1 MW of power are available for export. If a market exists at, say, 2.5 cents per kWh, this would represent an additional annual revenue of $3.17 MM for that case. There are also small differences in the relative distribution of products between cases as shown in Appendix F. No particular significance can be assigned to these differences, however, since no attempt was made to identify true differences in yield between cases. As mentioned in Section 4, the recovery or disposal of oxygenates is a problem requiring further study in both cases and the assumption of equal oxygenate yields may be an oversimplification.

6.1.4 Discussion

The development of the capital cost estimates is documented in Table 6.3 where total costs for each type of equipment and bulks are itemized for both cases. The F-T reactor system costs given in Subsection 6.1.1 represent 25% of the identified major equipment costs in the slurry reactor case and 29% in the fixed-bed reactor case.

An alternative case has been developed in which the number of slurry reactors is increased from 6 to 11. Reactor system costs are now comparable to the fixed-bed case. This

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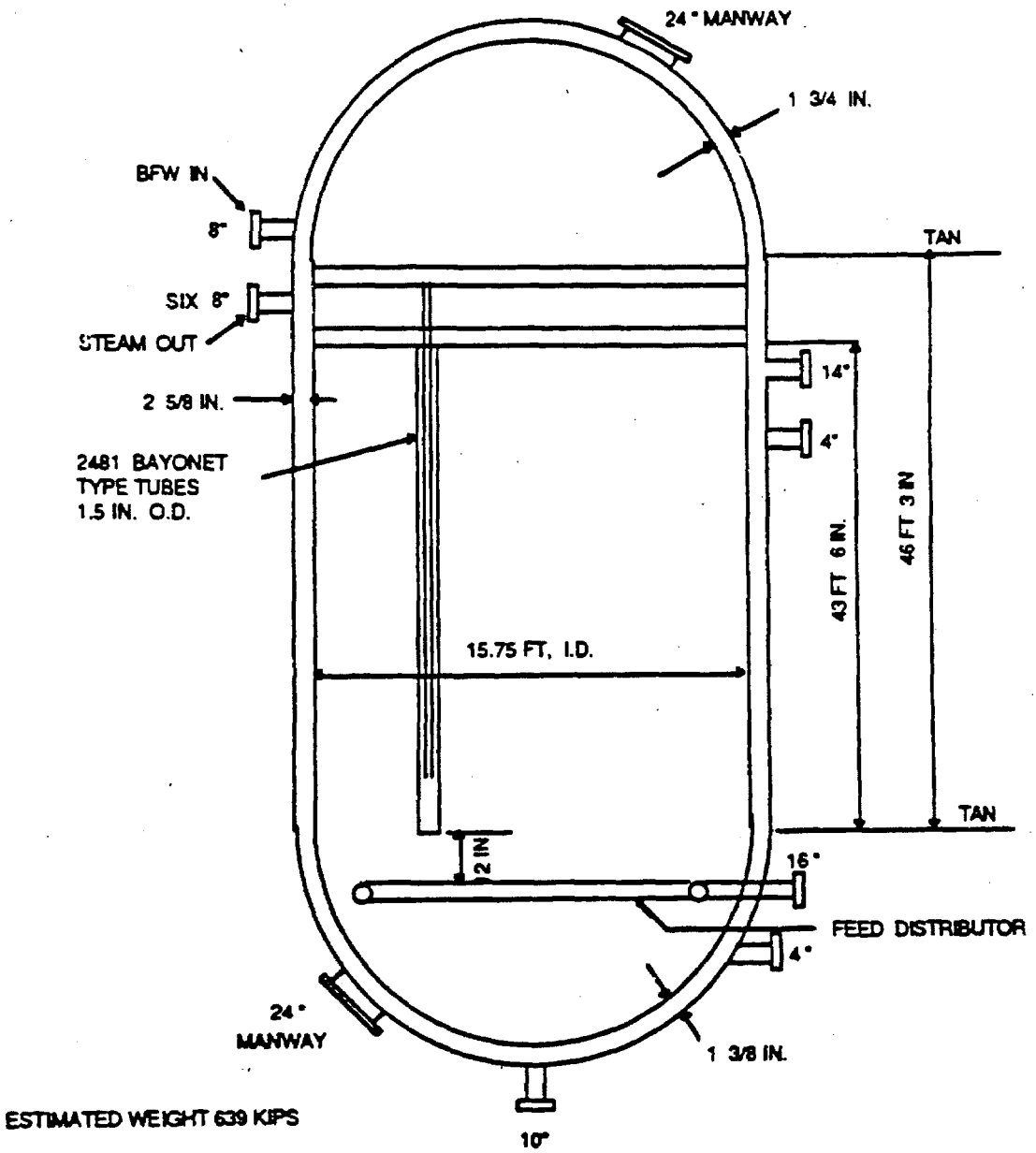
[7] Other fixed costs are deemed not to vary between cases.

