SYNTHESIS OF DIMETHYL ETHER AND ALTERNATIVE FUELS IN THE LIQUID PHASE FROM COAL-DERIVED SYNTHESIS GAS

Final Technical Report

Contractor

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Contents

INTRODUCTION.....	1
TASK 2 SYNTHESIS GAS CONVERSION TO DME AND METHANOL FOR DIRECT USE	3
Subtask 2.1 Process Evaluation Performance Target	3
Subtask 2.2 Definition of Preferred Catalyst System	3
Subtask 2.3 Process Variable Scans on the Preferred Catalyst System	3
Subtask 2.4 Life-Test on the Preferred Catalyst System	4
Subtask 2.5 Detailed Process and Economic Evaluations	4
Subtask 2.6 Plan for Demonstrating LPDME in the LaPorte AFDU	4
TASK 3 ONE-STEP LIQUID PHASE SYNTHESIS OF ALTERNATIVE FUELS	5
Subtask 3.2 Screen Novel Catalyst System	6
Subtask 3.3 Evaluation of Preferred Catalyst System	6
Subtask 3.4 Update Process and Economic Evaluations.....	7

INTRODUCTION

From September 1981 to October 1991, Air Products and Chemicals, Inc., with the support of the U.S. Department of Energy (DOE), Chem Systems Inc., and the Electric Power Research Institute (EPRI), developed the liquid phase methanol (LPMEOH) process. The process, which uses slurry reactor technology, has the potential to be a more cost-effective route to produce methanol from coal than conventional gas phase processes. Primarily because of the presence of the well-mixed liquid phase in the reactor, the process is particularly suited to coal-derived synthesis gas, normally rich in carbon monoxide.

Extensive testing of the technology started in laboratory reactors and was extended to a DOE-owned, nominal 10 ton-per-day (TPD) Process Development Unit (PDU) located at LaPorte, Texas. In 1990 DOE renamed the PDU the Alternative Fuels Development Unit (AFDU) to emphasize DOE's interest in extending slurry phase processing to the development of a range of alternative fuels. In addition to different simulated coal derived feedstocks, process variables such as temperature, pressure, and flow rate were investigated thoroughly in both the laboratory and AFDU. A 120-day catalyst activity maintenance test in 1989 confirmed superior methanol catalyst stability in the liquid phase reactor compared to conventional packed-bed reactors. An on-stream factor at pilot level of greater than 98% was achieved. Furthermore, significant effort was utilized to maximize reactor productivity. Very high catalyst slurry concentrations (up to 50 wt%) and high gas linear velocity (up to .72 ft/sec) were successfully demonstrated in the slurry bubble column reactor at LaPorte.

After these process and engineering development efforts, the LPMEOH process was deemed ready for the next scale of demonstration towards commercialization. Currently, a plan is in place to build a 200-500 TPD LPMEOH unit under DOE's Clean Coal III Program.

Through the mid-1980s, Air Products has brought the liquid phase approach to a number of other synthesis gas reactions where effective heat management is a key issue. In 1989, in response to DOE's PRDA No. DE-RA22-88PC88805, Air Products proposed a research and development program entitled "Synthesis of Dimethyl Ether and Alternative Fuels in the Liquid Phase from Coal Derived Syngas". The proposal aimed at extending the LPMEOH experience to convert coal-derived synthesis gas to other useful fuels and chemicals.

The work proposed included development of a novel one-step synthesis of dimethyl ether (DME) from syngas, and exploration of other liquid phase syntheses of alternative fuels directly from syngas. The one-step DME process, conceived in 1986 at Air Products as a means of increasing syngas conversion to liquid products, envisioned the concept of converting product methanol *in situ* to DME in a single reactor. The slurry reactor based liquid phase technology is ideally suited for such an application, since the second reaction (methanol to DME) can be accomplished by adding a second catalyst with dehydration activity to the methanol producing reactor. An area of exploration for other alternative fuels directly from syngas was single-step slurry phase synthesis of hydrocarbons via methanol and DME as intermediates. Other possibilities included the direct synthesis of mixed alcohols and mixed ethers in a slurry reactor.

Based on this proposal, DOE Contract No. DE-AC22-90PC89865 was awarded to Air Products on 21 December 1989. The primary objectives of this contract were to:

- Develop the technology for the single-step, slurry phase synthesis of DME.
- Investigate other potential slurry phase syntheses of alternative fuels.

For DME development, a further objective was to develop the process sufficiently for demonstration in the AFDU at LaPorte. The objective for other alternative fuels was to screen several novel catalyst systems and recommend preferred routes for further investigation.

The scope of work to be performed by Air Products was divided into three tasks as shown in Table 1. A detailed discussion of each task/subtask was presented in the Management Plan dated 21 February 1990.

TABLE 1 WORK BREAKDOWN STRUCTURE

Task 1. Project Management

Task 2. Synthesis Gas Conversion to DME/Methanol for Direct Use

- Subtask 2.1 - Process Evaluation Performance Targets
- Subtask 2.2 - Definition of Preferred Catalyst System
- Subtask 2.3 - Process Variable Scans on the Preferred Catalyst System
- Subtask 2.4 - Life-Test on the Preferred Catalyst System
- Subtask 2.5 - Detailed Process and Economic Evaluations
- Subtask 2.6 - Plan for Demonstration LPDME in the LaPorte AFDU

Task 3. One-Step Liquid Phase Synthesis of Alternative Fuels

- Subtask 3.1 - Product/Process Definition
- Subtask 3.2 - Screen Novel Catalyst Systems
- Subtask 3.3 - Evaluation of the Preferred Catalyst System
- Subtask 3.4 - Update Process and Economic Evaluations

The two topical reports shown below were issued separately to document experimental details from the laboratories.

1. Task 2.2 "Definition of Preferred Catalyst System", Task 2.3 "Process Variable Scans on the Preferred Catalyst System", Task 2.4 "Life-Test on the Preferred Catalyst System".
2. Task 3.2 "Screen Novel Catalyst Systems", Task 3.3 "Evaluation of the Preferred Catalyst System".

It is not intended to reproduce the content of each Topical Report in this Final Report. A brief summary of all subtasks is included. More detail can be found in the above mentioned Topical Reports.

TASK 2 SYNTHESIS GAS CONVERSION TO DME AND METHANOL FOR DIRECT USE

Subtask 2.1 Process Evaluation Performance Target

The objective of this subtask was to develop a process economic model of the one-step liquid phase DME (LPDME) process that can be used to set performance targets for catalysts and reactor systems. Under this subtask, a kinetic model was developed to describe the three reaction steps in the LPDME process,

Attachment A, a memo entitled "Kinetics of Methanol/DME Synthesis" provides details of the kinetic modeling.

Process flow diagrams for various LPDME applications are presented in Attachment B, "Economics of DME/Methanol Coproduction". Cost estimates were developed for three promising configurations.

Subtask 2.2 Definition of Preferred Catalyst System

Under this subtask, three methanol catalysts, five alumina-based dehydration catalysts, and one water-gas shift catalyst (all commercially available) were tested. A gas simulating the product from a Texaco coal gasifier, with a nominal composition of 51% CO, 35% H₂, 13% CO₂, and 1% N₂, was used as the feed for catalyst screening. Based on activity and slurry properties, BASF S3-86 and Catapal γ -alumina were selected as the preferred catalysts for the LPDME process. Depending on the application, these two catalysts can be physically mixed in different proportions to yield essentially any desirable combination of DME and methanol in the product stream.

Subtask 2.3 Process Variable Scans on the Preferred Catalyst System

The effects of pressure, temperature, gas-hourly-space-velocity (GHSV) and feed composition were studied. The catalyst system chosen for the process variable scans consisted of 81.25 wt% BASF S3-86 and 18.75 wt% Catapal γ -alumina. This proportion of catalysts was determined based on the maximum methanol equivalent productivity (MEP), which was defined as,

$$\text{MEP} = \text{methanol productivity} + 2 \times \text{DME productivity} \quad (4)$$

MEP measures not only syngas conversion to the methyl group but also relates directly to the heating value of the desirable product, since each molecule of DME contains almost twice the heating value of a molecule of methanol.

Subtask 2.4 Life-Test on the Preferred Catalyst System

Two catalyst activity-maintenance experiments were performed. One 620-hour experiment used the simulated Texaco gas as the feed. A subsequent 500-hour experiment utilized a simulated Shell gas as the feed. Both experiments employed a catalyst mixture of BASF S3-86 and Catapal γ -alumina in the optimized 81.25 to 18.75% weight proportion.

In both runs, the catalyst system experienced relatively fast deactivation compared to LPMEOH. Post mortem catalyst analyses by XRD and Raman Spectroscopy indicated copper crystallite growth on the catalyst but no graphitic carbon deposit.

Further studies on the catalyst deactivation phenomenon are strongly recommended.

Subtask 2.5 Detailed Process and Economic Evaluations

Process and economic evaluations for the LPDME process were conducted for three possible scenarios. Key conclusions from the study are:

- LPDME can produce DME more economically than in the conventional two-step process (methanol formation followed by dehydration), and the product holds promise as a feedstock for downstream chemical processes.
- DME and methanol mixtures as peak shaving fuels may be produced at a slightly lower cost than methanol alone. This cost difference is probably not sufficient to warrant the additional complexity associated with collecting and storing this fuel.
- Methanol with a small quantity of DME can be produced at essentially the same cost as methanol alone. The presence of DME may enhance the cold weather startability of methanol used as a fuel.

A detailed report is included as Attachment B.

Subtask 2.6 Plan for Demonstrating LPDME in the LaPorte AFDU

A plan for the demonstration of the LPDME process in the LaPorte AFDU was completed in conjunction with the follow-on Alternative Fuels DOE Contract No. DE-AC22-91PC90018. Since the latter contract included performance of the demonstration run, it makes more sense to report the plan in that contract; thus it is not included herein.

TASK 3 ONE-STEP LIQUID PHASE SYNTHESIS OF ALTERNATIVE FUELS

Subtask 3.1 Product/Process Definition

This task turned out to be very exploratory. The emphasis of Subtask 3.1 became product definition rather than process definition. Process definition was premature and inappropriate at this time.

Through a literature search, three syntheses showed potential for liquid phase application because of high acceptance probability of the product in the market place. These syntheses are:

1. Syngas conversion to hydrocarbons.
2. Syngas conversion to mixed alcohols.
3. Syngas conversion to mixed ethers.

Syngas conversion to hydrocarbons takes a similar approach to Mobil's MTG process. However, unlike the Mobil's multi-step process,

the current method employs only a single-step synthesis,

Conceptually, this single-step synthesis could be accomplished in a liquid phase reactor using a dual catalyst system consisting of a methanol catalyst and a zeolite catalyst (e.g. ZSM-5). The liquid phase would provide excellent temperature control for this highly exothermic reaction.

"Mixed alcohols" product has the potential to be an effective gasoline additive. All alcohols with a carbon number of less than 5 have relatively high octane numbers and provide oxygen for clean burning. Methanol alone as a fuel additive has poor water tolerance causing phase separation in wet gasoline. In addition, methanol has a very high blending Reid Vapor Pressure (RVP) which worsens the evaporative emissions. Commercially, Lurgi offers the Octamix process to produce mixed alcohols. The product from an Octamix unit is claimed to have a C₂+ alcohol to methanol ratio of 30 to 70 by weight. The presence of higher alcohols reduces the shortcomings of methanol as a gasoline additive. This 30 wt% higher alcohol content became the target of this subtask.

Many ethers are known in the industry as excellent gasoline blending agents. Methyl tert-butyl ether (MTBE), ethyl tert-butyl ether (ETBE), and tert-amyl methyl ether (TAME) are all well accepted in the market place. Below are the rationales for developing a single-step mixed ethers process.

- Ethers can easily be made from dehydration of alcohols.
- With the development of a mixed alcohols process, a mixed ethers process is established by adding a dehydration catalyst to the mixed alcohol reactor.
- A liquid phase reactor can accommodate the addition of a second catalyst easily, as demonstrated in the LPDME process.

Subtask 3.2 Screen Novel Catalyst System

Catalyst systems for the syntheses discussed above were evaluated.

For the one-step synthesis of hydrocarbons, a mixture of a methanol catalyst and ZSM-5 was used as the catalyst system. The stability of the slurrying liquid became a key issue. The liquids acceptable in the LPMEOH process cracked in the presence of ZSM-5 and hydrogen. Many attempts to mitigate the problem were tried with only limited success, including passivating the external acid site of the ZSM-5 catalyst and choosing different liquids. Although hydrocarbon products were observed, liquid stability remained an issue. The process is not considered viable without further work to identify a suitable liquid medium.

Two commercially available mixed alcohols catalysts and one laboratory prepared catalyst were tested for mixed alcohols synthesis. One of the commercial catalysts was the copper-based Octamix catalyst, and the other was a cobalt-based catalyst obtained through a cooperative secrecy agreement. The laboratory preparation was an alkali promoted BASF S3-86 catalyst. In addition, dual catalyst systems involving mixtures of a copper-based methanol catalyst and a higher alcohol forming co-catalyst were also evaluated. Among all the catalyst systems evaluated, the laboratory prepared alkali-promoted catalyst showed the highest activity for mixed alcohols.

Mixed ether synthesis was accomplished by adding a γ -alumina catalyst to the alkali-promoted BASF S3-86 catalyst. Various ethers were produced. The majority of the ingredients in the ether product were the statistical expectations from the mixed alcohols intermediate product. For example, DME, methyl iso-butyl ether, methyl ethyl ether, and methyl n-propyl ether were predominant ether products. Evaluation of these ethers as gasoline additives was conducted as part of the Alternative Fuels follow-on contract.

The screening of all these catalyst systems suggested the alkali-promoted BASF S3-86 for mixed alcohols synthesis showed the most promise for further investigation. In fact, preliminary tests showed that once-through conversions achieved the target of 30 wt% higher alcohol in the product.

Subtask 3.3 Evaluation of Preferred Catalyst System

The synthesis of mixed alcohols was chosen for further evaluation. Alkali-promoted S3-86 catalyst was studied for process variable effects. Temperature, GHSV, and gas composition were varied to observe the impact on catalyst productivity and selectivity. In addition, co-feeding methanol and ethanol with syngas was studied to simulate the recycle of light alcohols to increase the yield of higher alcohols.

Lurgi's Octamix catalyst was investigated in some detail because it is a "state-of-the-art" commercial catalyst for mixed alcohols synthesis. The effects of temperature, pressure, and gas composition were studied. The results obtained on the Lurgi catalyst provided a basis for performance comparison.

Subtask 3.4 Update Process and Economic Evaluations

As for Subtask 3.1, this subtask is premature at this time because the reaction system is still embryonic. Air Products has identified that one-step synthesis of mixed alcohols using an alkali-promoted catalyst deserves more detailed study. The optimization of the catalyst system requires more time and effort to finalize. The economics for mixed alcohols synthesis should be similar to that of methanol synthesis and should be done when the process becomes better defined.

ATTACHMENT A

Kinetics of MeOH/DME Synthesis

Memorandum

To: B. Bhatt Dept./Loc.: PSG R&D
From: D. M. Herron Dept./Ext.: PSG Process Engineering / 4765
Date: 7 May 1993
Subject: Kinetics of MeOH/DME Synthesis - REVISION OF 15 MARCH 1990 DOCUMENT

cc: D. M. Brown
T. H. Hsiung
D. W. Studer / Alt Fuels Modelling File (w/source code)
R. P. Underwood
R. F. Weimer

1.0 Summary -

I have completed calculations of the predicted product distribution from a CSTR for the reactions: syngas to methanol, methanol to dimethyl ether, and water-gas shift (all occurring simultaneously).

Based on the results of a single autoclave run (at one wt% MeOH catalyst), it is possible to estimate rate constants for each of these three reactions, then proceeded to predict the productivity and selectivity over the entire range of co-catalyst proportions.

The results of my calculations indicate the following:

- The maximum productivity (defined on an equivalent-methanol basis) will occur in the range of 70 wt% to 90 wt% MeOH catalyst. The MeOH rate constant can be determined accurately over the entire range of co-catalyst proportions.
- The DME selectivity (defined as the moles of DME divided by the moles of DME+MeOH) is fairly constant (around 0.8) in the range of 30 wt% to 70 wt% MeOH catalyst. Within this range it is difficult to accurately determine the DME reaction order and rate constant. Therefore, it will be best to screen DME catalysts at low proportions in the co-catalyst mix.
- The kinetics of the shift reaction will be difficult to determine with any accuracy (without the aid of water analysis in the reactor effluent). However, the most sensitive region is below 20 wt% MeOH catalyst.

I have examined runs 10454-74a, 74c, and 74e, and estimated rate constants. Based on this work, I offer the following comments:

- The activity of the MeOH catalyst (for the MeOH reaction) declined from 83% of expected (run 74a) to 50% of expected (run 74c) and finally to 40% of expected (run 74e).
- The activity of the DME catalyst and the MeOH catalyst (towards the shift reaction) both seemed to decline with time-on-stream. However, the rate constants are very sensitive to the measured composition of the reactor effluent - these observations need further experimental verification.

2.0 Model Development -

The reactions under consideration are:

2.1 Component Balances -

The component molar balances are:

FLOW RATE OUT - FLOW RATE IN (gmole/hr-kg total catalyst)	=	RATE OF PRODUCTION (gmole/hr-kg total catalyst)	
H ₂	=	- 2 E ₁	+ E ₃
CO	=	- E ₁	- E ₃
N ₂	=		
CO ₂	=		+ E ₃
MeOH	=	+ E ₁	- 2 E ₂
DME	=		+ E ₂
H ₂ O	=		+ E ₂ - E ₃
TOTAL	=	- 2 E ₁	

where E₁, E₂, and E₃ are the extents of reaction for rxn 1, 2, and 3 respectively.

The extent of reaction is defined by:

$$E_i = R_i X_i$$

where R_i is the rate of reaction (gmole/hr-kg catalyst i) and X_i is the weight fraction of catalyst i (kg catalyst i/kg total catalyst).

2.2 Reaction Rates -

The rates of reaction were assumed to be of the following (simplified) forms:

$$R_1 = k_1 (f_{\text{CO}})^{1/3} (f_{\text{H}_2})^{2/3} \left(1 - \frac{1}{K_{E1}} \frac{f_{\text{MeOH}}}{f_{\text{CO}} (f_{\text{H}_2})^2} \right) \quad (\text{rxn 1})$$

$$R_2 = k_2 (f_{\text{MeOH}})^N \left(1 - \frac{1}{K_{E2}} \frac{f_{\text{DME}} f_{\text{H}_2\text{O}}}{f_{\text{MeOH}}^2} \right) \quad (\text{rxn 2})$$

$$R_3 = k_3 (f_{\text{CO}}) (f_{\text{H}_2\text{O}}) \left(1 - \frac{1}{K_{E3}} \frac{f_{\text{CO}_2} f_{\text{H}_2}}{f_{\text{CO}} f_{\text{H}_2\text{O}}} \right) \quad (\text{rxn 3})$$

2.3 Solution Scheme -

For a CSTR, the rates of reaction are computed from the outlet component fugacities and the solution to this problem becomes straightforward:

- The inlet flow and composition, catalyst weights, reactor temperature and pressure, rate/equilibrium constants, and outlet fugacity coefficients are specified.
- The outlet flow and composition (the remaining unknowns) are computed by plugging the rate expressions into the component balances and solving the system of equations simultaneously (ie iteratively).

The solution scheme which I use is such that different forms of rate expressions may be used.

2.4 Equilibrium Constants -

The following expressions were used to describe the equilibrium constants:

$$\ln K_{EQ1} = 28.18 + 16251.2/T_R - 7.97 \ln(T_R) + 0.0032 T_R - 2.1 \times 10^{-7} T_R^2 \quad (\text{atm}^{-2})$$

$$\ln K_{EQ2} = -13.36 + 2835.2/T_K + 1.68 \ln(T_K) - 2.4 \times 10^{-4} T_K - 2.1 \times 10^{-7} T_K^2 \quad (-)$$

$$\ln K_{EQ3} = -4.33 + 8240/T_R \quad (-)$$

where T_R and T_K are temperature in $^{\circ}R$ and $^{\circ}K$, respectively. The functional forms for the equilibrium constants of rxn 1 and 3 were taken from reference 1; reference 2 was used as the source the equilibrium constant of rxn 2.

2.5 Rate Constants -

The rate constants k_1 , k_2 , and k_3 were estimated from selected autoclave data (Runs 10454-74A, 74C and 74E). Note that the order of the DME reaction, N , was assumed to be either 1 or 2. The results are summarized in the Tables 1, 2, and 3.

Also note that the experimental effluent composition was corrected for the presence of water (which was not measured in the lab). The water quantity in the effluent was estimated in such a way as to give the best overall agreement between the measured composition and that which would result from the indicated stoichiometry.

In general, the rate constants are shown to decrease with time-on-stream. In particular, the activity of the MeOH catalyst is decreasing substantially.

Once the rate constants have been obtained, it is possible to calculate productivity (in MeOH equivalent terms) and selectivity (towards DME) over a range of co-catalyst proportions. This exercise has been done - the results are plotted in Figure 1 (run 74A), Figure 2 (run 74C), and Figure 3 (run 74E). These results demonstrate that productivity and selectivity are greatest at the higher concentrations of MeOH catalyst.

3.0 Model Predictions / Sensitivity Studies -

Inspection of Figures 1 through 3 leads one to believe that the optimal MeOH catalyst concentration is between 70 wt% and 90 wt%. To challenge this conclusion, I have looked at the sensitivity of the trends to uncertainty in the various rate constants.

For this exercise, I have used the run 74a composition (CO-rich, CO₂ lean) as a basis.

3.1 Variation in MeOH Rate Constant -

I considered three levels of MeOH catalyst activity: 65%, 83% and 100% of expected. The "expected" rate constant is that which prevails under Liquid-Phase Methanol synthesis conditions (pre-exponential factor, $K_0=2.91 \times 10^6$ gmole/hr/kg-catalyst/atm). The results of my calculations are summarized in Figure 4 (Figure 4A for DME rxn order of 1, Figure 4B for DME rxn order of 2). These plots indicate:

- As MeOH catalyst activity increases the maximum productivity shifts to lower concentrations of MeOH catalyst.
- DME selectivity is unaffected by MeOH catalyst activity.

3.2 Variation in DME Rate Constant -

I considered three levels of DME catalyst activity. The results of my calculations are summarized in Figure 5 (Figure 5A for DME rxn order of 1, Figure 5B for DME rxn order of 2). These plots indicate:

- As DME catalyst activity increases the maximum productivity shifts to higher concentrations of MeOH catalyst.
- DME selectivity falls with decreasing DME catalyst activity.

3.3 Variation in Shift Rate Constant -

I considered three levels of Shift activity. The results of my calculations are summarized in Figure 6 (Figure 6A for DME rxn order of 1, Figure 6B for DME rxn order of 2). These plots indicate that shift reaction kinetics only impact selectivity and productivity modestly.

3.4 Variation in Space Velocity -

I considered three different space velocities. The results of my calculations are summarized in Figure 7 (Figure 7A for DME rxn order of 1, Figure 7B for DME rxn order of 2). These plots indicate:

- Productivity increases with increasing space velocity but the point of maximum productivity is largely unaffected by space velocity.
- DME selectivity falls with increasing space velocity. For greater selectivity, one is driven to lower concentrations of MeOH catalyst.

4.0 Conclusion -

These sensitivity studies strongly suggest that the commercial DME/MeOH synthesis should be carried out at 70 to 90 wt% MeOH catalyst.

A key piece of information which is missing is the kinetic expression of the DME reaction (reaction order in particular). As illustrated clearly in Figure 7, variation in DME selectivity is much stronger for first order kinetics.

D. M. Herron

5.0 References -

1. J. A. Ager, "LPMEOH Modeling", APCI memo to Distribution, 8 June 1989.
2. B. T. Diep and M. S. Wainwright, "Thermodynamic Equilibrium Constants for the Methanol-Dimethyl Ether-Water System", J. Chem. Eng. Data, 32, 330-333, 1987.

TABLE 1

Run 74A - 250 C, 52 atm, 2250 sL/kg-hr (total catalyst) - 50% MeOH Cat

	Measured		Corrected		Modelled	
	IN	OUT	IN	OUT	N=1	N=2
<u>Composition (mole %)</u>						
H ₂	39.8	22.2	39.6	21.8	21.4	21.3
CO	56.5	50.2	56.0	49.3	49.5	49.4
N ₂	1.2	1.8	1.2	1.8	1.8	1.7
CO ₂	3.4	14.2	3.4	13.9	14.1	14.1
MeOH		1.8		1.8	1.8	1.8
DME		10.7		10.5	10.4	10.4
H ₂ O		?		1.0	1.1	1.1
Flow (sL/hr)	67.6	48.0			46.6	46.5

Calculated Coefficients

MeOH	k ₁ (gmole/hr-kg cat-atm)	2.36	2.36
	Percent of typical	(82.5)	(82.5)
DME 1st order	k ₂ (gmole/hr-kg cat-atm)	28	
2nd order	k ₂ (gmole/hr-kg cat-atm ²)		35
Shift	k ₃ (gmole/hr-kg cat-atm ²)	0.95	0.95

TABLE 2

Run 74C - 250 C, 52 atm, 1500 sL/kg-hr (total catalyst) - 50% MeOH Cat

	Measured		Corrected		Modelled	
	IN	OUT	IN	OUT	N=1	N=2
<u>Composition (mole %)</u>						
H ₂	38.7	22.1	38.2	22.2	22.3	22.4
CO	47.7	39.4	47.1	39.6	39.5	39.6
N ₂	1.1	1.4	1.1	1.4	1.5	1.5
CO ₂	13.8	25.1	13.6	25.2	25.0	25.0
MeOH		1.6		1.6	1.6	1.6
DME		8.2		8.2	8.2	8.2
H ₂ O		?		1.8	1.8	1.7
Flow (sL/hr)	45.1	32.9			33.1	33.2

Calculated Coefficients

MeOH	k ₁ (gmole/hr-kg cat-atm) Percent of typical	1.40 (49.0)	1.40 (49.0)
DME 1st order	k ₂ (gmole/hr-kg cat-atm)	27	
2nd order	k ₂ (gmole/hr-kg cat-atm ²)		32
Shift	k ₃ (gmole/hr-kg cat-atm ²)	0.40	0.40

TABLE 3

Run 74E - 250 C, 52 atm, 1500 sL/kg-hr (total catalyst) - 50% MeOH Cat

	Measured		Corrected		Modelled	
	IN	OUT	IN	OUT	N=1	N=2
<u>Composition (mole %)</u>						
H ₂	36.5	23.2	36.9	23.7	23.7	23.7
CO	46.1	39.2	46.6	40.1	40.2	40.3
N ₂	1.0	1.3	1.0	1.3	1.3	1.3
CO ₂	15.3	24.6	15.5	25.2	25.4	25.4
MeOH		1.5		1.5	1.5	1.5
DME		6.5		6.6	6.6	6.6
H ₂ O		?		1.5	1.3	1.3
Flow (sL/hr)	45.1	34.5			34.8	34.8
<u>Calculated Coefficients</u>						
MeOH	k ₁ (gmole/hr-kg cat-atm)				1.12	1.12
	Percent of typical				(39.0)	(39.0)
DME 1st order	k ₂ (gmole/hr-kg cat-atm)				18	
2nd order	k ₂ (gmole/hr-kg cat-atm ²)					25
Shift	k ₃ (gmole/hr-kg cat-atm ²)				0.50	0.50

FIGURE 1

DME/MeOH Catalyst Performance – 250 C, 52.0 atm
(Based on run 74a – 40% H₂, 56% CO, 3% CO₂ – SV=2250)

FIGURE 2

DME/MeOH Catalyst Performance – 250 °C, 52.0 atm
(Based on run 74c – 38% H₂, 47% CO, 14% CO₂ – SV=1500)

FIGURE 3

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Based on run 74e - 37% H₂, 46% CO, 16% CO₂ - SV=1500)

LEGEND

PRODUCT N=1

SELECT N=1

PRODUCT N=2

SELECT N=2

NOTE:

The order of the DME Reaction is taken as either first (N=1) or as second (N=2)

FIGURE 4A

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Variation with MeOH Catalyst Rate Constant)

FIGURE 4B

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Variation with MeOH Catalyst Rate Constant)

FIGURE 5A

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Variation with DME Catalyst Rate Constant)

FIGURE 5B

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Variation with DME Catalyst Rate Constant)

FIGURE 6A

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Variation with Shift Rate Constant)

FIGURE 6B

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Variation with Shift Rate Constant)

FIGURE 7A

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Variation with Space Velocity)

FIGURE 7B

DME/MeOH Catalyst Performance - 250 C, 52.0 atm
(Variation with Space Velocity)

ATTACHMENT B

Economics of DME/MeOH Coproduction

ECONOMICS OF DME/MeOH COPRODUCTION

1. INTRODUCTION

The results of economic evaluations into the synthesis of DME and DME/MeOH mixtures from coal-derived syngas are presented. Several production scenarios were considered. These options were selected to investigate the merit of the three major reasons for producing DME:

- DME can be produced in small quantities with MeOH to increase the vapor pressure of the mixed fuel. This application is appropriate for improving the cold-start limitations sometimes associated with neat methanol as a motor fuel.
- DME/MeOH mixtures can be synthesized together to minimize the cost of producing a peak-shaving fuel in CGCC applications. The coproduction of the mixture can be accomplished with greater reactor productivity (expressed as BTU of fuel per unit of reactor volume) than MeOH alone. As a result, the plant size may be reduced for the same quantity of fuel produced.
- DME may be produced for use as a chemical feedstock in a single step. The current, two-step approach is to produce (or purchase) MeOH, then dehydrate it to form DME. By utilizing the single-step syngas-to-DME process, feedstock cost may be minimized.

To summarize the results:

- "Cold-start" methanol may be produced at essentially the same cost as methanol alone.
- DME/MeOH peak-shaving fuels may be produced at a slightly lower cost than methanol alone. This cost differential is probably not sufficient to warrant the additional processing complexity associated with collecting and storing this mixed fuel.
- DME can be produced in a single step more cheaply than in the two-step process.

2. BASIS OF STUDY

The basis for this evaluation is that the MeOH or DME/MeOH plant would be added on to a CGCC facility. A nominal 600 MW CGCC facility using Shell Gasifiers and producing a total flow of approximately 328 MMSCFD of synthesis gas was selected. Up to 50% of this gas could be processed in the syngas-to-fuel plant. The unconverted syngas would be recombined with the unprocessed gas and directed to a turbine for power production.

The syngas to the fuels plant was assumed to be available at 260 psig and have a composition of: 30% H₂, 61% CO, 2% CO₂, 6.8% N₂, the remainder H₂O and sulfur impurities. The syngas heating value was 272 BTU/SCF (LHV). Unconverted syngas was to be returned at 260 psig.

The syngas-to-plant was compressed to 1200 psig, treated to remove sulfur, passed to a liquid-phase slurry reactor, and further processed to recover the fuel products. The unconverted syngas was subsequently expanded to 260 psig to recover power then sent to the main power turbine.

Steam was produced in the slurry reactor at 230 psig. A portion of this steam was utilized for product recovery and fuel vaporization while the bulk was exported to the CGCC for credit. Cooling water was available at 80°F and returned at 100°F.

3. CASE STUDIES

The results from seven case studies are presented in this report. The important features are summarized below and in Tables 1A and 1B.

Methanol Only:

Case 1 - This is the base case; the product is methanol which is suitable as either a motor fuel or a turbine fuel. The economic desirability of the other cases are to be measured against this case. The flowsheet for this case is a typical LPMEOH scheme with 1:1 recycle of unconverted syngas. The methanol product is stabilized to remove some CO₂ from the MeOH to maintain a vapor pressure of 7.2 psia at 100°F.

Methanol with DME for "Cold-Start":

Case 2 - A small amount of alumina is added to the reactor slurry resulting in the production of some DME. The recycle ratio is 1:1 as in Case 1. Additional processing steps are added to recover DME and a distillation column is required to reject CO₂ and keep the DME in the MeOH product. The resulting product contains 1.8 wt% DME and has a vapor pressure of 9 psia at 100°F. The vapor pressure of the product can be easily adjusted by making slightly more or less DME.

DME/MeOH Mixtures for Peak-Shaving Fuel:

Case 3 - This case is aimed at coproducing a larger quantity of DME. To take advantage of the inherently higher reactor productivity, no recycle is used. This has the effect of reducing the size of the process equipment associated with the reactor train as well as eliminating the recycle compressor. The equipment used to recover DME is not significantly more complex than that used in Case 2. The

resulting product contains 61 wt% DME and has a vapor pressure of 109 psia at 100°F.

Case 4 - This case is also aimed at coproducing a larger quantity of DME. In contrast with Case 3, unconverted syngas is recycled at a rate of 1:1. The use of recycle plus the higher reactor productivity allows the syngas feed flow to be significantly reduced relative to Case 1. This has the effect of reducing the size of the front-end compression and sulfur cleanup. However, the equipment used to recover DME is fairly complex relative to Case 3. The resulting product contains 58 wt% DME and has a vapor pressure of 112 psia at 100°F. (The increase in vapor pressure relative to case 3 is due to the higher CO₂ content.)

DME for Chemical Feedstock:

Case 5 - This case is developed to examine the production of DME-only in the simplest and most cost effective manner. The mode of operation is once-through (no syngas recycle) and the coproduced MeOH is recovered and recycled to the slurry reactor. There is no attempt to recover DME from the syngas; rather the DME produced exits the plant as a vapor together with the unconverted H₂ and CO. This mode of operation has applicability when the downstream process uses both CO and DME as chemical feedstock and can tolerate CO₂.

Case 6 - In this case, DME is separated from the unconverted syngas and is recovered as a vapor along with CO₂. As in Case 5, the mode of operation is once-through and the coproduced MeOH is recovered and recycled to the slurry reactor. The DME recovery scheme is fairly simple, along the lines of Case 3. This mode of operation has applicability when the downstream process requires DME as chemical feedstock and can tolerate CO₂.

Case 7 - The final case collects DME as a concentrated liquid (98 wt%). As in Case 5, the mode of operation is once-through and the coproduced MeOH is recovered and recycled to the slurry reactor. The DME recovery scheme is more complex than that used in Case 6. This mode of operation has applicability when the downstream process requires only DME and/or DME must be temporarily stored (product vapor pressure is 133 psia at 100°F).

Table 1A highlights the various options and ranks them according to relative capital cost and relative product cost. Note that all the comparisons are "relative to case 1. Table 1B presents pertinent plant and performance data in more detail.

4. PROCESS DESCRIPTIONS

The process descriptions for each of the cases share many similarities. A detailed description will be presented for Case 1 (MeOH only); differences from Case 1 will be highlighted for the other cases.

4.1 Case 1:

Figure 1.1 depicts an overall block diagram for the Fuels Plant. A portion of syngas from the gasifier is diverted from the saturator/superheater and directed to feed compression and sulfur removal. Power is consumed in this unit operation to drive the feed compressor and operate the closed loop propane refrigeration unit (refrigeration is used to aid sulfur removal). The cleaned syngas is combined with recycle gas and passed into the Liquid-Phase Synthesis unit operation. Steam is generated from this operation as the reaction is exothermic. Most of the steam is exported to the CGCC plant while a portion is used to operate the distillation column in the Product Recovery unit operation. Within the Product Recovery unit the methanol product is separated from the unreacted syngas; the methanol is subsequently stabilized in a fractionator to remove CO₂. The stabilized methanol is then directed to storage. The syngas from Product Recovery unit is split into two streams; one is recycled to the front end, the remainder is expanded (to recover power) and combined with the syngas from the saturator. Since this plant is producing peak-shaving fuel, the liquid product is periodically pumped, vaporized with steam, and combined with the main gas flow to the turbine.

Figure 2 illustrates the feed/recycle compression and storage/vaporization area in greater detail. Syngas-to-plant is compressed in two stages from 260 psig to 1200 psig (in the 01.10A&B). The compressor discharge is subsequently cleaned in the Sulfur/Carbonyl Removal unit operation. The Sulfur Removal unit is shown here as a black box. This unit is fundamentally an absorber/stripper arrangement which uses methanol as the absorbent and propane refrigeration to assist in providing condenser duties. Carbonyl removal is accomplished with packed beds. Recycle syngas from downstream processing is compressed in the 01.20 compressor and combined with the cleaned feed and directed to the Synthesis Section (Figure 3).

Figure 3 shows the Synthesis Section. The cleaned and compressed syngas is preheated in the 21.11 exchanger. Water is pumped to pressure in the 10.64 pump then combined with the hot feed. The mixed stream passes through the 02.20 Sulfur Guard Bed and on to the 27.10 Slurry Reactor. Within the reactor, the syngas is partially converted to methanol; the heat of reaction is used to generate 230 psig steam. Catalyst is continually added and removed during operation to maintain a steady rate of production. The reactor effluent enters the 21.11 exchanger and is cooled to approximately 230°F. Slurry oil is condensed in the 21.11 and returned to the reaction via the 10.50 pump. The net reactor effluent exits the 21.11 exchanger and flows to the Liquids Recovery section (Figure 4.1).

Figure 4.1 represent the Liquids Recovery section. The methanol bearing reactor effluent is cooled, first in the 21.33 Economizer then in the 21.31 Methanol Condenser, and passed into the 22.10 HP Methanol Separator. From the 22.10, methanol is removed as a liquid and unconverted syngas as a vapor. A portion of this vapor is returned to the recycle compressor (Figure 2); the remainder is heated in the 21.33 exchanger then expanded to 260 psig in the 10.10 Purge Gas Expander. The liquid from the 22.10 separator is let-down to 260 psig to degas dissolved vapors (in the 22.12 separator). The degassed vapors are combined with the 10.10 expander discharge and sent to the CGCC turbine header. The liquid from the 22.12 separator is directed to the stabilization section. Within the stabilizer methanol is boiled in the 07.73 (using steam) and refluxed (in the 07.72) to remove the remnants of dissolved CO₂ (which is sent to the incinerator, Figure 7). The stabilized methanol is then pumped and delivered to Storage (Figure 2). Returning to Figure 2, methanol from storage (28.40) is

periodically removed and pumped to 260 psig in the 10.40 pump. The liquid is subsequently vaporized in the 21.80 vaporizer and transferred to the turbine header to be used as peak shaving fuel. Alternatively, the methanol may be withdrawn from storage and sold as motor fuel.

Figures 5, 6, and 7 illustrate the associated unit operations. Figure 5 is a schematic of the Catalyst Preparation and Slurry Reduction sections. This section is used to activate the catalyst with dilute syngas (nitrogen diluent). Reactor temperature is controlled with circulating utility oil. The reduction off-gas is sent to the incinerator (Figure 7). This unit operates continuously to facilitate the on-line catalyst addition/withdrawal. Figure 6 shows the Catalyst/Oil Separation section. This operation uses a centrifuge to produce a spent catalyst cake; the recovered oil is returned to the reactor. Figure 7 illustrates the General Utilities section. The equipment associated with propane refrigeration, gas incineration, and the steam system are shown.

Figures 2, 3, 5, 6, and 7 are common to all the Cases considered in this study. Only the Overall Block Diagram (Figures 1.x) and Product Recovery sections (Figures 4.x) differ. The Overall Block Diagrams are self-evident; the subsequent discussions will focus on product recovery.

4.2 Case 2:

This mode of operation is very similar to Case 1 in that plant flows and recycle rate are similar. A small fraction of alumina catalyst (0.3 wt%) has been added to synthesize a small fraction of DME.