increases the cost of the F-T synthesis section from $49 MM to $77 MM and increases total plant investment from $372 MM to $411 MM, cutting the differential in favor of the slurry case from $91 MM to $52 MM. Obviously, it is important to demonstrate that the reactor design conditions assumed for this study can be achieved.

The present estimate compares roughly as might be expected with MITRE's figures where a comparison can be made. MITRE's "plant construction cost" corresponds in scope to Bechtel's "total field cost" but is for a plant four times as large. MITRE's cost for Sulfur Removal, Shift, Fischer-Tropsch Synthesis, Autothermal Reforming and F-T Catalyst Preparation is $877 MM. Scaling down by the 0.9 or 1.0 capacity exponent and allowing 4% escalation, this corresponds to a range from $228 to $262 MM. Bechtel's cost for COS Hydrolysis, Acid Gas Removal, SynGas Compression, F-T synthesis and the F-T gas plant is $195 MM, at the field cost level, but in the alternate case with 11 reactors this is increased to $226 MM. MITRE would have used 14 or 15 reactors of roughly comparable dimensions, but running at lower pressure, for the same capacity.

Increased pressure improves the capacity of either the fixed-bed or slurry reactor. Because of the superficial velocity limitation, doubling pressure doubles the capacity of a given diameter slurry reactor (neglecting the area occupied by the cooling tubes). It also doubles the reactor wall thickness (excluding corrosion allowance). The fixed-bed reactor is more complicated but a good rule of thumb is that capacity increases as the square root of pressure, which keeps pressure drop constant. In this case, however, only the wall thickness of the heads and tube sheet are affected. For this reason, increasing pressure is expected to be more cost beneficial when using fixed-bed reactors. While the effect of pressure needs to be examined for both cases, the catalyst activity and selectivity data available to do so are extremely limited.