Figure 4.2 represents the Liquids Recovery section for Case 2. The MeOH/DME bearing reactor effluent is chilled in the 21.31 Methanol Condenser and passed into the 22.10 HP Methanol Separator. From the 22.10, methanol is removed as a liquid and unconverted syngas (plus DME) removed as a vapor. A portion of this vapor is returned to the recycle compressor (Figure 2); the remainder is split and cooled to -65°F, in the 21.32 Cross Exchanger and 21.33 Economizer, then recombined and passed into the 22.11 HP Liquids Separator. From the 22.11, additional methanol and much of the DME is recovered as a liquid; the light syngas is rejected as vapor. This vapor is then warmed in the 21.32 exchanger, expanded to 270 psig in the 10.10 Purge Gas Expander, then rewarmed in the 21.33 exchanger. The liquid from the 22.11 separator is let-down to 270 psig and warmed in the 07.72 exchanger then combined with the liquid from the 22.10 separator and introduced to the 22.12 MP separator to degas dissolved vapors. The liquid from the 22.12 is fractionated at 260 psig to remove dissolved CO₂. The 07.72 Condenser duty is supplied by vaporizing column feed liquid; reboil is provided with steam. The liquid bottoms product is chilled in the 21.71 Product Cooler then sent to Storage (Figure 2). The vapor overheads are combined with the degassed vapors from the 22.12 separator and with the 10.10 expander discharge and sent to the CGCC turbine header.

As in Case 1, the liquid may be subsequently vaporized in the 21.80 vaporizer (Figure 2) and transferred to the turbine header to be used as peak shaving fuel. Alternatively, this DME-containing methanol may be withdrawn from storage and sold as motor fuel.

4.3 Case 3:

The mode of operation is dissimilar to Case 2 in several ways. First, the reactor now contains 7.4 wt% alumina and consequently produces a significant amount of DME in addition to MeOH. Second, no syngas recycle is utilized. Finally, the Liquid Recovery section is more complex and the product liquid is stored at elevated pressure.

Figure 4.3 represents the Liquids Recovery section for Case 3. The MeOH/DME bearing reactor effluent is chilled in the 21.31 Methanol Condenser and passed into the 22.10 HP Methanol Separator. From the 22.10, MeOH and DME are recovered as liquid and unconverted syngas (plus DME) removed as vapor. This vapor is split and cooled, in the 21.32 Cross Exchanger and 21.33 Economizer, then recombined and passed into the 22.11 HP Liquids Separator. From the 22.11, additional MeOH and DME are recovered as a liquid and light syngas (plus DME) is rejected as vapor. The vapor is then expanded to 270 psig in the 10.10 Gas Expander, then introduced to the bottom of the 07.65 Absorber Column. The 07.65 column is refluxed with methanol. This allows the DME (and some CO₂) to be absorbed and recovered as bottoms. The vapor overhead from the 07.65 is warmed in the 21.32 exchanger. The liquid from the 07.65 is warmed in the 21.33 and 21.73 exchangers then combined with the liquids from the 22.10 and 22.11 separators to form the feed to the 07.71 Fractionation Column. Within the 07.71 column, CO₂ is rejected as vapor, the mixed DME/MeOH product is withdrawn as a side-stream, and purified methanol is withdrawn as bottoms. The condenser duty is provided in the 07.72 exchanger by vaporizing propane; the reboiler duty is provided in the 07.73 exchanger with steam. The vapor overhead is combined with the main gas flow from the 21.32 and sent to the CGCC turbine header. The liquid side draw is chilled in the 21.71 Product Cooler then sent to Storage (Figure 2). The liquid bottoms stream is cooled in the 21.72 Recycle Chiller, pumped in the 10.62, further cooled in the 21.73 exchanger, then returned to the 07.65 column as reflux.

As in Case 1, the liquid may be subsequently vaporized in the 21.80 vaporizer (Figure 2) and transferred to the turbine header to be used as peak shaving fuel. Alternatively, this liquid (containing 62 wt% DME) may be withdrawn from storage and sold as a crude product.

4.4 Case 4:

The mode of operation is similar to Case 3 in that a significant amount of DME is produced in addition to MeOH. However, Case 4 is based on 1:1 recycle and, as a consequence, requires a more complex Liquid Recovery section. Because reactor recycle is used, the DME content in the syngas-exit-reactor is lower in case 4 than in case 3. As a result, the recovery of DME is more difficult and greater levels of methanol-absorbent circulation (and refrigeration) is required. The number of unit operations is also greater as discussed below.

Figure 4.4 depicts the Liquids Recovery section for Case 4. The MeOH/DME bearing reactor effluent is chilled in the 21.31 Methanol Condenser and passed into the 22.10 HP Methanol Separator. From the 22.10, MeOH and DME are recovered as liquid and unconverted syngas (plus DME) removed as vapor. This vapor is split and cooled, in the 21.32 Cross Exchanger and 21.33 Economizer, then recombined and passed into the bottom of the 07.65 Absorber Column. The 07.65 column is refluxed with methanol (additional reflux is provided in the 07.66

condenser by vaporizing propane). This allows the DME (and some CO₂) to be absorbed and recovered as bottoms. The vapor overhead from the 07.65 is warmed in the 21.32 exchanger. A portion of the vapor flow is returned to the recycle compressor (Figure 2); the remainder is expanded to 270 psig in the 10.10 Purge Gas Expander.

The liquid from the 07.65 is warmed in the 07.74 and 21.73 exchangers then combined and sent to the 21.12 separator where dissolved, light components are degassed. The liquid from the 22.12 is further warmed in the 21.72 exchanger then directed to the 07.71 Fractionation Column. Within the 07.71 column, CO₂, DME, and some MeOH are rejected as a mixed vapor/liquid overhead product and purified methanol is withdrawn as bottoms. The condenser duty is provided in the 07.72 exchanger with cooling water; the reboiler duty is provided in the 07.73 exchanger with steam. The liquid from the bottom of the 07.71 is cooled in the 07.77 Reboiler and 21.72 Recycle Chiller, pumped in the 10.62, further cooled in the 21.73 exchanger, then returned to the 07.65 column as reflux.

The vapor overhead product from the 07.71 is chilled in the 07.74 exchanger and fed to a third column, the 07.75. The liquid overhead product from the 07.71 is combined with other liquid streams and fed to the 07.75 as well. Within the 07.75 Fractionator, CO₂ is rejected as vapor and the mixed DME/MeOH product is withdrawn as bottoms liquid. The condenser duty is provided in the 07.78 exchanger with vaporizing propane; the reboiler duty is provided in the 07.72 exchanger via cross exchange with the circulating methanol stream.

The vapor overhead product from the 07.74 is combined with the main gas flow from the 10.10 expander, then warmed in the 21.33 exchanger and finally sent to the CGCC turbine header. The liquid product is chilled in the 21.71 Product Cooler then sent to Storage (Figure 2).

As in Case 1, the liquid may be subsequently vaporized in the 21.80 vaporizer (Figure 2) and transferred to the turbine header to be used as peak shaving fuel. Alternatively, this liquid (containing 58 wt% DME) may be withdrawn from storage and sold as a crude product.

Relative to case 3, the case 4 recovery section is indeed more complex. Recall that case 3 utilizes a modest degree of methanol absorbent circulation to aid in DME recovery. In contrast, the absorbent circulation rate in case 4 is fairly high. As a result, it is no longer practical (economical) to strip the DME and CO₂ from the methanol in a column with a refrigerated condenser (as is used in case 3). Therefore, an additional distillation column has been added: the main column, 07.71, separates the DME and CO₂ from the circulating methanol (using cooling water for the condenser); the second column, 07.75, rejects CO₂ (using propane refrigeration to provide the condenser duty). Another important difference between case 3 and case 4 is that case 4 requires a propane refrigeration-supplied condenser on the 07.65 absorber column (this refrigeration is needed to chill the overhead). Clearly, case 4 utilizes refrigeration to a much more significant degree than case 3 and hence is both more complex and more costly. The benefit of case 4 over case 3 is that the reactor productivity is higher and therefore the flow through the front-end compression and pretreatment is reduced to 2/3 of case 3.

4.5 Case 5:

This mode of operation is the simplest of the All-DME plant schemes. No syngas recycle is utilized and the DME product is withdrawn as a vapor along with the unreacted syngas. All

that needs to be accomplished with the Liquids Recovery section is to recover MeOH so that it may be recycle to the reactor.

Figure 4.5 represents the Liquids Recovery section for Case 5. The MeOH/DME bearing reactor effluent is chilled in the 21.31 Methanol Condenser and 21.33 exchanger then passed into the 22.10 HP Methanol Separator. From the 22.10, methanol is removed as a liquid and unconverted syngas (plus DME) removed as a vapor. This vapor is expanded to 270 psig in the 10.10 Purge Gas Expander, then introduced to the 22.12 separator where the last remnants of methanol drop-out as liquid. The vapor off the 22.12 is rewarmed in the 21.33 exchanger. The liquid from the 22.12 separator is let-down to 270 psig, combined with the liquid from the 22.10 separator, then warmed in the 07.72 exchanger before being introduced to the 07.71 column. Within the 07.71 column, CO₂ and DME are rejected as a vapor overhead product and purified methanol is withdrawn as bottoms. The condenser duty is provided in the 07.72 exchanger via cross exchange; the reboiler duty is provided in the 07.73 exchanger with steam. The liquid bottoms product is pumped to pressure in the 10.62 then recycled to the reactor as supplemental feed (Figure 3). The vapor overheads are combined with the main syngas flow from the 21.33 and exit the plant for further processing.

4.6 Case 6:

This mode of operation is similar to Case 5 except the DME product is withdrawn as a vapor along with CO₂ (separated from the syngas stream).

Figure 4.6 represents the Liquids Recovery section for Case 6. Note that it is very similar to the flowsheet for Case 3. The major exception being that internal circulation of methanol is not utilized - thereby simplifying the design (and operation). Also note that the vapor overhead from the 07.71 constitutes the mixed DME/CO₂ product. The 07.71 bottoms are returned to the reactor as in Case 5.

4.7 Case 7:

This mode of operation is similar to Case 6 except the DME product is withdrawn as a liquid, purified product from a second column.

Figure 4.7 represents the Liquids Recovery section for Case 7. Note the similarity to the flowsheet for Case 6. However, note that the vapor overhead from the 07.71 constitutes a CO₂ purge to the gas turbine; the 07.71 bottoms contain MeOH and DME. This bottoms stream is passed to the 07.75 column to affect a DME-MeOH separation. The DME is withdrawn from the top of the 07.75 column, condensed in the 07.74 exchanger and sent to storage. The 07.75 bottoms contain the MeOH and as such are returned to the reactor as in Case 6.

5. PROCESS SPECIFICATIONS

The process data required to carry-out the economic evaluation were developed and are attached:

- Overall material balances are included in Table 2.
- An equipment list is attached as Table 3.
- Budgetary-quality equipment specifications are contained in Table 4.
- The utility summary is presented in Table 5.

The information contained in the above tables were utilized to estimate the cost of producing fuels from each of the seven operating options. The methodology used and estimation of cost is presented next.

6. COST ESTIMATES

6.1 Production Cost Estimate Basis

The capital costs are based on instantaneous 1992 Construction at a Gulf Coast location. These costs represent a battery limits erected facility added on to a large CGCC power plant.

The direct capital costs were broken-down by unit as follows:

- **Feed Compression and Gas Pretreatment**
Feed Compressor including Driver and Coolers
Bulk Acid Gas Removal, Absorber, Stripper, Pumps, Heat Exchangers
Refrigeration Unit
- **Basic Synthesis Section**
Recycle Compressor
Purge Gas Expander
Carbonyl and Sulfur Guard Beds
Reactor
Product Condenser and HP Separator
Stabilizer
Process Pumps
- **Extra Synthesis Equipment**
Equipment added to recover/separate DME
- **Slurry Reduction**
Heated Reduction Vessel
Catalyst Feed Equipment
- **Spent Catalyst Recovery**
Slurry Centrifuge
Oil Storage
- **Storage Loading and Vaporization**

- General Utilities
 - Steam Drum
 - Slurry Tank
 - Flare Knockout Drum

The production costs were computed using the following factors:

Syngas	\$ 4.00 per MM BTU, HHV
Steam Credit	\$ 3.15 per M lb
Electric Power	\$ 0.05 per kW-hr
Cooling Water	\$ 0.10 per M gal
Make-up Oil	\$ 2.87 per gal
Catalyst	\$10.09 per lb
Catalyst/Oil Disposal	\$ 1.00 per lb
Adsorbents	\$ 0.98 per lb carbon, \$ 2.56 per lb ZnO
Adsorbent Disposal	\$ 1.00 per lb
Operating Labor	10 persons @ \$50,000/yr
Plant Overhead	2 times Operating Labor
Maintenance	2% of Capital Investment
Taxes and Insurance	2% of Capital Investment

A Capital Charge Rate of 14% of investment was used which includes a ROI as well as depreciation and interest charges. This is typical for 'Utility Economics'.

6.2 Production Cost Comparison

Table 6 summarizes the unit cost of production for Cases 1-7. The Capital Cost break-down by section is contained in the upper portion of Table 6; the Unit Costs are presented in the lower portion. Conclusions drawn from these results are discussed below:

- **DME Coproduction for Use as a Turbine Fuel for Power Plant Peaking:**

The relevant options to consider for peak shaving are Cases 1-4. For these cases the Capital investment lies between \$ 63.7 MM (Case 3) and \$70.8 MM (Case 4). The appropriate unit costs range from \$8.76/MM BTU, LHV (Case 3) to \$9.10/MM BTU, LHV (Case 4).

The superiority of Case 3 (DME, no syngas recycle) over Case 4 (DME, 1:1 syngas recycle) demonstrates that the preferred method of coproducing DME is without the use of recycle. This can be explained as being a consequence of the relative difficulty in condensing the volatile DME from the syngas vapor.

Nevertheless, DME coproduction can reduce product cost as demonstrated by the favorable economics of Case 3 over either Case 1 or 2. Specifically, the conclusion is that one could produce and store a high pressure DME (61 wt%)/MeOH mixture at 25¢/MM BTU cheaper than it would be to store neat methanol. This represents a 2.8%

reduction in fuel cost. It could also be represented as an increase in gross ROI from 14% to 16.4%.

As it turned out, coproduction of DME carries only a very slight savings in capital, the major savings comes from lower utility requirements. Since the Case 3 scheme is more complex than the LPMEOH case, and since it introduced a 100 psi liquid storage requirement, it may be difficult to justify further development of DME/MeOH mixtures as a turbine fuel; however, the results do show that the mix can be made at a lower cost than methanol alone. Therefore in cases where mixtures of DME/MeOH would have been derived from a separate dehydration of methanol step (e.g. the MTG process), this process should be economically attractive. Furthermore, DME coproduction plants can operate with lower syngas throughputs than LPMEOH plants (for equivalent production). In applications where the desired production rates would force an LPMEOH plant into parallel pretreatment units, DME coproduction plants would clearly be favored.

- **DME as a Cold Start Additive to Methanol:**

Cases 1 and 2 can be modified slightly to produce the new cases 1A and 2A for the transportation fuel application. To accomplish this, the storage pump and vaporizer are removed and the utilities adjusted to eliminate the steam required to vaporize the product. This reduces the cost of each case by about 28 cents/MM BTU, LHV.

A comparison of Case 2A and Case 1A indicates that at a 14% ROI, the DME mix yields a lower required selling price (\$8.65/MM BTU) than that of the neat methanol value of (\$8.73/ MM BTU). In addition the DME will have a value as an additive. One might use MTBE as a guide to determine an additive value.

If MTBE sells at 1.00\$/gal when unleaded gasoline sells for 0.62\$/gal, this would represent a premium of about 90% over its fuel value (ie.MTBE's value as an additive is approximately 1.9 times its value as a fuel.).

If the DME in the product were valued at 1.9 times fuel value, the mix would be valued at approximately 1.4% higher than fuel value. If one uses \$8.73/MM BTU for the basis, then the mix would have a value of $1.014 \times \$8.73/\text{MM BTU} = \$8.85/\text{MM BTU}$. The difference between 8.85 and 8.65 represents a 20¢/MM BTU advantage for the mix; this would result in an increase in the ROI from 14% to 15.9%.

The conclusion is that there would be economic incentive to produce such a mixture. Furthermore, there would be little technical risk in such a process.

- **DME Production for Chemical Feedstock:**

Depending on the downstream application, it may be desirable to produce DME in any one of the three forms represented by Cases 5-7. The cost differential between Case 5 and Case 6 reflects the cost of separating DME from the unreacted syngas; the differential between Case 6 and Case 7 reflects the cost of recovering DME as a purified liquid.

It is also interesting to compare the cost differential between Case 1 and Case 7. Both these cases produce relatively pure liquid products (Case 1 - MeOH, Case 7 - DME). On a weight basis, the cost of methanol from Case 1 is \$0.083/lb. If this methanol could be dehydrated to DME at zero cost, the cost of that DME would be \$0.115/lb DME.

An estimate of the cost of methanol dehydration is approximately \$0.017/lb DME (1992 basis). This processing cost was estimated from an SRI study of a similar operation (PEP 111) where methanol was dehydrated to produce ethylene. Combining the methanol dehydration cost with the Case 1 MeOH production cost yields a DME-from-methanol cost of $0.115 + 0.017 = \$0.132/\text{lb DME}$. Therefore, the estimated cost to produce DME from Case 7 (only \$0.118/lb) represents a significant savings.

FIGURE 1.1: BLOCK DIAGRAM FOR CASE 1

ISSUED FOR ESTIMATE ONLY

	FEED GAS	TURBINE GAS	LIQUID PRODUCT	MED PRES STEAM
STREAM #	10	163	210	423
FLOW (LBMOL/HR)	17684	12151	2138	4004
MOL. WT.	20.49	25.72	32.65	18.02
PRESSURE (PSIG)	260	260	0	230
HHV (MMBTU/HR)	1957	1143	717	~
LHV (MMBTU/HR)	1861	1100	635	~

CASE 1 (OPTION A)

AIR PRODUCTS
GENERIC LP PLANT METHANOL PRODUCTION BLOCK DIAGRAM	
DRG. NO. GEN-CC1A	REV. 1
ATT-1016	REV. 1

DMH-8/4/92

FIGURE 1.2: BLOCK DIAGRAM FOR CASE 2

ISSUED FOR ESTIMATE ONLY

	FEED GAS	TURBINE GAS	LIQUID PRODUCT	MED PRES STEAM
STREAM #	10	163	210	423
FLOW (LBMOL/HR)	17684	12151	2237	3448
MOL. WT.	20.49	25.91	32.95	18.02
PRESSURE (PSIG)	260	260	0	230
HHV (MMBTU/HR)	1957	1110	762	~
LHV (MMBTU/HR)	1861	1070	676	~

DMH-8/4/92

CASE 2 (OPTION C)

AIR PRODUCTS
GENERIC LP PLANT MeOH W/LOW DME BLOCK DIAGRAM	
DRG. NO. GEN-CC1C	SHEET 1
ATT-1016	REV. 0

FIGURE 1.3: BLOCK DIAGRAM FOR CASE 3

ISSUED FOR ESTIMATE ONLY

	FEED GAS	TURBINE GAS	LIQUID PRODUCT	MED PRES STEAM
STREAM #	10	163	210	423
FLOW (LBMOL/HR)	15762	10739	1455	4753
MOL. WT.	20.49	26.16	39.44	18.02
PRESSURE (PSIG)	260	260	95	230
HHV (MMBTU/HR)	1744	960	684	~
LHV (MMBTU/HR)	1659	923	618	~

CASE 3 (OPTION L)

<p align="center">AIR PRODUCTS</p>	
<p align="center">GENERIC LP PLANT MeOH/DME COPRODUCT BLOCK DIAGRAM</p>	
DES. NO.	GEN-CC1L
REV.	0

DMH-8/4/92

FIGURE 1.4: BLOCK DIAGRAM FOR CASE 4

ISSUED FOR ESTIMATE ONLY

	FEED GAS	TURBINE GAS	LIQUID PRODUCT	MED PRES STEAM
STREAM #	10	163	210	423
FLOW (LBMOL/HR)	10600	5960	1492	3892
MOL. WT.	20.49	30.51	39.09	18.02
PRESSURE (PSIG)	260	260	98	230
HHV (MMBTU/HR)	1173	388	687	~
LHV (MMBTU/HR)	1115	372	620	~

CASE 4 (OPTION P)