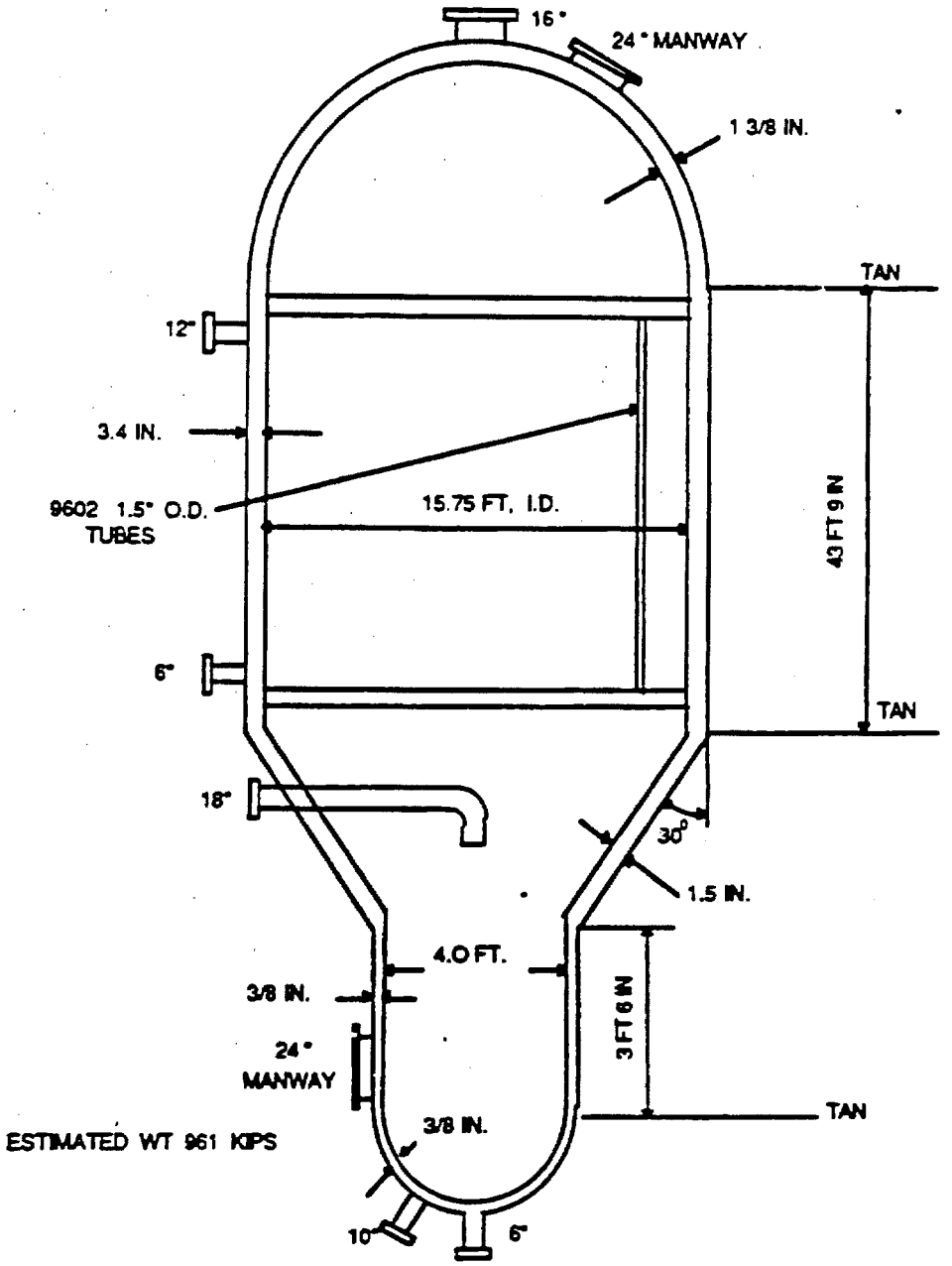
FIGURE 6.1
SLURRY REACTOR DESIGN STUDIES
FISCHER - TROPSCH PLANT
SLURRY REACTOR



NOTES:
(1)DESIGN CONDITIONS- TUBESIDE 580 PSIG,550 F ; SHELLSIDE 460 PSIG 550 F
(2) METALLURGY: SHELL - SA516 GR 70 WITH 1/8" C.A. ; TUBES C.S. CHROMIZED;TUBESHEET A516 GR70 WITH 1/8" C.
(3) ALL NOZZLES ARE 400 LB CLASS

Figure 6.2

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



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

Table 6.1

Capital Cost Comparison
Fischer-Tropsch Cases - Selected Units
$Millions

	Slurry Reactor		Fixed-Bed Reactor	
Process Plants:				
From Equipment Lists				
Shift Conversion	NA		22.4	
COS Hydrolysis	15.3		NA	
Acid Gas Removal	43.3		65.7	
SynGas Compression	7.2		11.6	
F-T Synthesis	49.0		82.1	
F-T Gas Plant	11.3		32.4	
CO2 Removal	47.6		14.3	
Subtotal from Lists	**173.8**		**228.4**	
From Cost-Capacity Curves		BPSD		BPSD
Catalytic Polymerization	4.6	1249	5.9	1674
HGO Hydrotreater	0.5	356	0.4	249
Gasoline Alkylation	5.7	1409	7.5	2136
Subtotal from Curves	**10.8**		**13.8**	
Subtotal Onsites	**184.6**		**242.2**	
Offsites:		Capacity		Capacity
Power Generation	20.4	61.4 MW	16.5	42.2 MW
Cooling Water	20.0	221 Mgpm	18.9	209 Mgpm
Waste Water Treatment	15.8	1900 gpm	18.3	3200 gpm
Raw Water Treatment	12.6	6640 gpm	14.2	7788 gpm
Sour Water Stripping	9.9	461 gpm	17.4	1036 gpm
Subtotal Offsites	**78.7**		**85.2**	
Total Direct Cost	**263.3**		**327.4**	
Contractor's Indirects	34.3		43.3	
Total Field Cost	**297.6**		**370.7**	
Eng'ng + Cont. @ 25%	74.4		92.7	
Total Project Cost	**372.0**		**463.4**	
Cost Differential		91.4		

Table 6.2
Fischer-Tropsch Operating Costs
Selected Cost Items
90% On-Stream Factor

		Slurry Reactor		Fixed-Bed Reactor	
Variable Costs	Unit Cost	Quantity/hr	$MM/yr	Quantity/hr	$MM/yr
Fuel Gas	$2.50/MMBtu	631 MMBtu	12.44	892 MMBtu	17.58
Raw Water	$0.08/MGal	398 MGal	0.25	467 MGal	0.30
Catalysts					
COS Hydr.	$377/CF	18.5 CF	2.30	NA	
CO Shift	$242.5/CF	NA		8.9 CF	0.71
F-T	$2.23/Lb.	11.8 Mlb. (60 day life)	8.69	5.3 Mlb. (1 year life)	3.90
Subtotal Catalysts			**10.99**		**4.61**
Chemicals					
Selexol	$2.00/lb.	98.3 lb.	0.07	NA	
Rectisol	$0.06/lb.	NA		5506 lb.	0.11
MEA	$0.56/lb.	4277 lb.	0.79	645 lb.	0.12
Offsites Chemicals (unit cost is cost per gpm treated)					
Water Trtg.			0.90		2.10
Raw	$10/yr	6638 gpm		7788 gpm	
Demin.	$1086/yr	657 gpm		1743 gpm	
BFW	$1518/yr	84 gpm		84 gpm	
Cooling Twr.	$907/yr	1304 gpm	1.25	1273 gpm	1.22
Effl. Trtg.	$43/yr	866 gpm	0.04	859 gpm	0.04
Subtotal Chemicals			**3.05**		**3.59**
Total Variable Costs			**26.73**		**26.08**
Fixed Costs	@3% of Investment/yr.		**10.66**		**13.27**
Total Selected Operating Costs			**37.39**		**39.35**
Total with 1 yr F-T life			**30.29**		

Table 6.3

Fischer-Tropsch Equipment Cost Summary
(Including Installation Labor)

All Plants

	Slurry Reactor $MM	Fixed-Bed Reactor $MM
Major Equipment		
Pressure Vessels	40.32	54.16
Tanks	1.31	0.55
Exchangers	30.08	43.15
Fired Heaters	0.50	0.90
Pumps and Drivers	2.11	1.96
Compressors	18.83	24.65
Package Equipment	2.57	
Total Major Equipment	**95.71**	**125.38**
Bulks	**78.15**	**103.05**
Total Direct Cost	**173.87**	**228.43**

6.2 Methanol and Mixed Alcohols

The block flow diagram for both types of reactors is given in Figure 4.1. As described in Section 4.1, only the methanol synthesis loop differs between cases. Methanol production is identical at 1488 tonnes per day (1640 short tons per day). The process flow diagram and equipment list for the methanol cases can be found in Appendix E.