AIR PRODUCTS
GENERIC LP PLANT MeOH/DME COPRODUCT BLOCK DIAGRAM	
DRG NO. GEN-CC1P	REV. 1
ATT-1016	REV. 1

DMH-8/4/92

FIGURE 1.5: BLOCK DIAGRAM FOR CASE 5

ISSUED FOR ESTIMATE ONLY

	FEED GAS	S.G. FRACTION	DME FRACTION	MED PRES STEAM
STREAM #	10	163	163	423
FLOW (LBMOL/HR)	12398	7252	1125	5877
MOL. WT.	20.49	29.16	46.07	18.02
PRESSURE (PSIG)	260	260	260	230
HHV (MMBTU/HR)	1372	551	709	~
LHV (MMBTU/HR)	1305	535	645	~

CASE 5 (OPTION Q)

AIR PRODUCTS
GENERIC LP PLANT DME ONLY, W/SG BLOCK DIAGRAM	
DRG. NO.	GEN-CC1Q
REV. NO.	1
ATT-1016	0

DMH-10/10/92

FIGURE 1.6: BLOCK DIAGRAM FOR CASE 6

42

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	FEED GAS	TURBINE GAS	LIQUID PRODUCT	MED PRES STEAM
STREAM /	10	139	160	423
FLOW (LBMOL/HR)	12398	6377	1996	5413
MOL. WT.	20.49	27.52	43.91	18.02
PRESSURE (PSIG)	260	260	260	230
HHV (MMBTU/HR)	1372	571	688	~
LHV (MMBTU/HR)	1305	552	628	~

CASE 6 (OPTION R)

AIR PRODUCTS
GENERIC LP PLANT DME ONLY, W/CO2 BLOCK DIAGRAM	
DRG. NO. GEN-CC1R	REV. 1
ATT-1016	REV. 0

DMH-10/10/92

FIGURE 1.7: BLOCK DIAGRAM FOR CASE 7

ISSUED FOR ESTIMATE ONLY

	FEED GAS	TURBINE GAS	LIQUID PRODUCT	MED PRES STEAM
STREAM #	10	163	180	423
FLOW (LBMOL/HR)	12398	7300	1072	5296
MOL. WT.	20.49	29.29	45.96	18.02
PRESSURE (PSIG)	260	260	118	230
HHV (MMBTU/HR)	1372	597	661	~
LHV (MMBTU/HR)	1305	578	602	~

CASE 7 (OPTION S)

AIR PRODUCTS
GENERIC LP PLANT LIQUID DME PRODUCT BLOCK DIAGRAM	
DRG. NO. GEN-CC1S	DEPT NO. 1
ATT-1016	REV. NO. 0

DMH-10/10/92

FIGURE 2: FEED/RECYCLE COMPRESSION, SULFUR/CARBONYL REMOVAL, LIQUID STORAGE (ALL CASES)

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DAH-7/10/92

AIR PRODUCTS		
GENERIC LP PLANT COMPRESSION AND STORAGE		
DRG. NO.	GEN-CC02	SHEET NO. 2
	ATT-1012	REV. NO. 1

FIGURE 3: SYNTHESIS SECTION (ALL CASES)

AIR PRODUCTS

**GENERIC LP PLANT
SYNTHESIS SECTION**

PAGE NO GEN-CC03

ATT-1012

ISSUED FOR ESTIMATE ONLY

044-8/17/02

FIGURE 4.1: LIQUID RECOVERY SECTION FOR CASE 1

AIR PRODUCTS	GENERIC LP PLANT	
	METHANOL PRODUCTION	
	LIQUIDS RECOVERY	
	GEN-CC4A	ATT-1012

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DATE: 8/17/92

FIGURE 4.2: LIQUID RECOVERY SECTION FOR CASE 2

PRODUCTS AIR GENERIC LP PLANT MeOH W/LOW DME LIQUIDS RECOVERY	
GEN-CC4C	REV NO 4
ATT-1012	REVISION 0

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DSH-5/17/93

FIGURE 4.3: LIQUID RECOVERY SECTION FOR CASE 3

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DUH-7/10/92

AIR PRODUCTS
GENERIC LP PLANT	
MeOH/DME COPRODUCT	
LIQUIDS RECOVERY	
GEN-CC4L	SHEET NO 4
ATT-1012	REVISED 0

FIGURE 4.4: LIQUID RECOVERY SECTION FOR CASE 4

ISSUED FOR ESTIMATE ONLY

DUH-7/10/92

AIR PRODUCTS	
GENERIC LP PLANT	
MeOH/DME COPRODUCT	
LIQUIDS RECOVERY	
GEN-CC4P	SHEET NO 4
ATT-1012	REVISION 0

FIGURE 4.5: LIQUID RECOVERY SECTION FOR CASE 5

AIR PRODUCTS
GENERIC LP PLANT	
DME RECOVERED IN S.G.	
MeOH RECYCLED	
GEN-CC4Q	SHEET NO 4
ATT-1012	REVISION 0

DWH-8/8/92

ISSUED FOR ESTIMATE ONLY

FIGURE 5: CATALYST PREPARATION AND ACTIVATION (ALL CASES)

AIR PRODUCTS	
PRODUCTS	
GENERIC LP PLANT	
SLURRY REDUCTION	
SECTION	
DRG NO.	GEN-CC05
REV NO.	5
ATT-1012	

ISSUED FOR ESTIMATE ONLY

DATE: 8/17/92

FIGURE 6: CATALYST AND OIL SEPARATION (ALL CASES)

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DATE: 8/17/92

AIR PRODUCTS
GENERIC LP PLANT	
CATALYST RECOVERY	
SECTION	
DRG. NO.	GEN-CC06
REV. NO.	6
ATT-1012	0

FIGURE 7: GENERAL UTILITIES -- PROPANE REFRIGERATION, INCINERATOR, STEAM SUPPLY (ALL CASES)

AIR PRODUCTS	
PRODUCTS LP PLANT	
GENERAL UTILITIES	
SECTION	
DRAWING NO.	GEN-CC07
REV.	7
DATE	7/10/82
ATTACHMENT	ATT-1012

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DATE: 7/10/82

TABLE 1A : SUMMARY OF CASE STUDIES

Case	Process Objective	Special Features	Reactor Recycle Ratio	Relative Cleanup & Compression Size	Relative Reactor Train Size	Relative Reactor Volume	Ranking of Product Recovery Complexity	Relative Capital Cost (\$)	Relative Unit Cost (\$/BTU LHV)
1	Produce methanol-only	Base-case plant with methanol stabilizer column.	1:1	1.00	1.00	1.00	1	1.00	1.00
2	Methanol with a small amount of DME to increase vapor pressure.	Partial condensation process for DME recovery, CO ₂ degasser, gas expander for refrigeration.	1:1	1.00	1.00	1.00	4	1.04	0.99
3	DME & Methanol coproduction to minimize production cost of turbine fuel	Partial condensation with absorber fluid assist (using methanol). Refrigeration by expander with propane	0	0.89	0.45	0.88	5	0.99	0.97
4	DME & Methanol coproduction to minimize production cost of turbine fuel	Absorber fluid circulation process. Refrigeration by expander with propane assist.	1:1	0.60	0.60	1.19	7	1.10	1.01
5	Produce DME-only. DME product taken as vapor with unreacted syngas	Partial condensation for methanol knock-out. Methanol recycle to reactor. Expander refrigeration.	0	0.70	0.35	1.59	2	0.83	0.93
6	Produce DME-only. DME product taken as vapor with CO ₂ impurity.	Partial condensation for DME recovery. DME/CO ₂ recovery column. Expander refrigeration. Methanol	0	0.70	0.35	1.59	3	0.87	0.99
7	Produce DME-only. DME product taken as purified liquid.	Like case 6 except staged distillation columns for DME recovery and liquefaction. Methanol recycle to reactor.	0	0.70	0.35	1.59	6	1.05	1.05

ISSUED FOR ESTIMATE

TABLE 1B: SUMMARY OF PROCESS CONDITIONS FOR CASES 1-7

	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6	Case 7
Reactor Feed Flows							
Plant Feed Flow (SCFD)	164	164	146	98	115	115	115
Syngas Recycle Rate (SCFD)	164	164	0	98	0	0	0
Water Injection Rate (gph)	3,053	3,053	1,785	2,704	1,075	1,057	1,057
MeOH Recycle Rate (gph)	0	0	0	0	564	578	581
Reactor							
Diameter (ft)	13.6	13.6	11.7	13.6	13.2	13.2	13.2
Height (ft)	76.1	76.1	90.7	90.6	120.8	120.8	120.8
Temperature (F)	482	482	482	482	482	482	482
Pressure (psig)	1060	1060	1060	1060	1060	1060	1060
Gas Space Velocity (sL/kg-hr)	6250	6250	2850	2950	1320	1320	1320
Alumina Loading (wt%)	0	0.3	7.4	7.4	13	13	13
Catalyst Usage Rate (lb/yr)	151110	151110	149533	195523	383321	383321	383321
Performance							
MeOH Production (TPD)	828	861	418	403	nil	nil	nil
DME Production (TPD)	nil	16	240	262	622	590	579
Product LHV (MMBTU/hr)	635	676	618	620	645	628	602
Product Composition (mole%)							
Hydrogen	0.00	0.00	0.00	0.00	10.05	0.34	0.00
Carbon Monoxide	0.00	0.00	0.00	0.00	43.77	5.64	0.00
Carbon Dioxide	0.12	0.03	2.30	2.60	22.53	39.50	1.80
Nitrogen	0.00	0.00	0.00	0.00	10.08	0.77	0.00
Dimethyl Ether	0.01	1.31	52.02	48.90	13.43	53.50	97.79
Methyl Formate	1.00	1.34	0.53	0.55	0.01	0.01	0.02
Water	1.16	0.82	3.65	3.81	0.03	0.09	0.12
Methanol	95.96	94.81	40.82	43.39	0.11	0.14	0.27
Ethanol	1.07	1.04	0.42	0.45	0.00	0.00	0.00
Propanol	0.33	0.31	0.13	0.14	0.00	0.00	0.00
Butanol	0.33	0.31	0.13	0.14	0.00	0.00	0.00
Utilities							
Steam Production (lb/hr at 230 psig)	25858	14044	63247	46732	105900	97586	85092
Cooling Water Flow (gph)	664500	809460	472500	589560	254640	326640	432060
Electric Power (MW)	14.60	16.16	14.23	13.21	11.77	12.54	13.53

ISSUED FOR ESTIMATE

TABLE 2.1: CASE 1

GENERIC LP PLANT - MATERIAL BALANCE
LPM - MV FUEL AND PS FUEL

Stream	Feed Gas	Water Injection	Turbine Gas	Purge Gas	Liquid Product
F/S Number	10	432	163	212	210
Pressure (psia)	275	800	275	15	15
Temperature (°F)	110	240	96	100	100
Flow (lbmol/hr)	17684	1415	12151	151	2138
Molecular Weight	20.51	18.02	25.72	40.35	32.65
Composition (mol%)					
Hydrogen	30.01	0	17.77	0.38	0.000
Carbon Monoxide	61.02	0	57.26	5.52	0.000
Carbon Dioxide	2.00	0	14.23	67.42	0.120
Nitrogen	6.81	0	9.90	0.83	0.000
Dimethyl Ether	0	0	0.020	0.390	0.009
Methyl Formate	0	0	0.055	1.95	0.996
Water	0.16	100	0.003	0.076	1.159
Methanol	0	0	0.760	23.30	95.96
Ethanol	0	0	0.006	0.116	1.072
Propanol	0	0	0.001	0.011	0.329
Butanol	0	0	0.000	0.004	0.332
COS	0.002	0	0	0	0
H2S	5E-04	0	0	0	0
Mineral Oil	0	0	0	0	0.03
HHV (MMBtu/hr)	1,957	0	1,143	14	717
LHV (MMBtu/hr)	1,861	0	1,100	13	635

ISSUED FOR ESTIMATE

TABLE 2.2: CASE 2

**GENERIC LP PLANT - MATERIAL BALANCE
LPDME - MV FUEL WITH COLD START**

Stream	Feed <u>Gas</u>	Water <u>Injection</u>	Turbine <u>Gas</u>	Purge <u>Gas</u>	Liquid <u>Product</u>
F/S Number	10	432	163	212	210
Pressure (psia)	275	800	275		15
Temperature (°F)	110	240	15		100
Flow (lbmol/hr)	17684	1415	12151	0	2237
Molecular Weight	20.51	18.02	25.91		32.95
Composition (mol%)					
Hydrogen	30.01	0	17.73		0.000
Carbon Monoxide	61.02	0	56.71		0.000
Carbon Dioxide	2.00	0	15.49		0.032
Nitrogen	6.81	0	9.91		0.000
Dimethyl Ether	0	0	0.139		1.310
Methyl Formate	0	0	0.003		1.336
Water	0.16	100	0.000		0.818
Methanol	0	0	0.019		94.81
Ethanol	0	0	0.000		1.038
Propanol	0	0	0.000		0.312
Butanol	0	0	0.000		0.312
COS	0.002	0	0		0
H2S	5E-04	0	0		0
Mineral Oil	0	0	0		0.03
HHV (MMBtu/hr)	1,957	0	1,110		762
LHV (MMBtu/hr)	1,861	0	1,070		676

ISSUED FOR ESTIMATE

TABLE 2.3: CASE 3

**GENERIC LP PLANT - MATERIAL BALANCE
LPDME - PS FUEL ONLY - HIGH REC - OPTION 1**

Stream	Feed <u>Gas</u>	Water <u>Injection</u>	Turbine <u>Gas</u>	Purge <u>Gas</u>	Liquid <u>Product</u>
F/S Number	10	432	163	212	210
Pressure (psia)	275	800	275		109
Temperature (°F)	110	240	74		100
Flow (lbmol/hr)	15762	827.5	10739	0	1455
Molecular Weight	20.51	18.02	26.16		39.44
Composition (mol%)					
Hydrogen	30.01	0	18.07		0.000
Carbon Monoxide	61.02	0	54.18		0.000
Carbon Dioxide	2.00	0	17.51		2.295
Nitrogen	6.81	0	10.00		0.000
Dimethyl Ether	0	0	0.244		52.024
Methyl Formate	0	0	0.003		0.525
Water	0.16	100	0.000		3.651
Methanol	0	0	0.008		40.82
Ethanol	0	0	0.000		0.421
Propanol	0	0	0.000		0.126
Butanol	0	0	0.000		0.126
COS	0.002	0	0		0
H2S	5E-04	0	0		0
Mineral Oil	0	0	0		0.02
HHV (MMBtu/hr)	1,744	0	960		684
LHV (MMBtu/hr)	1,659	0	923		618

ISSUED FOR ESTIMATE

TABLE 2.4: CASE 4

GENERIC LP PLANT - MATERIAL BALANCE
LPDME - PS FUEL ONLY - HIGH REC - OPTION 2

Stream	Feed <u>Gas</u>	Water <u>Injection</u>	Turbine <u>Gas</u>	Purge <u>Gas</u>	Liquid <u>Product</u>
F/S Number	10	432	163	212	210
Pressure (psia)	275	800	275		112
Temperature (°F)	110	240	83		100
Flow (lbmol/hr)	10600	1253	5960	0	1492
Molecular Weight	20.51	18.02	30.51		39.09
Composition (mol%)					
Hydrogen	30.01	0	12.98		0.000
Carbon Monoxide	61.02	0	38.21		0.000
Carbon Dioxide	2.00	0	36.23		2.604
Nitrogen	6.81	0	12.11		0.000
Dimethyl Ether	0	0	0.436		48.896
Methyl Formate	0	0	0.007		0.552
Water	0.16	100	0.000		3.809
Methanol	0	0	0.030		43.39
Ethanol	0	0	0.000		0.450
Propanol	0	0	0.000		0.136
Butanol	0	0	0.000		0.136
COS	0.002	0	0		0
H2S	5E-04	0	0		0
Mineral Oil	0	0	0		0.03
HHV (MMBtu/hr)	1,173	0	388		687
LHV (MMBtu/hr)	1,115	0	372		620

ISSUED FOR ESTIMATE

TABLE 2.5: CASE 5

GENERIC LP PLANT - MATERIAL BALANCE
LPDME - DME AS VAPOR IN SG, MEOH RECYCLE

Stream	Feed <u>Gas</u>	Water <u>Injection</u>	Turbine <u>Gas</u>	Purge <u>Gas</u>	Product <u>Fraction</u>
F/S Number	10	432	163	212	163
Pressure (psia)	275	800	275		275
Temperature (°F)	110	240	120		120
Flow (lbmol/hr)	12398	498.4	7252	0	1125
Molecular Weight	20.51	18.02	29.16		46.07
Composition (mol%)					
Hydrogen	30.01	0	11.60		0.000
Carbon Monoxide	61.02	0	50.56		0.000
Carbon Dioxide	2.00	0	26.02		0.000
Nitrogen	6.81	0	11.64		0.000
Dimethyl Ether	0	0	0.00		100.0
Methyl Formate	0	0	0.01		0.000
Water	0.16	100	0.03		0.000
Methanol	0	0	0.13		0.00
Ethanol	0	0	0.00		0.000
Propanol	0	0	0.00		0.000
Butanol	0	0	0.00		0.000
COS	0.002	0	0		0
H2S	5E-04	0	0		0
Mineral Oil	0	0	0		0.00
HHV (MMBtu/hr)	1,372	0	551		709
LHV (MMBtu/hr)	1,305	0	535		645

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TABLE 2.6: CASE 6

GENERIC LP PLANT - MATERIAL BALANCE
LPDME - DME AS VAPOR W/CO₂, MEOH RECYCLE

Stream	Feed <u>Gas</u>	Water <u>Injection</u>	Turbine <u>Gas</u>	Purge <u>Gas</u>	Vapor <u>Product</u>
F/S Number	10	432	139	212	160
Pressure (psia)	275	800	275		275
Temperature (°F)	110	240	120		115
Flow (lbmol/hr)	12398	489.7	6377	0	1996
Molecular Weight	20.51	18.02	27.52		43.91
Composition (mol%)					
Hydrogen	30.01	0	13.09		0.338
Carbon Monoxide	61.02	0	55.84		5.635
Carbon Dioxide	2.00	0	17.16		39.503
Nitrogen	6.81	0	13.00		0.771
Dimethyl Ether	0	0	0.91		53.5
Methyl Formate	0	0	0.00		0.014
Water	0.16	100	0.00		0.092
Methanol	0	0	0.00		0.14
Ethanol	0	0	0.000		0.000
Propanol	0	0	0.000		0.000
Butanol	0	0	0.000		0.000
COS	0.002	0	0		0
H ₂ S	5E-04	0	0		0
Mineral Oil	0	0	0		0.00
			100		
HHV (MMBtu/hr)	1,372	0	571		688
LHV (MMBtu/hr)	1,305	0	552		628

ISSUED FOR ESTIMATE

TABLE 2.7: CASE 7

GENERIC LP PLANT - MATERIAL BALANCE
LPDME - DME AS LIQUID, MEOH RECYCLE

Stream	Feed Gas	Water Injection	Turbine Gas	Purge Gas	Liquid Product
F/S Number	10	432	163	212	180
Pressure (psia)	275	800	275		133
Temperature (°F)	110	240	62		100
Flow (lbmol/hr)	12398	489.7	7300	0	1072
Molecular Weight	20.51	18.02	29.29		45.96
Composition (mol%)					
Hydrogen	30.01	0	11.52		0.000
Carbon Monoxide	61.02	0	50.32		0.000
Carbon Dioxide	2.00	0	25.53		1.795
Nitrogen	6.81	0	11.57		0.000
Dimethyl Ether	0	0	1.06		97.79
Methyl Formate	0	0	0.00		0.017
Water	0.16	100	0.00		0.120
Methanol	0	0	0.00		0.27
Ethanol	0	0	0.00		0.000
Propanol	0	0	0.00		0.000
Butanol	0	0	0.00		0.000
COS	0.002	0	0		0
H2S	5E-04	0	0		0
Mineral Oil	0	0	0		0.00
HHV (MMBtu/hr)	1,372	0	597		661
LHV (MMBtu/hr)	1,305	0	578		602