In order to achieve capacity, the slurry reactor is operated at a pressure of 10,000 kPa and a recycle to fresh feed ratio of 2.2. The fixed-bed reactor operates at 5600 kPa. While the fixed-bed design is for a recycle to fresh feed ratio of 3 to 1, Lurgi has advised that a 4 to 1 ratio is required but that the pressure balance can remain as shown. This is not reflected in the equipment lists but adjustments have been made to the final cost estimate which allow for the resulting change in capacity of the recycle compressor and exchangers.

Methanol presents an entirely different situation than Fischer-Tropsch since conversion per pass is limited by equilibrium. To achieve high ultimate conversions to methanol a recycle operation is required. High pressure is needed to achieve capacity in a slurry reactor and this is a serious disadvantage since the vessel walls become very thick and feed gas compression is required. Recycle operation, however, suits the fixed-bed design very well. With a stoichiometric feed gas, operation at the pressure level available from a Texaco gasifier is possible and has actually been demonstrated at the Tennessee-Eastman facility.

The design of the slurry methanol reactor is provided in Figure 6.3, but the design of the tubular fixed-bed reactor is proprietary to Lurgi. Overall reactor dimensions are 4.8 m I.D. by 7.5 m T-T height for the fixed-bed reactor and 4.8 m I.D. by 16.25 m T-T height for the slurry reactor. The slurry reactor shell is designed for 11000 kPa (1600 psig), whereas the fixed-bed shell is designed for a maximum steam pressure of 4700 kPa (675 psig). Delivered cost of the fixed-bed reactor is estimated at $4.6 MM, that of the slurry reactor, at $7.9 MM. This includes shipping at $1000/ton. The slurry reactor synthesis loop requires a feed gas compressor and auxiliary slurry handling equipment such as holding tanks, a cyclone, filters, pumps and a catalyst prereduction system. All in all the slurry reactor system is projected to cost $41 MM and the fixed-bed system $23 MM. The breakdown on costs is given in Table 6.4.