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**TABLE 3: GENERIC LP PLANT DESIGN
EQUIPMENT LIST**

ITEM NUMBER	F.S.	DESCRIPTION
01.10A	2	FEED GAS COMPRESSOR - FIRST STAGE
01.10B	2	FEED GAS COMPRESSOR - SECOND STAGE
01.20	2	RECYCLE COMPRESSOR
01.30	7	VENT STREAMS TO INCINERATOR COMPRESSOR
01.35	7	01.30 INLET KNOCKOUT POT
01.40	5	REDUCTION N2 COMPRESSOR
01.60	2	INSTRUMENT AIR COMPRESSOR
02.10A&B	2	CARBONYL GUARD BEDS
02.20	3	SULFUR GUARD BED
02.83	2	02.10 REGEN HEATER
07.65	4	ABSORBER COLUMN
07.66	4	07.65 COLUMN CONDENSER
07.71	4	FRACTIONATION COLUMN
07.72	4	07.71 COLUMN CONDENSER
07.73	4	07.71 COLUMN REBOILER
07.74	4	07.71 COLUMN OVERHEAD CHILLER
07.75	4	CO2 REJECTION COLUMN
07.76	4	07.75 COLUMN CONDENSER
07.77	4	07.75 COLUMN REBOILER
10.10	4	PURGE GAS EXPANDER
10.30	7	INCINERATOR LIQUID FEED PUMP
10.40	2	FUEL TO VAPORIZOR PUMP
10.50	3	CONDENSED OIL CIRCULATION PUMP
10.52	7	EMERGENCY SLURRY RETURN PUMP
10.53	5	UTILITY OIL PUMP
10.54	5	REDUCTION OIL PUMP
10.56	6	FRESH OIL PUMP
10.58	6	SLURRY FILTRATION CIRCULATION PUMP
10.59	5	REDUCED SLURRY TRANSFER PUMP
10.60	3	OIL MAKEUP PUMP
10.62 A&B	4	METHANOL TO STORAGE/RECYCLE PUMPS
10.64	3	H2O INJECTION PUMP
10.65	4	07.65 COLUMN REFLUX PUMP
10.70		OILY WATER SUMP PUMP
15.10	7	INCINERATOR
15.20	7	STEAM DRUM
15.40	5	UTILITY OIL HEATER

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Page 1

**TABLE 3: GENERIC LP PLANT DESIGN
EQUIPMENT LIST**

ITEM NUMBER	F.S.	DESCRIPTION
21.01	2	01.10 COMPRESSOR INTERCOOLER
21.02	2	01.10 COMPRESSOR RECYCLE CHILLER
21.03	2	01.20 COMPRESSOR RECYCLE CHILLER
21.11	3	REACTOR FEED/PRODUCT DEPHLEGMATOR
21.20	3	REACTOR INTERNAL HEAT EXCHANGER
21.31	4	METHANOL PRODUCT CONDENSER
21.32	4	RECOVERY SECTION CROSS EXCHANGER
21.33	4	RECOVERY SECTION ECONOMIZER
21.40	5	REDUCTION FEED/PRODUCT ECONOMIZER
21.41	5	REDUCTION GAS COMPRESSOR COOLERS
21.53	5	UTILITY OIL AIRCOOLER
21.60	2	INSTRUMENT AIR COMPRESSOR COOLERS
21.61	7	01.30 RECYCLE COOLER
21.71	4	LIQUID PRODUCT COOLER
21.72	4	LIQUID RECYCLE COOLER
21.73	4	LIQUID RECYCLE CROSS EXCHANGER
21.80	2	LIQUID FUEL VAPORIZOR
22.02	2	FEED GAS COMPRESSOR SECOND STAGE KNOCKOUT DRUM
22.10	4	H.P. METHANOL SEPARATOR
22.11	4	H.P. LIQUIDS SEPARATOR
22.12	4	M.P. FLASH POT/ABSORBER
22.13	4	L.P. FLASH POT
22.40	5	REDUCTION CONDENSATE ACCUMULATOR
22.50	6	RECLAIMED OIL FILTER
22.51	4	LIQUID PRODUCT/RECYCLE FILTER
22.52	2	TURBINE FUEL FILTER
22.53	2	02.10 GUARD BED REGEN TO TURBINE FILTER
22.54	2	02.10 GUARD BED REGEN HEATER OUTLET FILTER
22.90		OILY/WATER SEPARATOR
27.10	3	LPMEOH REACTOR
28.10	6	OIL STORAGE TANK
28.30	5	SLURRY MIX TANK & REDUCTION VESSEL
28.31	5	SLURRY MIX TANK AGITATOR
28.35	5	CATALYST BIN, HOPPER & FEEDER
28.40	2	LIQUID FUEL STORAGE
28.45	6	SLURRY CENTRIFUGE
28.46	6	SLURRY FILTRATION SURGE POT
28.50	7	FLARE KNOCKOUT POT
28.53	5	UTILITY OIL EXPANSION TANK

**TABLE 3: GENERIC LP PLANT DESIGN
EQUIPMENT LIST**

ITEM NUMBER	F.S.	DESCRIPTION
28.60	7	SLURRY TANK
28.61	7	SLURRY TANK AGITATOR
30.00	2	SULFUR REMOVAL UNIT
40.10		PROPANE COMPRESSOR
40.20		PROPANE STORAGE
40.21		PROPANE MP FLASH POT
40.30		PROPANE CONDENSER

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
01.10A FEED GAS COMPRESSOR - FIRST STAGE							
TYPE	CENTRIFUGAL	SAME AS	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
INLET P (PSIA)	275		275	275	275	275	275
OUTLET P (PSIA)	575		575	575	575	575	575
ADIABATIC EFFICIENCY (%)	75		75	75	75	75	75
GAS M.W. (LB/LB-MOLE)	20.51		20.51	20.51	20.51	20.51	20.51
COMPRESSION POWER (HP)	8741		7791	5240	6128	6128	6128
MOTOR POWER (HP)	9201.05	9201.05	8201.05	5515.79	6450.53	6450.53	6450.53
FLOW (#MOLE/HR)	17684	17684	15762	10600	12398	12398	12398
MATERIAL	CS		CS	CS	CS	CS	CS
01.10B FEED GAS COMPRESSOR - SECOND STAGE							
TYPE	CENTRIFUGAL	SAME AS	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
INLET P (PSIA)	570		570	570	570	570	570
OUTLET P (PSIA)	1200		1200	1200	1200	1200	1200
ADIABATIC EFFICIENCY (%)	75		75	75	75	75	75
GAS M.W. (LB/LB-MOLE)	20.51		20.51	20.51	20.51	20.51	20.51
COMPRESSION POWER (HP)	8837		7876	5297	6195	6195	6195
MOTOR POWER (HP)	9302.11	9302.11	8290.53	5575.79	6521.05	6521.05	6521.05
FLOW (#MOLE/HR)	17684	17684	15762	10600	12398	12398	12398
MATERIAL	CS		CS	CS	CS	CS	CS
01.20 RECYCLE COMPRESSOR							
TYPE	RECIPROCATING	RECIPROCATING	N/A	RECIPROCATING	N/A	N/A	N/A
INLET P (PSIA)	1035	1045		1025			
OUTLET P (PSIA)	1100	1100		1100			
TOTAL OUTLET FLOW (#MOLE/HR)	17684	17684		10600			
OUTLET M.W. (LB/LB-MOLE)	25.69	25.8		27.37			
ADIABATIC EFFICIENCY (%)	75	75		75			
COMPRESSION POWER (HP)	628	527		411			
MOTOR POWER (HP)	661.053	554.737	0	432.632	0	0	0
MATERIAL	CS	CS		CS			
01.30 INCINERATOR FEED COMPRESSOR							
TYPE	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING
INLET P (PSIA)	13.2	13.2	13.2	13.2	13.2	13.2	13.2
OUTLET P (PSIA)	23	23	23	23	23	23	23
TOTAL OUTLET FLOW (#MOLE/HR)	191	40	39.5633	51.7316	101.405	101.405	101.405
OUTLET M.W. (LB/LB-MOLE)	40.3	40.3	40.3	40.3	40.3	40.3	40.3
POLYTROPIC EFFICIENCY (%)	70	70	70	70	70	70	70
TOTAL POWER (HP)	74.3095	15.5709	15.4009	20.1481	39.4945	39.4045	39.4945
MATERIAL	CS	CS	CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
01.35 INCINERATOR COMP SUCTION KNOCKOUT POT							
TYPE	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL
VESSEL ID (FT)	1.86829	0.85498	0.8503	0.97231	1.36131	1.36131	1.36131
T-T HEIGHT (FT)	6	4.3	4.3	4.3	5	5	5
DESIGN T (F)	150	150	150	150	150	150	150
DESIGN P (PSIG)	50	50	50	50	50	50	50
VAPOR Kv	0.186	0.186	0.186	0.186	0.186	0.186	0.186
VESSEL MATERIAL	CS	CS	CS	CS	CS	CS	CS
01.40 REDUCTION N2 COMPRESSOR							
TYPE	RECIPROCATING	SAME AS	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING
INLET P (PSIA)	13.2		13.2	13.2	13.2	13.2	13.2
OUTLET P (PSIA)	145		145	145	145	145	145
ADIABATIC EFFICIENCY (%)	75		75	75	75	75	75
GAS M.W. (LB/LB-MOLE)	28.01		28.01	28.01	28.01	28.01	28.01
POWER (HP)	463.139	463.139	458.083	599.973	1174.11	1174.11	1174.11
FLOW (#MOLE/HR)	266.08	266.08	263.175	344.119	674.545	674.545	674.545
MATERIAL	CS		CS	CS	CS	CS	CS
01.60 INSTRUMENT AIR COMPRESSOR							
TYPE	CENT - 3 STG	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS
INLET P (PSIA)	13.2						
OUTLET P (PSIA)	114						
OUTLET FLOW (#MOLE/HR)	224.785	224.785	224.785	224.785	224.785	224.785	224.785
OUTLET FLOW (SCFH)	85300						
M.W. (LB/LB-MOLE)	28.01						
ADIABATIC EFFICIENCY (%)	70						
TOTAL POWER (HP)	376.982	376.982	376.982	376.982	376.982	376.982	376.982
MATERIAL	CS						
02.10 A CARBONYL GUARD BEDS							
TYPE	VERT PACKED	SAME AS	VERT PACKED	VERT PACKED	VERT PACKED	VERT PACKED	VERT PACKED
DIAMETER (FT)	3.3		3.3	3	3	3	3
HEIGHT T-T (FT)	13.8264		12.3237	10.0281	11.7291	11.7291	11.7291
BED VOLUME (FT^3)	118.257		105.404	70.8848	82.9084	82.9084	82.9084
SPACE VELOCITY (HR^-1)	638		638	638	638	638	638
DESIGN T (F)	200		200	200	200	200	200
DESIGN P (PSIG)	1320		1320	1320	1320	1320	1320
DESIGN FE(CO)5 LEVEL (PPB)	100		100	100	100	100	100
DESIGN NI(CO)4 LEVEL (PPB)	50		50	50	50	50	50
MATERIAL	CS		CS	CS	CS	CS	CS
ASSUMED ONSTREAM FACTOR (%)	80		80	80	80	80	80

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
02.20 SULFUR GUARD BED							
TYPE	VERT PACKED	SAME AS	VERT PACKED	VERT PACKED	VERT PACKED	VERT PACKED	VERT PACKED
NUMBER OF UNITS	1		1	1	1	1	1
DIAMETER (FT)	8		6.25	6.75	5.5	5.5	5.5
HEIGHT T-T (FT)	22.6217		16.5175	19.0468	16.7772	16.7772	16.7772
L/D	2.82771		2.64281	2.82175	3.0504	3.0504	3.0504
BED VOLUME (FT ³)	1137.09		506.752	681.584	398.598	398.598	398.598
SPACE VELOCITY (HRA ⁻¹)	228.488		228.488	228.488	456.976	456.976	456.976
DESIGN T (F)	600		600	600	600	600	600
DESIGN P (PSIG)	1320		1320	1320	1320	1320	1320
MATERIAL	SS CLAD ON CS		SS CLAD ON CS	SS CLAD ON CS	SS CLAD ON CS	SS CLAD ON CS	SS CLAD ON CS
DESIGN SULFUR LEVEL (PPM)	0.3		0.3	0.3	0.3	0.3	0.3
ASSUMED ONSTREAM FACTOR (%)	80		80	80	80	80	80
02.10 CARBONYL REGEN HEATER							
TYPE	ELECTRIC	SAME AS	ELECTRIC	ELECTRIC	ELECTRIC	ELECTRIC	ELECTRIC
HEATER KW	22.7418	22.7418	20.2701	13.6317	15.9439	15.9439	15.9439
REGENERATION TEMPERATURE (F)	250		250	250	250	250	250
TIME TO HEAT (HRS)	8		8	8	8	8	8
TIME ON HOLD (HRS)	10		10	10	10	10	10
SUP. VELOCITY IN 02.10 (FT/SEC)	0.25		0.25	0.25	0.25	0.25	0.25
REGEN FLOW WITH SYNGAS (LBMOLE/H)	21.3773		19.0539	12.8138	14.9873	14.9873	14.9873
07.65 ABSORBER COLUMN							
TYPE	N/A	N/A	TRAYED - 1 SECT	TRAYED - 1 SECT	N/A	N/A	N/A
TRAYED SECTION							
DIAMETER (IN)			80	86			
OVERALL HEIGHT (FT)			24	32			
ASSUMED EFFICIENCY (%)			50	50			
ACTUAL TRAY COUNT			12	16			
OVERALL HEIGHT (FT)			33	42			
DESIGN T (F)			-150	-30			
DESIGN P (PSIG)			350	1320			
MATERIAL			SS	SS			

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
07.65 07.65 COLUMN CONDENSER							
TYPE	N/A	N/A	N/A	KETTLE	N/A	N/A	N/A
DUTY (MMBTU/HR)				9.2			
U ASSUMED (BTU/HR-FT ² -F)				100			
LMTD (F)				28.8246			
AREA (FT ²) + 20% S.F.				3830.08			
SHELL DIAMETER (IN)				90			
SHELL LENGTH (FT)				20.1173			
TUBE LENGTH (FT)				18.1173			
NUMBER OF TUBES				646			
TUBE OD (IN)				1.25			
SHELL: DESIGN T (F)				-50			
DESIGN P (PSIG)				50			
MATERIAL				SS			
TUBES: DESIGN T (F)				-50			
DESIGN P (PSIG)				1320			
MATERIAL				SS			
07.71 FRACTIONATION COLUMN							
TYPE	N/A	TRAYED - 2 SECT	TRAYED - 2 SECT	TRAYED - 2 SECT	TRAYED - 2 SECT	TRAYED - 2 SECT	TRAYED - 2 SECT
TOP SECTION							
DIAMETER (IN)		18	28	36	26	52	43
OVERALL HEIGHT (FT)		21	21	21	25	29	29
ASSUMED EFFICIENCY (%)		50	50	50	50	50	50
ACTUAL TRAY COUNT		8	8	8	10	12	12
BOTTOM SECTION							
DIAMETER (IN)		54	50	62	34	52	46
OVERALL HEIGHT (FT)		28	58	52	28	24	24
ASSUMED EFFICIENCY (%)		50	50	50	50	50	50
ACTUAL TRAY COUNT		10	24	22	10	8	8
OVERALL HEIGHT (FT)		49	77	73	53	53	53
DESIGN T (F)		350	-20/350	350	350	350	-20/350
DESIGN P (PSIG)		350	350	350	350	350	350
MATERIAL		SS	CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
07.72 07.71 COLUMN CONDENSER							
TYPE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
DUTY (MMBTU/HR)	1.03	0.45	1.65	10	0.94	2.75	5.19
U ASSUMED (BTU/HR-FT ² -F)	30	50	75	50	75	50	75
LMTD (F)	55.8	22.4	38.9705	28.6	68	27.9988	35.3283
AREA (FT ²) + 20% S.F.	738.351	482.143	677.438	8391.61	221.176	2357.24	2350.52
SHELL DIAMETER (IN)	20	20	20	54	20	37	39
SHELL HEIGHT (FT)	29.2841	17.5925	22.6931	32.5662	10.7766	23.1507	21.195
TUBE HEIGHT (FT)	19.2841	12.5925	17.6931	25.5662	5.77663	16.1507	14.195
NUMBER OF TUBES	117	117	117	1003	117	446	506
TUBE OD (IN)	1.25	1.25	1.25	1.25	1.25	1.25	1.25
COOLING WATER FLOW (GPM)	99.9563	0	0	970.45	0	266.874	0
SHELL: DESIGN T (F)	250	-100	-100	250	250	350	-50
DESIGN P (PSIG)	100	350	350	50	350	50	50
MATERIAL	CS	SS	SS	CS	CS	CS	SS
TUBES: DESIGN T (F)	250	-100	-100	250	250	250	-20
DESIGN P (PSIG)	100	350	350	350	350	350	350
MATERIAL	CS	SS	SS	CS	CS	CS	SS
07.73 07.71 COLUMN REBOILER							
TYPE	KETTLE	KETTLE	KETTLE	KETTLE	KETTLE	KETTLE	KETTLE
DUTY (MMBTU/HR)	4.44	14.78	13.1	23.27	4.7	11.82	6.83
U ASSUMED (BTU/HR-FT ² -F)	65	65	65	65	65	65	35
LMTD (F)	241.472	72.4068	64.3952	59.3864	72.4068	70.4042	224.47
AREA (FT ²) + 20% S.F.	339.458	3768.45	3755.65	7233.98	1198.36	3099.47	1043.22
SHELL DIAMETER (IN)	42	90	90	108	62	90	62
SHELL LENGTH (FT)	10.7907	20.1062	20.0447	25.3673	14.9855	16.8919	13.3044
TUBE LENGTH (FT)	8.7907	18.1062	18.0447	23.3673	12.9855	14.8919	11.3044
NUMBER OF TUBES	118	636	636	946	282	636	282
TUBE OD (IN)	1.25	1.25	1.25	1.25	1.25	1.25	1.25
STEAM FLOW (LB/HR)	5222	17402	15421	27394	5534	13915	8042
SHELL DESIGN T (F)	250	350	350	350	350	350	350
DESIGN P (PSIG)	100	350	350	350	350	350	350
MATERIAL	CS	CS	CS	CS	CS	CS	CS
TUBES DESIGN T (F)	450	450	450	450	450	450	450
DESIGN P (PSIG)	350	350	350	350	350	350	350
MATERIAL	CS	CS	CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
07.74 07.75 COLUMN OVERHEAD CHILLER							
TYPE	N/A	N/A	N/A	SHELL & TUBE	N/A	N/A	SHELL & TUBE
DUTY (MMBTU/HR)				3.72			8.8
U ASSUMED (BTU/HR-FT ² -F)				75			75
LMTD (F)				23			60
AREA (FT ²) + 20% S.F.				2587.83			2346.67
SHELL DIAMETER (IN)				36			36
SHELL HEIGHT (FT)				18.9636			17.1964
TUBE HEIGHT (FT)				18.9636			17.1964
NUMBER OF TUBES				417			417
TUBE OD (IN)				1.25	0	0	1.25
COOLING WATER FLOW (GPM)							853.996
				250			250
SHELL DESIGN T (F)				350			50
DESIGN P (PSIG)				CS			CS
MATERIAL							
				250			250
TUBES DESIGN T (F)				350			350
DESIGN P (PSIG)				CS			CS
MATERIAL							
07.75 DME PURIFICATION COLUMN							
TYPE	N/A	N/A	N/A	TRAYED	N/A	N/A	TRAYED
TRAYED SECTION				44			50
DIAMETER (IN)				52			52
OVERALL HEIGHT (FT)				50			50
ASSUMED EFFICIENCY (%)				24			24
ACTUAL TRAY COUNT							
				62			62
OVERALL HEIGHT (FT)				-20/350			350
DESIGN T (F)				350			350
DESIGN P (PSIG)				SS			CS
MATERIAL							