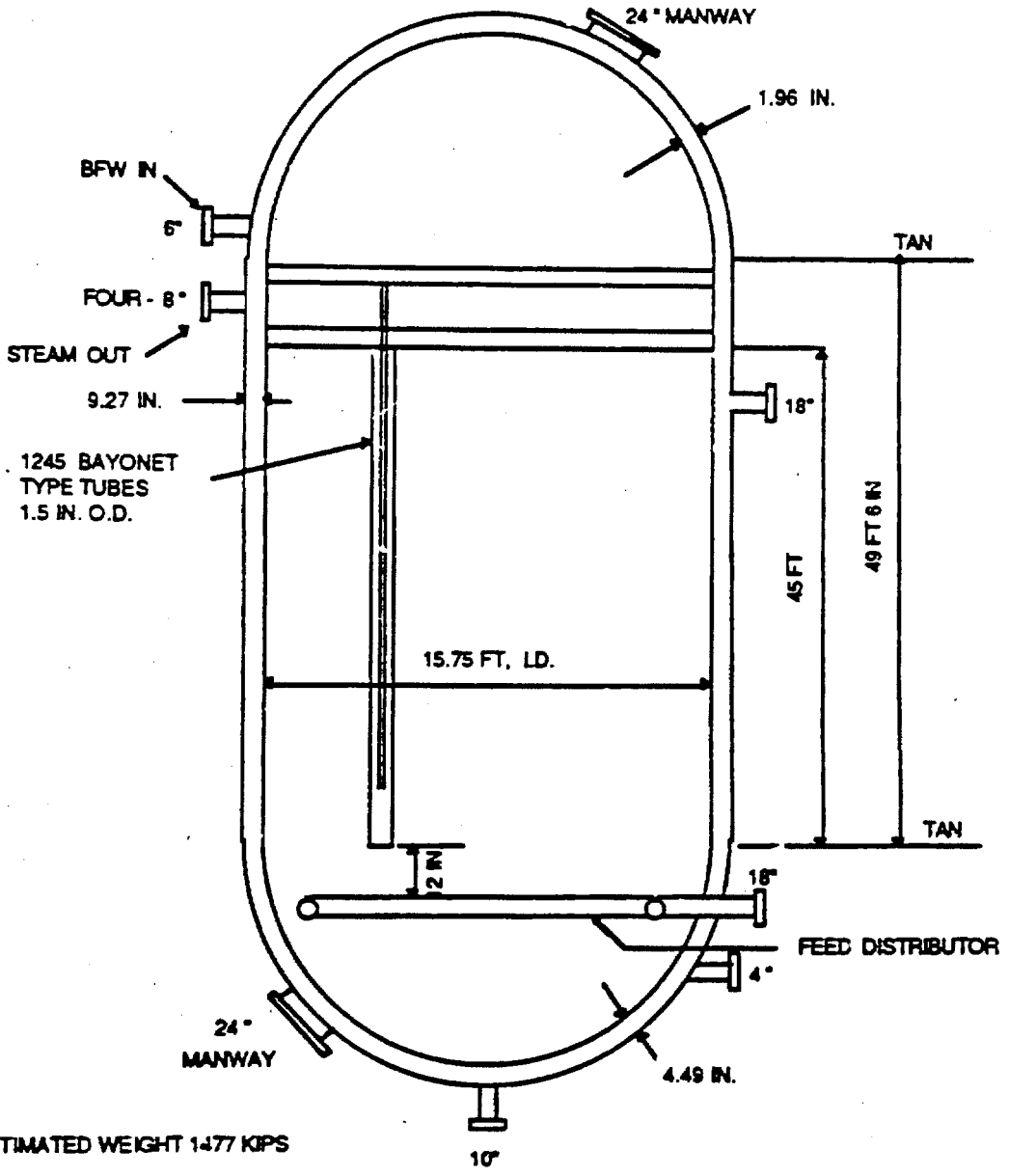
Several comments are necessary on this result. The slurry reactor has not been proposed seriously for recycle type methanol operations but rather has been aimed at coproduction of methanol and power via a once-through operation, at low conversion, on gas without H_2/CO ratio adjustment. Pressure can be on the order of 5600 kPa with little reduction in capacity since there is no recycle. In addition, it is understood that Air Products feels that superficial velocity can be increased up to 0.25 m/s. All these factors will reduce the cost.

It would be of interest to compare the fixed-bed and slurry reactors for once-through methanol operation if appropriate data can be obtained from the licensors of the technology. While it is conceivable that the fixed-bed reactor could be operated under such conditions there is no publicly available data on which to base a design.

The fixed-bed reactor is operated under low H_2/CO ratio conditions in Lurgi's Octamix™ process and the design of such a system is provided in Appendix E. This proposed new technology is a low space velocity, recycle operation intentionally producing mixed alcohols as a superior automotive fuel. As described in Section 4.2, the economic potential of the slurry reactor for this type of operation depends on whether design conditions can be altered in the direction of higher temperature and higher space velocity as shown in Table 4.2. It would also be essential for an economically competitive design to be able to run at a

lower operating pressure than the 99 atmospheres specified by Lurgi, because of the effect of pressure on the shell thickness of a slurry reactor. Bechtel's conclusion is that the mixed alcohols application does not appear worth pursuing further.

Figure 6.3
SLURRY REACTOR DESIGN STUDIES
METHANOL PLANT
SLURRY REACTOR



NOTES:
(1)DESIGN CONDITIONS- TUBESIDE 675 PSIG,550 F ; SHELLSIDE 1600 PSIG 550 F
(2) METALLURGY: SHELLSIDE - 304SS CLAD SA220; TUBES SA669,TUBESHEET SA204C +309SS OVERLAY ; TOP HEAD- A516GR 70 W 1/8" CA

Table 6.4

Methanol Synthesis Section
Equipment Cost Summary
(Including Installation Labor)