73

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
07.76 07.75 COLUMN CONDENSER							
TYPE	N/A	N/A	N/A	SHELL & TUBE	N/A	N/A	SHELL & TUBE
DUTY (MMBTU/HR)				3.28			3.95
U ASSUMED (BTU/HR-FT ² -F)				75			75
LMTD (F)				28.4504			70
AREA (FT ²) + 20% S.F.				1844.62			902.857
SHELL DIAMETER (IN)				30			30
SHELL LENGTH (FT)				26.0326			15.2945
TUBE LENGTH (FT)				21.0328			10.2945
NUMBER OF TUBES				268			268
TUBE OD (IN)				1.25	0	0	1.25
COOLING WATER FLOW (GPM)							383.328
SHELL: DESIGN T (F)				-50			350
DESIGN P (PSIG)				50			50
MATERIAL				SS			CS
TUBES: DESIGN T (F)				-50			350
DESIGN P (PSIG)				350			350
MATERIAL				SS			CS
07.77 07.75 COLUMN REBOILER							
TYPE	N/A	N/A	N/A	KETTLE	N/A	N/A	KETTLE
DUTY (MMBTU/HR)				7.7			11.11
U ASSUMED (BTU/HR-FT ² -F)				65			65
LMTD (F)				105.954			65.3968
AREA (FT ²) + 20% S.F.				1341.66			3136.36
SHELL DIAMETER (IN)				54			90
SHELL LENGTH (FT)				22.499			17.0692
TUBE LENGTH (FT)				20.499			15.0692
NUMBER OF TUBES				200			636
TUBE OD (IN)				1.25	5534	13915	1.25
STEAM FLOW (LB/HR)							8042
SHELL: DESIGN T (F)				350			350
DESIGN P (PSIG)				350			350
MATERIAL				CS			CS
TUBES: DESIGN T (F)				350			350
DESIGN P (PSIG)				350			350
MATERIAL				CS			CS

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
10.10 PURGE GAS EXPANDER							
TYPE	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
INLET P (PSIA)	1025	1025	1035	1025	1045	1035	1035
OUTLET P (PSIA)	275	285	285	285	285	280	280
ADIABATIC EFFICIENCY (%)	75	75	75	75	75	75	75
GAS M.W. (LB/LB-MOLE)	25.67	25.4	26.5	27.4	31.32	28.55	28.55
POWER (HP)	5672	3414	3185	1512	2953	1953	1953
FLOW (#MOLE/HR)	12091	11791	10896	4720	8382	6789	6789
MATERIAL	CS	SS	SS	SS	SS	SS	SS
10.30 INCINERATOR LIQUID FEED PUMP							
TYPE	CENTRIFUGAL	SAME AS	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
INLET P (PSIA)	13.2		13.2	13.2	13.2	13.2	13.2
HEAD (FT)	49.8		49.8	49.8	49.8	49.8	49.8
EFFICIENCY (%)	50		50	50	50	50	50
FLOWRATE (GPM)	7		8,92358	9,05304	17,7458	17,7458	17,7458
BRAKE HORSEPOWER (BHP)	5	5	4,94542	6,46645	12,6756	12,6756	12,6756
MATERIAL	CS		CS	CS	CS	CS	CS
10.40 FUEL TO VAPORIZER PUMP							
75 TYPE	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	N/A	N/A	CENTRIFUGAL
FLOW (GPM)	712	752	629	635	0	0	613
HEAD (FT)	806	806	568	557			560
DENSITY (LB/FT ³)	48.7	48.7	45.4	45.7			40
VISCOSITY (CP)	0.4	0.4	0.4	0.4			0.4
POWER (BHP)	226.98	239.648	131.5	130.6	0	0	111
EFFICIENCY (%)	50	50	50	50			50
DESIGN T (F)	150	150	150	150			150
SUCTION P (PSIA)	15	15	109	111	0	0	133
MATERIAL	CS	CS	CS	CS			CS
10.50 COND. OIL CIRCULATION PUMP							
TYPE	CENTRIFUGAL	SAME AS	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
FLOW (GPM)	65,0415		28,9862	38,9066	22,7998	22,7998	45,5996
HEAD (FT)	156.1		156.1	156.1	156.1	156.1	156.1
DENSITY (LB/FT ³)	40		49	49	49	49	49
VISCOSITY (CP)	2.22		2.22	2.22	2.22	2.22	2.22
POWER (BHP)	100,064	100,064	44,5941	59,9794	35,0768	35,0766	70,1533
EFFICIENCY (%)	32.34		32.34	32.34	32.34	32.34	32.34
DESIGN T (F)	200		200	200	200	200	200
SUCTION P (PSIA)	1033		1033	1033	1033	1033	1033
MATERIAL	CS		CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
10.52	<u>EMERGENCY SLURRY RETURN PUMP</u>							
	TYPE	CENTRIFUGAL	SAME AS	SAME AS	SAME AS	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
	SLURRY TRANSFER TIME (HRS)	8		8	8	8	8	8
	VOL OF SLURRY TO BE TRANS (GAL)	35838		35446.8	46349	60716	60716	60716
	INLET P (PSIA)	14		14	14	14	14	14
	OUTLET P (PSIA)	40		40	40	40	40	40
	HEAD (FT)	78.5		78.5	78.5	78.5	78.5	78.5
	EFFICIENCY (%)	35.5		35.5	35.5	35.5	35.5	35.5
	FLOWRATE (GPM)	74.6825		73.8474	96.5603	126.492	126.492	126.492
	BRAKE HORSEPOWER (BHP)	14.9263	14.9263	14.7633	19.304	25.2878	25.2878	25.2878
	MATERIAL	CS		CS	CS	CS	CS	CS
10.53	<u>UTILITY OIL PUMP</u>							
	TYPE	CENTRIFUGAL	SAME AS	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
	INLET P (PSIA)	58		58	58	58	58	58
	OUTLET P (PSIA)	108		108	108	108	108	108
	HEAD (FT)	53		53	53	53	53	53
	EFFICIENCY (%)	50		50	50	50	50	50
	FLOWRATE (GPM)	600		593.45	775.975	1521.07	1521.07	1521.07
	BRAKE HORSEPOWER (BHP)	48	48	47.478	62.078	121.686	121.686	121.686
	MATERIAL	CS		CS	CS	CS	CS	CS
96	DESIGN T (F)	500		500	500	500	500	500
	DESIGN P (PSIG)	150		150	150	150	150	150
10.54	<u>REDUCTION OIL PUMP</u>							
	TYPE	CENTRIFUGAL	SAME AS	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
	INLET P (PSIA)	19		19	19	19	19	19
	OUTLET P (PSIA)	39		39	39	39	39	39
	HEAD (FT)	58.2		58.2	58.2	58.2	58.2	58.2
	EFFICIENCY (%)	36.6		36.6	36.6	36.6	36.6	36.6
	FLOWRATE (GPM)	83.6		82.6874	108.119	211.936	211.936	211.936
	BRAKE HORSEPOWER (BHP)	8	8	7.91267	10.3463	20.281	20.281	20.281
	MATERIAL	CS		CS	CS	CS	CS	CS
10.56	<u>FRESH OIL PUMP</u>							
	TYPE	RECIPROCATING	SAME AS	SAME AS	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING
	INLET P (PSIA)	13.2			13.2	13.2	13.2	13.2
	HEAD (FT)	87.4			87.4	87.4	87.4	87.4
	FLOWRATE (GPM)	52.2			67.5098	132.333	132.333	132.333
	POWER (HP)	4	4	4	5.17316	10.1405	10.1405	10.1405
	MATERIAL	CS			CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
10.58	<u>SLURRY FILTRATION CIRCULATION PUMP</u>							
	TYPE	CENTRIFUGAL	SAME AS	SAME AS	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
	INLET P (PSIA)	13.2			13.2	13.2	13.2	13.2
	HEAD (FT)	58.2			58.2	58.2	58.2	58.2
	FLOWRATE (GPM)	5.2			6.72511	13.1826	13.1826	13.1826
	BRAKE HORSEPOWER (BHP)	2	2	2	2.58658	5.07024	5.07024	5.07024
	MATERIAL	CS			CS	CS	CS	CS
10.59	<u>REDUCED SLURRY TRANSFER PUMP</u>							
	TYPE	RECIPROCATING	SAME AS	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING
	INLET P (PSIA)	14		14	14	14	14	14
	OUTLET P (PSIA)	1128		1128	1128	1128	1128	1128
	HEAD (FT)	3200		3200	3200	3200	3200	3200
	EFFICIENCY (%)	32		32	32	32	32	32
	FLOWRATE (GPM)	8		7.91267	10.3463	20.281	20.281	20.281
	BRAKE HORSEPOWER (BHP)	24	24	23.738	31.039	60.8429	60.8429	60.8429
	MATERIAL	CS		CS	CS	CS	CS	CS
10.60	<u>OIL MAKEUP PUMP</u>							
77	TYPE	RECIPROCATING	SAME AS	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING	RECIPROCATING
	CALCD MAKEUP FLOW (GPM)	0.45484		0.2027	0.27283	0.15944	0.15944	0.31888
	DESIGN FLOWRATE (GPM)	5		2	3	2	2	3
	HEAD (FT)	3093.3		3093.3	3093.3	3093.3	3093.3	3093.3
	DENSITY (LB/FT ³)	50.1		50.1	50.1	50.1	50.1	50.1
	VISCOSITY (CP)	15.71		15.71	15.71	15.71	15.71	15.71
	POWER (BHP)	22.5	22.5	9	13.5	9	9	13.5
	EFFICIENCY (%)	31.5		31.5	31.5	31.5	31.5	31.5
	DESIGN T (F)	200		200	200	200	200	200
	SUCTION P (PSIA)	17		17	17	17	17	17
	MATERIAL	CS		CS	CS	CS	CS	CS
10.62	<u>METHANOL TO STORAGE/RECYCLE PUMP</u>							
	TYPE	CENTRIFUGAL	N/A	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
	DESIGN FLOWRATE (GPM)	177		69	268	23.5	24.1	24.2
	HEAD (FT)	30		90	2240	3023	3103	3023
	DENSITY (LB/FT ³)	48.7		49.3	49.1	40.9	40.1	40.2
	VISCOSITY (CP)	0.5		0.5	0.5	0.5	0.5	0.5
	POWER (BHP)	1.70198	0	2.4	239	24	24	24
	EFFICIENCY (%)	50		50	50	50	50	50
	DESIGN T (F)	200		200	200	350	350	350
	SUCTION P (PSIA)	15		265	285	275	275	275
	MATERIAL	CS		CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
10.64 WATER INJECTION PUMP							
TYPE	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL	CENTRIFUGAL
DESIGN FLOWRATE (GPM)	53.6	53.6	31.4	48.1	18.9	18.6	18.6
HEAD (FT)	714	714	714	725	714	714	714
DENSITY (LB/FT ³)	59.3	59.3	59.3	58.5	59.3	59.3	59.3
VISCOSITY (CP)	0.138	0.138	0.136	0.136	0.136	0.136	0.136
POWER (BHP)	18.4	18.4	10.8	16.8	6.5	6.39683	6.39683
EFFICIENCY (%)	50	50	50	50	50	50	50
DESIGN T (F)	550	550	550	550	550	550	550
SUCTION P (PSIA)	800	800	800	800	800	800	800
MATERIAL	CS	CS	CS	CS	CS	CS	CS
10.65 07.65 REFLUX PUMP							
TYPE	N/A	N/A	N/A	CENTRIFUGAL	N/A	N/A	N/A
DESIGN FLOWRATE (GPM)				318			
HEAD (FT)				120			
DENSITY (LB/FT ³)				55.8			
VISCOSITY (CP)				0.8			
POWER (BHP)	0	0	0	17.2274	0	0	0
EFFICIENCY (%)				50			
DESIGN T (F)				-20			
SUCTION P (PSIA)				800			
MATERIAL				CS			
10.75 OILY WATER SUMP PUMP							
TYPE	CENTRIFUGAL	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS
INLET P (PSIA)	13.2						
HEAD (FT)	109.3						
EFFICIENCY (%)	33.0						
FLOWRATE (GPM)	10						
BRAKE HORSEPOWER (BHP)	40	40	40	40	40	40	40
MATERIAL	CS						
15.10 INCINERATOR							
TYPE							
HEAT LOAD (MMBTU/HR)	46.7447	9.70047	0.68261	12.6606	24.0175	24.8175	24.8175
EXCESS AIR (%)	50	50	50	50	50	50	50
OPERATING TEMPERATURE (F)	1800	1800	1800	1800	1800	1800	1800
WASTE GAS FLOW (LBMOL/HR)	15898.2	3329.47	3293.13	4305.98	8440.62	8440.62	8440.62
CH ₄ FUEL GAS FLOW (LBMOL/HR)	14.0737	2.94737	2.91519	3.8118	7.47194	7.47194	7.47194
COMB. AIR BLOWER POWER (HP)	50.2632	10.5263	10.4114	13.6136	26.6855	26.6855	26.6855

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
15.20	STEAM DRUM							
	TYPE	HORIZONTAL	HORIZONTAL	HORIZONTAL	HORIZONTAL	HORIZONTAL	HORIZONTAL	HORIZONTAL
	RESPONSE TIME (MIN)	30	30	30	30	30	30	30
	NORMAL % FULL	50	50	50	50	50	50	50
	BFW FEED RATE (LBS/HR)	77898	80098	101754	98193	112220	112287	112298
	SAT. WATER DENSITY (LBS/GAL)	7.18984	7.18984	7.18984	7.18984	7.18984	7.18984	7.18984
	TOTAL VOLUME (GAL)	10830	11140	14150	13660	15610	15620	15620
	DIAMETER (FT)	8	8	8.5	8.5	10	10	10
	LENGTH (FT)	28.8043	29.6288	33.337	32.1826	26.5712	26.5882	26.5882
	DESIGN PRESSURE (PSIG)	660	660	660	660	660	660	660
	DESIGN TEMPERATURE (F)	600	600	600	600	600	600	600
	MATERIAL	CS	CS	CS	CS	CS	CS	CS
15.40	UTILITY OIL HEATER							
	TYPE	ELECTRIC	SAME AS	ELECTRIC	ELECTRIC	ELECTRIC	ELECTRIC	ELECTRIC
	HEATER KW	275	275	271.098	355.655	697.158	697.158	697.158
	DESIGN T (F)	500		500	500	500	500	500
	DESIGN P (PSIG)	150		150	150	150	150	150
	MATERIAL	CS		CS	CS	CS	CS	CS
21.01	01.10 COMPRESSOR INTERCOOLER							
79	TYPE	SHELL & TUBE	SAME AS	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
	DUTY (MMBTU/HR)	24.02		21.4094	14.3979	16.8401	16.8401	16.8401
	U ASSUMED (BTU/HR-FT ² -F)	60		60	60	60	60	60
	LMOT (F)	74.6		74.6	74.6	74.6	74.6	74.6
	AREA (FT ²) + 20 % S.F.	6439.68		5739.78	3860.02	4514.77	4514.77	4514.77
	SHELL DIAMETER (IN)	54		48	42	48	48	48
	SHELL LENGTH (FT)	19.6193		22.2582	19.9246	17.5078	17.5078	17.5078
	TUBE LENGTH (FT)	19.6193		22.2582	19.9246	17.5078	17.5078	17.5078
	NUMBER OF TUBES	1003		788	592	788	788	788
	TUBE OD (IN)	1.25		1.25	1.25	1.25	1.25	1.25
	COOLING WATER FLOW (GPM)	2330	2330	2076.76	1396.63	1633.53	1633.53	1633.53
	SHELL DESIGN T (F)	250		250	250	250	250	250
	SHELL DESIGN P (PSIG)	100		100	100	100	100	100
	SHELL MATERIAL	CS		CS	CS	CS	CS	CS
	TUBES DESIGN T (F)	300		300	300	300	300	300
	TUBES DESIGN P (PSIG)	660		660	660	660	660	660
	TUBES MATERIAL	CS		CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
21.02	<u>01.10 COMPRESSOR RECYCLE CHILLER</u>							
	TYPE	SHELL & TUBE	SAME AS	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
	DUTY (MMBTU/HR)	24.02		21.4094	14.3979	16.8401	16.8401	16.8401
	U ASSUMED (BTU/HR-FT^2-F)	60		60	60	60	60	60
	LMDT (F)	74.6		74.6	74.6	74.6	74.6	74.6
	AREA (FT^2) + 20 % S.F.	6430.68		5739.78	3860.02	4514.77	4514.77	4514.77
	SHELL DIAMETER (IN)	54		48	42	48	48	42
	SHELL LENGTH (FT)	19.6193		22.2582	19.9246	17.5078	17.5078	23.3042
	TUBE LENGTH (FT)	19.6193		22.2582	19.9246	17.5078	17.5078	23.3042
	NUMBER OF TUBES	1003		788	592	788	788	592
	TUBE OD (IN)	1.25		1.25	1.25	1.25	1.25	1.25
	COOLING WATER FLOW (GPM)	2330		2076.76	1396.63	1633.53	1633.53	1633.53
	SHELL DESIGN T (F)	250		250	250	250	250	250
	DESIGN P (PSIG)	100		100	100	100	100	100
	MATERIAL	CS		CS	CS	CS	CS	CS
	TUBES DESIGN T (F)	300		300	300	300	300	300
	DESIGN P (PSIG)	350		350	350	350	350	350
	MATERIAL	CS		CS	CS	CS	CS	CS
21.03	<u>01.20 COMPRESSOR RECYCLE CHILLER</u>							
08	TYPE	SHELL & TUBE	SHELL & TUBE	N/A	SHELL & TUBE	N/A	N/A	N/A
	DUTY (MMBTU/HR)	1.6	1.34		1.05			
	U ASSUMED (BTU/HR-FT^2-F)	60	60		60			
	LMDT (F)	40	40		40			
	AREA (FT^2) + 20 % S.F.	800	670		525			
	SHELL DIAMETER (IN)	28	28		28			
	SHELL LENGTH (FT)	10.3149	8.63869		9.43695			
	TUBE LENGTH (FT)	10.3149	8.63869		9.43695			
	NUMBER OF TUBES	237	237		170			
	TUBE OD (IN)	1.25	1.25		1.25			
	COOLING WATER FLOW (GPM)	155.204	129.983		101.853			
	SHELL DESIGN T (F)	250	250		250			
	DESIGN P (PSIG)	100	100		100			
	MATERIAL	CS	CS		CS			
	TUBES DESIGN T (F)	250	250		250			
	DESIGN P (PSIG)	1320	1320		1320			
	MATERIAL	CS	CS		CS			

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
21.11	REACTOR FEED/PRODUCT DEPHLEGMATOR							
	TYPE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
	DUTY (MMBTU/HR)	53.95	54.14	22.65	33.33	18.01	18.01	18
	U ASSUMED (BTU/HR-FT ² -F)	30	30	30	30	30	30	30
	LMTD (F)	80	79.2	81.4	77	80.3	80.4	80.4
	AREA (FT ²) + 20% S.F.	26975	27343.4	11130.2	17314.3	8971.36	8960.2	8955.22
	SHELL DIAMETER (IN)	108	108	72	84	72	72	72
	SHELL HEIGHT (FT)	29.7578	30.0277	28.7494	31.1043	25.1127	25.0939	25.0855
	TUBE HEIGHT (FT)	19.7578	20.0277	18.7494	21.1043	15.1127	15.0939	15.0855
	NUMBER OF TUBES	4172	4172	1814	2507	1814	1814	1814
	TUBE OD (IN)	1.25	1.25	1.25	1.25	1.25	1.25	1.25
	SHELL DESIGN T (F)	600	600	600	600	600	600	600
	SHELL DESIGN P (PSIG)	1320	1320	1320	1320	1320	1320	1320
	SHELL MATERIAL	SS CLAD	SS CLAD	SS CLAD	SS CLAD	SS CLAD	SS CLAD	SS CLAD
	TUBES DESIGN T (F)	600	600	600	600	600	600	600
	TUBES DESIGN P (PSIG)	1320	1320	1320	1320	1320	1320	1320
	TUBES MATERIAL	SS	SS	SS	SS	SS	SS	SS
21.20	REACTOR INTERNAL HEAT EXCHANGER							
181	HEAT DUTY (MMBTU/HR)	80	82.2	104.5	100.8	115.3	115.3	115.3
	STEAM PRODUCED (LBS/HR)	94643	97313	123606	119300	136343	136424	136438
	U _{overall} (BTU/HR-FT ² -F) with factor	113	113	104	105	96	96	96
	LMTD (F)	83	83	82.7	82.7	82.7	82.7	82.7
	AREA (FT ²)	8529.69	8784.28	12150	11608.2	14522.9	14522.9	14522.9
	PIPES PRODUCED STEAM P (PSIG)	230	230	230	230	230	230	230
	EXTERNAL DESIGN P (PSIG)	1320	1320	1320	1320	1320	1320	1320
	INTERNAL DESIGN P (PSIG)	750	750	750	750	750	750	750
	DESIGN T (F)	600	600	600	600	600	600	600
	MATERIAL	304 SS	304 SS	304 SS	304 SS	304 SS	304 SS	304 SS
	# PIPES	297	304	361	352	339	339	339
	PIPE LENGTH (FT)	57.7371	57.9588	67.6625	66.2981	86.1253	86.1253	86.1253
	NOMINAL PIPE OD (IN)	1.9	1.9	1.9	1.9	1.9	1.9	1.9
	% OF SLURRY VOLUME	4	4.1	6.6	4.8	5.3	5.3	5.3