	Slurry Reactor		Fixed-Bed Reactor
Major Equipment	$MM		$MM
Pressure Vessels	8.81		5.05
Tanks	0.09		0.24
Exchangers	0.71		1.68
Pumps and Drivers	0.28		0.04
Compressors	5.86		2.26
Package Equipment	0.43		
Total Major Equipment	**16.18**		**9.02**
Bulks	**13.24**		**7.39**
Total Direct Cost	**29.42**		**16.41**
Contractor's Indirects	3.43		1.94
Total Field Cost	**32.85**		**18.35**
Eng'ng + Cont. @ 25%	8.21		4.59
Total Project Cost	**41.06**		**22.94**
Cost Differential		18.12	

Table of Nomenclature

a	gas-liquid interfacial area, m^{-1}
Bo	Bodenstein Number = $d_R^2 \cdot \rho_L \cdot g/\sigma_L$
c	heat capacity of the fluid, Btu/(lb·°F)
C	pressure drop coefficient in $ft \cdot hr^2/in^2$
C'Cat	catalyst concentration, kg/m^3
C_{HG}	hydrogen concentration in gas phase, kg mole/m^3
C^*_{HL}	hydrogen concentration,liquid, in equilibrium with gas, kg mole/m^3
C_{HL}	hydrogen concentration in the liquid phase, kg mole/m^3
d	effective particle diameter, ft
d_R	I.D. of reactor, cm
D	internal tube diameter (fixed-bed), ft
D_A	diffusivity of component A, m^2/s
D_L	Axial dispersion coefficient, cm^2/s
D_H	diffusivity of hydrogen, m^2/s
f	friction factor dependent on the modified Reynolds Number, dG/μ
Fr_G	Froude Number = $u_G^2/(g \cdot d_R)$
g	gravitational acceleration in consistent units
G	superficial mass velocity, $lb/(h \cdot ft^2)$.
GHSV	Gas hourly space velocity, Nm^3 (H_2+CO)/[h · m^3 reactor volume], (reactor volume is expanded slurry height times cross section area)
h	heat transfer coefficient, Btu/(h.ft^2·°F) or W/(m^2·s)
H	solubility coefficient of hydrogen = C_{HG}/C^*_{HL}
He	Henry's law constant, kPa·cm^3/mol
I	Inlet ratio of CO/H_2
k	thermal conductivity, Btu/(h·ft^2·°F/ft) or equivalent SI units
k_A	overall reaction rate constant defined by $1/K_A = 1/k_L a + 1/k_r \epsilon_L$
k_r	rate constant in volume/(unit volume · time) for H_2. $k_r = k_H = k'_H \cdot (kgCat/m^3)$
k_L	liquid side mass transfer coefficient, m/s
k_H	effective reaction rate constant for hydrogen consumption, s^{-1} (note that to agree with space velocity in $Nm^3/[s \cdot kgCat]$, $k_H = k_H' \cdot C'Cat$
k_H'	rate constant for hydrogen in volume/(mass catalyst · time), $m^3/[kg \cdot s]$
L	Length of expanded slurry bed or of fixed-bed, m or ft
P	pressure, kPa
r	rate of hydrogen consumption, $r = k_H \cdot C_{HL}$, kg moles/$[m^3 \cdot s]$
St	Stanton Number = $-K_A/SV$
SV	Space velocity in actual m^3 inlet gas/$[s \cdot m^3]$
T	temperature, °K
u_G	superficial gas velocity, cm/s or m/s
u_{Go}	inlet superficial gas velocity
U	Usage ratio of CO/H_2
X_H	hydrogen fractional conversion per pass (If U = I, $X_H = X_{CO}$)
α	contraction factor, α=[m^3/s($X_{H_2+CO=1}$)-m^3/s(inlet)]/[m^3/s(inlet)]

α^*	contraction factor modified for H_2 conversion, $\alpha^* = \alpha \cdot (1+U)/(1+I)$
ΔP	pressure drop, psi or equivalent SI units
ε_G	fractional gas hold-up
ε_L	fractional liquid hold-up
μ_L	liquid viscosity, poise or lb/(h·ft)
ρ_L	liquid density, g/cm^3 or lb/ft^3
σ_L	surface tension, N/m