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
21.31	METHANOL PRODUCT CONDENSER							
	TYPE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
	DUTY (MMBTU/HR)	65.67	81.09	29.99	40.59	11.12	20.75	20.74
	U ASSUMED (BTU/HR-FT ² -F)	60	60	60	60	60	60	60
	LMTD (F)	63.7	72.9	70.1	68.5	128	65.3	65.6
	AREA (FT ²) + 20% S.F.	20616.5	22246.9	8556.35	11851.1	1737.5	6355.28	6323.17
	SHELL DIAMETER (IN)	78	78	60	72	35	54	54
	SHELL LENGTH (FT)	29.4281	31.7522	21.1368	19.9638	12.7324	19.3622	19.2644
	TUBE LENGTH (FT)	29.4281	31.7522	21.1368	19.9638	12.7324	19.3622	19.2644
	NUMBER OF TUBES	2141	2141	1237	1814	417	1003	1003
	TUBE OD (IN)	1.25	1.25	1.25	1.25	1.25	1.25	1.25
	COOLING WATER FLOW (GPM)	6370	7865	2908	3952	1079	2012	2011
	SHELL DESIGN T (F)	250	250	250	250	250	250	250
	DESIGN P (PSIG)	100	100	100	100	100	100	100
	MATERIAL	CS	CS	CS	CS	CS	CS	CS
	TUBES DESIGN T (F)	400	400	400	400	400	400	400
	DESIGN P (PSIG)	1320	1320	1320	1320	1320	1320	1320
	MATERIAL	304 SS	304 SS	304 SS	304 SS	304 SS	304 SS	304 SS
21.32	RECOVERY SECTION CROSS EXCHANGER							
82	TYPE	N/A	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	N/A	SHELL & TUBE	SHELL & TUBE
	DUTY (MMBTU/HR)		10.02	9.86	11.07		7.76	7.8
	U ASSUMED (BTU/HR-FT ² -F)		60	60	60		60	60
	LMTD (F)		31.45	35.6	23.6		19	19
	AREA (FT ²) + 20% S.F.		6372.02	5539.33	9381.36		8168.42	8210.53
	SHELL DIAMETER (IN)		54	54	66		60	60
	SHELL LENGTH (FT)		19.4132	16.8763	18.8601		20.1785	20.2826
	TUBE LENGTH (FT)		19.4132	16.8763	18.8601		20.1785	20.2826
	NUMBER OF TUBES		1003	1003	1520		1237	1237
	TUBE OD (IN)		1.25	1.25	1.25		1.25	1.25
	SHELL DESIGN T (F)		-100	-100	-20		-20	-20
	DESIGN P (PSIG)		1320	1320	1320		350	350
	MATERIAL		SS	SS	CS		SS	CS
	TUBES DESIGN T (F)		-100	-100	-20		-20	-20
	DESIGN P (PSIG)		1320	1320	1320		1320	1320
	MATERIAL		SS	SS	CS		SS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
21.33 RECOVERY SECTION ECONOMIZER							
TYPE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
DUTY (MMBTU/HR)	16.08	9.8	2.47	6.14	6.8	6.88	6.9
U ASSUMED (BTU/HR-FT ² -F)	60	60	60	60	60	60	60
LMTD (F)	58.07	36.1	42.7	21.9	70.9	24	23.84
AREA (FT ²) + 20% S.F.	5538.14	5429.38	1156.91	5607.31	1918.19	5733.33	5788.59
SHELL DIAMETER (IN)	54	54	54	54	33	54	54
SHELL LENGTH (FT)	16.8727	16.5413	13.1912	17.0834	16.419	17.4674	17.6357
TUBE LENGTH (FT)	16.8727	16.5413	13.1912	17.0834	16.419	17.4674	17.6357
NUMBER OF TUBES	1003	1003	268	1003	357	1003	1003
TUBE OD (IN)	1.25	1.25	1.25	1.25	1.25	1.25	1.25
SHELL DESIGN T (F)	400	-100	-100	-100	20	-100	-100
SHELL DESIGN P (PSIG)	1320	350	350	350	350	350	350
SHELL MATERIAL	CS	SS	SS	SS	CS	SS	SS
TUBES DESIGN T (F)	400	-100	-100	-100	20	-100	-100
TUBES DESIGN P (PSIG)	1320	1320	1320	1320	1320	1320	1320
TUBES MATERIAL	304 SS	SS	SS	SS	CS	SS	SS
21.40 REDUCTION FEED/PRODUCT ECONOMIZER							
TYPE	SHELL & TUBE	SAME AS	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
DUTY (MMBTU/HR)	0.30078		0.29748	0.38897	0.76246	0.76246	0.76246
U ASSUMED (BTU/HR-FT ² -F)	50		50	50	50	50	50
LMTD (F)	96.06		96.06	96.06	96.06	96.06	96.06
AREA (FT ²) + 30% S.F.	81.405		80.5163	105.28	206.371	206.371	206.371
SHELL DESIGN T (F)	500		500	500	500	500	500
SHELL DESIGN P (PSIG)	150		150	150	150	150	150
SHELL MATERIAL	304 SS		304 SS	304 SS	304 SS	304 SS	304 SS
TUBES DESIGN T (F)	500		500	500	500	500	500
TUBES DESIGN P (PSIG)	150		150	150	150	150	150
TUBES MATERIAL	304 SS		304 SS	304 SS	304 SS	304 SS	304 SS
21.41 01.40 REDUCTION GAS COMPRESSOR COOLERS							
TYPE	SHELL & TUBE	SAME AS	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
STAGE 1 COOLER DUTY (MMBTU/HR)	0.22572		0.22326	0.29193	0.57224	0.57224	0.57224
COOLING WATER FLOW (GPM)	21.9054		21.6663	28.3301	55.5329	55.5329	55.5329
STAGE 2 COOLER DUTY (MMBTU/HR)	0.22572		0.22326	0.29193	0.57224	0.57224	0.57224
COOLING WATER FLOW (GPM)	21.9054		21.6663	28.3301	55.5329	55.5329	55.5329
TOTAL CW FLOW (GPM)	43.8109	43.8109	43.3326	56.6602	111.066	111.066	111.066

ISSUED FOR ESTIMATE

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
21.53	<u>UTILITY OIL AIR COOLER</u>							
	TYPE	AIR COOLER	SAME AS	AIR COOLER	AIR COOLER	AIR COOLER	AIR COOLER	AIR COOLER
	DUTY (MMBTU/HR)	0.5		0.49454	0.64665	1.26756	1.26756	1.26756
	AREA (FT ²) + 20% S.F.	1538		1521.21	1989.08	3899.02	3899.02	3899.02
	FAN MOTOR (BHP)	5.28514	5.28514	5.22744	6.83522	13.3985	13.3985	13.3985
	TUBES DESIGN T (F)	500		500	500	500	500	500
	DESIGN P (PSIG)	150		150	150	150	150	150
	MATERIAL	CS		CS	CS	CS	CS	CS
21.60	<u>INSTRUMENT AIR COMPRESSOR COOLERS</u>							
	TYPE	SHELL & TUBE	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS
	STAGE 1 COOLER DUTY (MMBTU/HR)	0.34253						
	COOLING WATER FLOW (GPM)	33.2408						
	STAGE 2 COOLER DUTY (MMBTU/HR)	0.29971						
	COOLING WATER FLOW (GPM)	29.0857						
	AFTERCOOLER DUTY (MMBTU/HR)	0.29971						
	COOLING WATER FLOW (GPM)	29.0857						
21.61	<u>01.30 RECYCLE COOLER</u>							
	TYPE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
	DUTY (MMBTU/HR)	0.18932	0.03965	0.03922	0.05128	0.10051	0.10051	0.10051
	U ASSUMED (BTU/HR-FT ² -F)	60	60	60	60	60	60	60
	LMTD (F)	27.26	27.26	27.26	27.26	27.26	27.26	27.26
	AREA (FT ²) + 20% S.F.	138.9	29.0891	28.7715	37.6206	73.7444	73.7444	73.7444
	SHELL DIAMETER (IN)	12	12	8	8	12	8	8
	SHELL LENGTH (FT)	11.1697	2.3392	7.32661	9.58002	5.93016	18.7788	18.7788
	TUBE LENGTH (FT)	11.1697	2.3392	7.32661	9.58002	5.93016	18.7788	18.7788
	NUMBER OF TUBES	38	38	12	12	38	12	12
	TUBE OD (IN)	1.25	1.25	1.25	1.25	1.25	1.25	1.25
	COOLING WATER FLOW (GPM)	18.4317	3.86004	3.81791	4.99216	9.78568	9.78568	9.78568
	SHELL DESIGN T (F)	250	250	250	250	250	250	250
	DESIGN P (PSIG)	100	100	100	100	100	100	100
	MATERIAL	CS	CS	CS	CS	CS	CS	CS
	TUBES DESIGN T (F)	300	300	300	300	300	300	300
	DESIGN P (PSIG)	50	50	50	50	50	50	50
	MATERIAL	CS	CS	CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
21.71	LIQUID PRODUCT COOLER							
	TYPE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	N/A	N/A	N/A
	DUTY (MMBTU/HR)	2.52	13.04	3.19	3.39			
	U ASSUMED (BTU/HR-FT ² -F)	60	60	60	60			
	LMTD (F)	34.4	91.9	45.3	46.5			
	AREA (FT ²) + 20% S.F.	1455.12	2837.87	1408.39	1458.06			
	SHELL DIAMETER (IN)	27	31	27	27			
	SHELL LENGTH (FT)	18.8906	27.7045	18.1591	18.7996			
	TUBE LENGTH (FT)	18.8906	27.7945	18.1591	18.7996			
	NUMBER OF TUBES	237	312	237	237			
	TUBE OD (IN)	1.25	1.25	1.25	1.25	0	0	0
	COOLING WATER FLOW (GPM)	244	1265	309	328.373			
	SHELL DESIGN T (F)	250	250	250	250			
	SHELL DESIGN P (PSIG)	100	100	100	100			
	SHELL MATERIAL	CS	CS	CS	CS			
	TUBES DESIGN T (F)	250	350	250	250			
	TUBES DESIGN P (PSIG)	100	350	350	350			
	TUBES MATERIAL	CS	CS	CS	CS			
21.72	LIQUID RECYCLE COOLER							
	TYPE	N/A	N/A	SHELL & TUBE	SHELL & TUBE	N/A	N/A	N/A
	DUTY (MMBTU/HR)			5.21	13.23			
	U ASSUMED (BTU/HR-FT ² -F)			60	60			
	LMTD (F)			93.6	46.2			
	AREA (FT ²) + 20% S.F.			1113.25	5727.27			
	SHELL DIAMETER (IN)			24	54			
	SHELL LENGTH (FT)			20.0108	17.4489			
	TUBE LENGTH (FT)			20.0108	17.4489			
	NUMBER OF TUBES			170	1003			
	TUBE OD (IN)			1.25	1.25			
	COOLING WATER FLOW (GPM)	0	0	506	0	0	0	0
	SHELL DESIGN T (F)			350	350			
	SHELL DESIGN P (PSIG)			100	350			
	SHELL MATERIAL			CS	CS			
	TUBES DESIGN T (F)			350	350			
	TUBES DESIGN P (PSIG)			350	350			
	TUBES MATERIAL			CS	CS			

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
21.73 LIQUID RECYCLE CROSS EXCHANGER							
TYPE	N/A	N/A	SHELL & TUBE	SHELL & TUBE	N/A	N/A	N/A
DUTY (MMBTU/HR)			2.53	3.93			
U ASSUMED (BTU/HR-FT ² -F)			60	60			
LMTD (F)			21.7	24.1			
AREA (FT ²) + 20% S.F.			2331.8	3261.41			
SHELL DIAMETER (IN)			31	37			
SHELL LENGTH (FT)			22.838	22.3456			
TUBE LENGTH (FT)			22.838	22.3456			
NUMBER OF TUBES			312	446			
TUBE OD (IN)			1.25	1.25			
SHELL DESIGN T (F)			-80	20			
DESIGN P (PSIG)			350	350			
MATERIAL			SS	CS			
TUBES DESIGN T (F)			-80	20			
DESIGN P (PSIG)			350	1320			
MATERIAL			SS	CS			
21.80 LIQUID FUEL VAPORIZOR							
TYPE	KETTLE	KETTLE	KETTLE	KETTLE	N/A	N/A	KETTLE
DUTY (MMBTU/HR)	153	159	74.6	79			34.9
U ASSUMED (BTU/HR-FT ² -F)	100	100	100	100			100
LMTD (F)	70	70	124	121			230
AREA (FT ²) + 20% S.F.	26228.6	27257.1	7219.35	7834.71			1820.87
SHELL DIAMETER (IN)	156	156	84	84			66
SHELL LENGTH (FT)	37.4351	38.9032	37.2647	40.4411			15.5859
TUBE LENGTH (FT)	37.4351	38.9032	37.2647	40.4411			15.5859
NUMBER OF TUBES	2141	2141	592	592			357
TUBE OD (IN)	1.25	1.25	1.25	1.25			1.25
STEAM FLOW (#/HR)	185096	192354	89490	93517	0	0	41342
STEAM PRESSURE (PSIG)	230	230	230	230	0	0	230
SHELL DESIGN T (F)	350	350	350	350			350
DESIGN P (PSIG)	350	350	350	350			350
MATERIAL	CS	CS	CS	CS			CS
TUBES DESIGN T (F)	500	500	500	500			500
DESIGN P (PSIG)	350	350	350	350			350
MATERIAL	CS	CS	CS	CS			CS

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
22.02 FEED GAS COMP 2ND STAGE KNOCKOUT							
TYPE	VERTICAL	SAME AS	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL
VESSEL I.D. (FT)	8		7.25	6	6.5	6.5	6.5
T-T HEIGHT (FT)	12.6687		12.1	10.9	11.4	11.4	11.4
DESIGN T (F)	300		300	300	300	300	300
DESIGN P (PSIG)	600		600	600	600	600	600
VAPOR Kv	0.19		0.19	0.19	0.2	0.2	0.2
MATERIAL	CS		CS	CS	CS	CS	CS
22.10 H.P. METHANOL SEPARATOR							
TYPE	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL
VESSEL ID (FT)	10	9.7	8.1	7.5	5.3	5.3	5.3
T-T HEIGHT (FT)	13.5	13.3	10.4	11.5	9.7	9.6	9.6
DESIGN T (F)	400.0	400	400	400	400	400	400
DESIGN P (PSIG)	1320	1320	1320	1320	1320	1320	1320
VAPOR Kv	0.19	0.2	0.2	0.2	0.2	0.2	0.2
VESSEL MATERIAL	SS CLAD ON CS	SS CLAD ON CS	CS	SS CLAD ON CS	SS CLAD ON CS	CS	SS CLAD ON CS
22.11 H.P. LIQUIDS SEPARATOR							
87 TYPE	N/A	VERTICAL	VERTICAL	VERTICAL	N/A	VERTICAL	VERTICAL
VESSEL ID (FT)		5	5.7	6.4		4.5	4.5
T-T HEIGHT (FT)		9.5	10	10.6		9	9
DESIGN T (F)		-100	-20	-20		20	20
DESIGN P (PSIG)		1320	1320	1320		1320	1320
VAPOR Kv		0.21	0.21	0.2		0.2	0.2
MATERIAL		SS	CS	CS		CS	CS
22.12 M.P. FLASH POT/ABSORBER							
TYPE	VERTICAL	VERTICAL	N/A	VERTICAL	VERTICAL	VERTICAL	VERTICAL
VESSEL ID (FT)	2.5	2.5		3.5	6.9	5.5	5.5
T-T HEIGHT (FT)	8.6	8.2		9	11.7	10.5	10.5
DESIGN T (F)	200.0	200		100	-20	-100	-100
DESIGN P (PSIG)	350.0	350		350	350	350	350
VAPOR Kv	0.0	0.03		0.06	0.2	0.2	0.2
MATERIAL	CS	CS		CS	CS	SS	SS

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
22.13	<u>L.P. FLASH POT</u>							
	TYPE	VERTICAL	N/A	N/A	N/A	N/A	N/A	N/A
	VESSEL ID (FT)	2.5						
	T-T HEIGHT (FT)	9.1						
	DESIGN T (F)	200.0						
	DESIGN P (PSIG)	100.0						
	VAPOR Kv	0.1						
	MATERIAL	CS						
22.40	<u>REDUCTION CONDENSATE ACCUMULATOR</u>							
	TYPE	VERTICAL TANK	SAME AS	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK
	VESSEL ID (FT)	1.66667		1.65754	1.89538	2.65368	2.65368	2.65368
	T-T HEIGHT (FT)	7.66667		7.62471	8.71876	12.2069	12.2069	12.2069
	DESIGN T (F)	250		250	250	250	250	250
	DESIGN P (PSIG)	150		150	150	150	150	150
	VAPOR Kv	0.199		0.199	0.199	0.199	0.199	0.199
	MATERIAL	CS		CS	CS	CS	CS	CS
22.50	<u>RECLAIMED OIL FILTER</u>							
∞	TYPE	CARTRIDGE	SAME AS	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE
	FLOWRATE (GPM)	5		5	6.46645	12.6756	12.6756	12.6756
	PARTICLE SIZE RATING (MICRON)	5		5	5	5	5	5
	DESIGN T (F)	250		250	250	250	250	250
	DESIGN P (PSIG)	25		25	25	25	25	25
	MATERIAL	CS		CS	CS	CS	CS	CS
22.51	<u>LIQUID PRODUCT/RECYCLE FILTER</u>							
	TYPE	CARTRIDGE	N/A	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE
	FLOWRATE (GPM)	177		69	268	23.5	24.1	24.2
	PARTICLE SIZE RATING (MICRON)	5		5	5	5	5	5
	DESIGN T (F)	250		250	250	250	250	250
	DESIGN P (PSIG)	100		350	350	350	350	350
	MATERIAL	CS		CS	CS	CS	CS	CS
22.52	<u>TURBINE FUEL FILTER</u>							
	TYPE	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE	N/A	N/A	CARTRIDGE
	FLOWRATE (GPM)	712	752	829	635			613
	PARTICLE SIZE RATING (MICRON)	5	5	5	5			5
	DESIGN T (F)	150	150	150	150			150
	DESIGN P (PSIG)	350	350	350	350			350
	MATERIAL	CS	CS	CS	CS			CS

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
22.53	<u>02.10 GUARD BED REGEN TO TURBINE FILTER</u>							
	TYPE	CARTRIDGE	SAME AS	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE
	FLOWRATE (LBMOL/HR)	21.3773		19.0539	12.8138	14.9873	14.9873	14.9873
	PARTICLE SIZE RATING (MICRON)	5		5	5	5	5	5
	FILTER DIAMETER (IN)	6		6	5	6	6	5
	FILTER HEIGHT (IN)	18		18	15	18	18	15
	DESIGN T (F)	500		500	500	500	500	500
	DESIGN P (PSIG)	600		600	600	600	600	600
	MATERIAL	304 SS		304 SS	304 SS	304 SS	304 SS	304 SS
22.54	<u>02.10 GUARD BED REGEN HEATER OUTLET FILTER</u>							
	TYPE	CARTRIDGE	SAME AS	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE	CARTRIDGE
	FLOWRATE (LBMOL/HR)	21.3773		19.0539	12.8138	14.9873	14.9873	14.9873
	PARTICLE SIZE RATING (MICRON)	5		5	5	5	5	5
	FILTER DIAMETER (IN)	6		6	6	6	6	6
	FILTER HEIGHT (IN)	18		18	18	18	18	18
	DESIGN T (F)	500		500	500	500	500	500
	DESIGN P (PSIG)	1320		1320	1320	1320	1320	1320
	MATERIAL	304 SS		304 SS	304 SS	304 SS	304 SS	304 SS
22.90	<u>OILY WATER SEPARATOR</u>							
	TYPE		SAME AS	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS
	DESIGN % FULL	80						
	TOTAL VOLUME (GAL)	10000						
	WIDTH AND DEPTH (FT)	10						
	LENGTH (FT)	18.709						
	MATERIAL	CONCRETE						

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
27.10 LPMEOH REACTOR							
TYPE	SLURRY	SLURRY	SLURRY	SLURRY	SLURRY	SLURRY	SLURRY
FEED RATE (LB-MOL/HR)	36782	36783	16590	22453	13155	13144	22453
DESIGN S.V. (SL/HR-KG)	6250	6250	2850	2950	1320	1320	1320
NORMAL OPERATING T (F)	482	482	482	482	482	482	482
NORMAL OPERATING P (PSIA)	1060	1060	1060	1060	1060	1060	1060
DESIGN SUP. GAS VEL. (FT/S)	0.75	0.75	0.47	0.47	0.29	0.29	0.29
DESIGN GAS HOLDUP (%)	47	42	43	41	43	43	43
DESIGN CATALYST CONC. (WT% OXIDE)	40	40	40	40	40	40	40
CATALYST INVENTORY (kg)	59850	59852	59197	77404	101347	101347	101347
DIAMETER (FT)	13.6	13.6	11.7	13.6	13.2	13.2	13.2
T-T HEIGHT (FT)	78.1	78.1	90.7	90.6	120.8	120.8	120.8
SLURRY HEIGHT (FT)	64.5	64.5	80.8	79	110	110	110
SLURRY L/D	4.74265	4.74265	6.90598	5.80882	8.33333	8.33333	8.33333
GASSED SLURRY VOLUME (GAL)	67508	67548	60822	81444	106132	106132	106132
MATERIAL	SS CLAD ON CS	SS CLAD ON CS	SS CLAD ON CS	SS CLAD ON CS	SS CLAD ON CS	SS CLAD ON CS	SS CLAD ON CS
DESIGN T (F)	600	600	600	600	600	600	600
DESIGN P (PSIG)	1320	1320	1320	1320	1320	1320	1320
28.10 OIL STORAGE TANK							
DESCRIPTION	VERTICAL TANK	SAME AS	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK
CAPACITY (GAL)	5000		4945.42	6466.45	12675.6	12675.6	12675.6
VESSEL ID (FT)	6.5		6.5	7	6.5	6.5	7
T-T HEIGHT (FT)	20.1416		19.9217	22.4606	51.0614	51.0614	44.0274
DESIGN T (F)	200		200	200	200	200	200
DESIGN P (PSIG)	10		10	10	10	10	10
MATERIAL	CS		CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
28.30	<u>SLURRY MIX TANK & REDUCTION VESSEL</u>							
	DESCRIPTION	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS
	VESSEL ID (FT)	5.5	5.5	6.25476	8.75714	8.75714	8.75714	8.75714
	T-T HEIGHT (FT)	11.1596	11.0923	11.1596	11.1596	11.1596	11.1596	11.1596
	TOTAL VESSEL VOL (FT ³)	265.132	263.534	342.893	672.142	672.142	672.142	672.142
	LB OXIDE PER RED'N	1816	1796.18	2348.62	4603.78	4603.78	4603.78	4603.78
	KG OXIDE PER RED'N	824.464	815.464	1066.27	2090.12	2090.12	2090.12	2090.12
	STARTING WT% OXIDE	35.2941	35.2941	35.2941	35.2941	35.2941	35.2941	35.2941
	ENDING WT% OXIDE	40	40	40	40	40	40	40
	GAL OIL PER RED'N	475.32	470.131	614.726	1204.99	1204.99	1204.99	1204.99
	UNGASSED SLURRY VOL (FT ³)	87.8043	86.8458	113.557	222.595	222.595	222.595	222.595
	ASSUMED % GAS HOLDUP	40	40	40	40	40	40	40
	GASSED SLURRY VOL (FT ³)	146.341	144.743	189.261	370.991	370.991	370.991	370.991
	SLURRY HEIGHT (FT)	8.15956	6.09231	6.15956	6.15956	6.15956	6.15956	6.15956
	SLURRY L/D	1.11992	1.10769	0.98478	0.70338	0.70338	0.70338	0.70338
	FREEBOARD HEIGHT (FT)	5	5	5	5	5	5	5
	FREEBOARD L/D	0.90909	0.90909	0.79939	0.57096	0.57096	0.57096	0.57096
16	JACKET HEIGHT (FT)	5.15956	5.09231	5.15956	5.15956	5.15956	5.15956	5.15956
	JACKET H.T. AREA (FT ²)	89.1507	87.9889	101.385	141.946	141.946	141.946	141.946
	REDUCTION GAS FLOW (SCFH)	82724	81820.9	106986	209715	209715	209715	209715
	SUP. GAS VEL. (FT/S)	0.20615	0.2039	0.20615	0.20615	0.20615	0.20615	0.20615
	SPACE VELOCITY (SL/HR-KG OXIDE)	2841.54	2841.54	2841.54	2841.54	2841.54	2841.54	2841.54
	DESIGN T (F)	500	500	500	500	500	500	500
	DESIGN P (PSIG)	150	150	150	150	150	150	150
	MATERIAL	304 SS	304 SS	304 SS	304 SS	304 SS	304 SS	304 SS
28.31	<u>SLURRY MIX TANK & REDUCTION VESSEL AGITATOR</u>							
	TYPE	MOTOR & IMP	SAME AS	MOTOR & IMP	MOTOR & IMP	MOTOR & IMP	MOTOR & IMP	MOTOR & IMP
	AGITATOR MOTOR HP	20	20	19.7817	25.8658	50.7024	50.7024	50.7024
	DESIGN T (F)	500	500	500	500	500	500	500
	DESIGN P (PSIG)	150	150	150	150	150	150	150
	MATERIAL	304 SS	304 SS	304 SS	304 SS	304 SS	304 SS	304 SS

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
28.35 CATALYST BIN, HOPPER & FEEDER								
DESCRIPTION			SAME AS	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS
BIN CROSS SECTIONAL AREA (FT^2)	25			25	30	35	35	35
BIN HEIGHT (FT)	6.29327			6.29327	6.55549	10.1142	10.1142	10.1142
BIN VOLUME (FT^3)	157.332			157.332	196.665	353.996	353.996	353.996
BIN CAP (KG OXIDE POWDER)	4000			4000	5000	9000	9000	9000
TYPICAL LOAD (KG OXIDE POWDER)	1816			1796.18	2348.62	4603.78	4603.78	4603.78
FILL TIME/BATCH (MIN)	15			15	15	15	15	15
FEEDER MOTOR (BHP)	20	20		19.7817	25.8658	50.7024	50.7024	50.7024
MATERIAL	CS			CS	CS	CS	CS	CS
28.40 LIQUID FUEL STORAGE								
TYPE	VERTICAL TANK	VERTICAL TANK		HORIZONTAL	HORIZONTAL	N/A	N/A	HORIZONTAL
NUMBER	1	1		20	20			20
HOLD TIME (DAYS PER TANK)	5	5		0.25	0.25			0.25
DESIGN % FULL	80	80		80	80			80
VESSEL ID (FT)	82	82		12	12			12
HEIGHT (FT)	41	41		83.5	84.3			82
DESIGN PRESSURE (PSIG)	ATMOSPHERIC	ATMOSPHERIC		110	110			150
DESIGN TEMP (F)	200	200		200	200			200
MATERIAL	CS	CS		CS	CS			CS
28.45 SLURRY CENTRIFUGE								
CONE DIAMETER (FT)	2.33333	SAME AS		2.32056	2.65354	3.71515	3.71515	3.71515
CONE LENGTH (FT)	2.5			2.5	2.5	2.5	2.5	2.5
MOTOR POWER (BHP)	58	58		55.6935	63.6849	89.1638	89.1636	89.1636
MATERIAL	SS			SS	SS	SS	SS	SS
28.46 SLURRY CENTRIFUGE SURGE POT								
HOLDUP TIME (MIN)	5	SAME AS		5	5	5	5	5
LIQUID FLOWRATE (GPM)	5.2			5.17154	5.91359	8.27947	8.27947	8.27947
DESIGN % FULL	30			30	30	30	30	30
VESSEL VOLUME (GAL)	90			90	100	140	140	140
VESSEL DIAMETER (FT)	2			2	2	2.2	2.2	2.2
T-T HEIGHT (FT)	3.02042			3.82942	4.25401	4.02304	4.02304	4.02304
DESIGN T (F)	200			200	200	200	200	200
DESIGN P (PSIG)	5			5	5	5	5	5
MATERIAL	CS			CS	CS	CS	CS	CS

TABLE 4: EQUIPMENT SPECIFICATIONS

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
28.50 FLARE KNOCKOUT POT							
TYPE	HOR TANK	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS	SAME AS
VESSEL I.D. (FT)	6.7082						
T-T LENGTH (FT)	9						
VESSEL VOLUME (GAL)	2379.6						
DESIGN LIQUID HOLDUP TIME (MIN)	30						
DESIGN LIQUID HEIGHT (FT)	1.4						
VAPOR VELOCITY (FT/SEC)	11.38						
LIQUID DROPOUT VELOCITY (FT/SEC)	2.28						
DESIGN T (F)	500						
DESIGN P (PSIG)	75						
MATERIAL	CS						
28.53 UTILITY OIL EXPANSION TANK							
TYPE	VERTICAL	SAME AS	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL
WORKING VOLUME (GAL)	450		445.088	581.981	1140.8	1140.8	1140.8
NOMINAL VOLUME (GAL)	692.308		684.75	895.355	1755.08	1755.08	1755.08
% LEVEL HOT	75		75	75	75	75	75
% LEVEL COLD	10		10	10	10	10	10
VESSEL DIAMETER (FT)	3.5		3.5	4	3.5	3.5	4
T-T HEIGHT (FT)	9.81092		9.51491	9.52542	24.3877	24.3877	18.6718
DESIGN T (F)	500		500	500	500	500	500
DESIGN P (PSIG)	150		150	150	150	150	150
MATERIAL	CS		CS	CS	CS	CS	CS
28.60 SLURRY TANK							
TYPE	VERTICAL TANK	SAME AS	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK	VERTICAL TANK
DESIGN % FULL	80		80	80	80	80	80
TOTAL SLURRY VOLUME (GAL)	35838		35464	46372	60495	60495	60495
VESSEL VOLUME (GAL)	44707.5		44330	57965	75618.8	75618.8	75618.8
VESSEL I.D. (FT)	15		15	15	18	18	18
T-T LENGTH (FT)	33.8861		33.5325	43.8464	39.7224	39.7224	39.7224
DESIGN T (F)	500		500	500	500	500	500
DESIGN P (PSIG)	50		50	50	50	50	50
MATERIAL	CS		CS	CS	CS	CS	CS
28.61 SLURRY TANK AGITATOR							
MOTOR HP	37.3157	37	36.9263	48.284	62.9894	62.9894	62.9894
DESIGN T (F)	500		500	500	500	500	500
DESIGN P (PSIG)	75		75	75	75	75	75

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
30.00	<u>SULFUR REMOVAL UNIT</u>							
	E-50 FT2	17500	17500	15598	10489.7	12269	12269	12269
	E-51 FT2	5660	5660	5044.84	3392.67	3968.15	3968.15	3968.15
	E-52 FT2	20700	20700	18450.2	12407.8	14512.5	14512.5	14512.5
	E-53 FT2	7700	7700	6943.34	4669.42	5461.46	5461.46	5461.46
	F-50 DIA (FT)	8.5	8.5	7.57617	5.095	5.95923	5.95923	5.95923
	T-T (FT)	39	39	39	39	39	39	39
	F-51 DIA (FT)	9.6	9.6	8.55662	5.75435	6.73042	6.73042	6.73042
	T-T (FT)	48	48	48	48	48	48	48
	F-53 DIA (FT)	6.4	6.4	5.70441	3.83624	4.48695	4.48695	4.48695
	T-T (FT)	8	8	7.5	6.2	7.5	7.5	6.2
	F-54 DIA (FT)	6.7	6.7	5.97181	4.01606	4.69727	4.69727	4.69727
	T-T (FT)	8.2	8.2	7.7	6.5	7.7	7.7	6.5
	P-50 GPM	865	865	770.987	518.491	606.439	606.439	606.439
	BHP	432	432	385.048	258.946	302.869	302.869	302.869
40.00	<u>PROPANE REFRIGERATION SKID</u>							
94	TYPE	2-STG PROPANE	2-STG PROPANE	2-STG PROPANE	2-STG PROPANE	2-STG PROPANE	2-STG PROPANE	2-STG PROPANE
	REFRIGERATION LOAD (TONS)	1041.67	1041.67	1065.95	1664.68	730.298	730.298	1162.93
	BOILING TEMPERATURE (F)	-21	-21	-21	-21	-21	-21	-21
	CONDENSING TEMPERATURE (F)	120	120	120	120	120	120	120
40.10	<u>COMPRESSOR</u>							
	GAS M.W. (LB/LB-MOLE)	44	44	44	44	44	44	44
	MATERIAL	SS/CS	SS/CS	SS/CS	SS/CS	SS/CS	SS/CS	SS/CS
	MOTOR POWER (HP)	2913.61	2913.61	2981.54	4656.16	2042.69	2042.69	3252.78
	1ST STAGE							
	INLET P (PSIA)	25	25	25	25	25	25	25
	OUTLET P (PSIA)	78	78	78	78	78	78	78
	ADIABATIC EFFICIENCY (%)	75	75	75	75	75	75	75
	COMPRESSION POWER (HP)	1044.17	1044.17	1068.51	1668.66	732.053	732.053	1165.72
	FLOW (#MOLE/HR)	1989.18	1989.18	2035.56	3178.86	1394.59	1394.59	2220.74
	2ND STAGE							
	INLET P (PSIA)	78	78	78	78	78	78	78
	OUTLET P (PSIA)	248	248	248	248	248	248	248
	ADIABATIC EFFICIENCY (%)	75	75	75	75	75	75	75
	COMPRESSION POWER (HP)	1723.76	1723.76	1763.95	2754.69	1208.5	1208.5	1924.42
	FLOW (#MOLE/HR)	3023.84	3023.84	3094.34	4832.32	2119.07	2119.07	3375.84

TABLE 4: EQUIPMENT SPECIFICATIONS

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
40.20	STORAGE TANK							
	TYPE	HORIZONTAL	HORIZONTAL	HORIZONTAL	HORIZONTAL	HORIZONTAL	HORIZONTAL	HORIZONTAL
	VESSEL ID (FT)	6	6	6	7	5.5	5.5	7
	T-T HEIGHT (FT)	22.6502	22.6502	23.1782	26.5935	18.8982	18.8982	18.5781
	DESIGN T (F)	150	150	150	150	150	150	150
	DESIGN P (PSIG)	265	265	265	265	265	265	265
	VAPOR Kv							
	VESSEL MATERIAL	CS	CS	CS	CS	CS	CS	CS
40.21	FLASH POT							
	TYPE	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL	VERTICAL
	VESSEL ID (FT)	4.20338	4.20338	4.25208	5.31368	3.51951	3.51951	4.44129
	T-T HEIGHT (FT)	10.4	10.4	10.6	12	10.4	10.6	12
	DESIGN T (F)	150	150	150	150	150	150	150
	DESIGN P (PSIG)	110	110	110	110	110	110	110
	VAPOR Kv	0.19	0.19	0.19	0.19	0.19	0.19	0.19
	VESSEL MATERIAL	CS	CS	CS	CS	CS	CS	CS
40.30	CONDENSER							
	TYPE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE	SHELL & TUBE
	DUTY (MMBTU/HR)	19.5513	19.5513	20.0071	31.2444	13.7071	13.7071	21.8273
	U ASSUMED (BTU/HR-FT ² -F)	75	75	75	75	75	75	75
	LMDT (F)	31.9146	31.9146	31.9146	31.9146	31.9146	31.9146	31.9146
	AREA (FT ²) + 20 % S.F.	9801.79	9801.79	10030.3	15664	6871.89	6871.89	10942.8
	SHELL DIAMETER (IN)	66	66	66	78	54	54	66
	SHELL LENGTH (FT)	19.7053	19.7053	20.1647	22.3567	20.9361	20.9361	21.9992
	TUBE LENGTH (FT)	19.7053	19.7053	20.1647	22.3567	20.9361	20.9361	21.9992
	NUMBER OF TUBES	1520	1520	1520	2141	1003	1003	1520
	TUBE OD (IN)	1.25	1.25	1.25	1.25	1.25	1.25	1.25
	COOLING WATER FLOW (GPM)	1896.03	1896.03	1940.24	3030	1329.28	1329.28	2116.75
	SHELL DESIGN T (F)	300	300	300	300	300	300	300
	SHELL DESIGN P (PSIG)	265	265	265	265	265	265	265
	SHELL MATERIAL	CS	CS	CS	CS	CS	CS	CS
	TUBES DESIGN T (F)	300	300	300	300	300	300	300
	TUBES DESIGN P (PSIG)	100	100	100	100	100	100	100
	TUBES MATERIAL	CS	CS	CS	CS	CS	CS	CS

TABLE 5: UTILITY SUMMARY

ELECTRIC USAGE SUMMARY

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
Equip. # Description							
01.10A Feed Gas Compressor - first stage	9,201	9,201	8,201	5,516	6,451	6,451	6,451
01.10B Feed Gas Compressor - second stage	9,302	9,302	8,291	5,576	6,521	6,521	6,521
Feed Gas Compressor - lube oil pump	143	143	127	86	100	100	100
Feed Gas Compressor - seal oil booster pump	143	143	127	86	100	100	100
Feed Gas Compressor -coalescer gear pump	6	6	5	3	4	4	4
01.20 Recycle Compressor	681	555	0	433	0	0	0
Recycle Compressor - lube oil pump	32	27	0	21	0	0	0
01.30 Vent Streams to Incinerator Compressor	74	16	15	20	39	39	39
01.40 Reduction N2 Compressor	268	266	263	344	675	675	675
Reduction N2 Compressor - lube oil pump	13	13	13	17	33	33	33
01.60 Instrument Air Compressor	377	377	377	377	377	377	377
Instrument Air Compressor - lube oil pump	18	18	18	18	18	18	18
02.83 02.10 Carbonyl Regen Heater	1	1	1	1	1	1	1
10.10 Purge Gas Expander	-5,672	-3,414	-3,185	-1,512	-2,953	-1,953	-1,953
Expander - lube oil pump	44	26	25	12	23	15	15
Expander - seal oil booster pump	44	26	25	12	23	15	15
Expander -coalescer gear pump	2	1	1	0	1	1	1
10.30 01.30 Incinerator Liquid Feed Pump	5	5	5	6	13	13	13
10.40 Fuel to Vaporizer Pump	57	60	33	33	0	0	28
10.50 Condensed Oil Circulation Pump	100	100	45	60	35	35	70
10.52 Emergency Slurry Return Pump	0	0	0	0	0	0	0
10.53 Utility Oil Pumps	48	48	47	62	122	122	122
10.54 Reduction Oil Pump	0	0	0	0	0	0	0
10.56 Fresh Oil Pump	0	0	0	0	0	0	0
10.58 Slurry Filtration Circulation Pump	2	2	2	3	5	5	5
10.59 Reduced Slurry Transfer Pump	0	0	0	0	0	0	0
10.60 Oil Makeup Pump	23	23	9	14	9	9	14
10.62 MECH to Storage/Recycle Pump	2	0	2	239	24	24	24
10.64 H2O Injection Pump	18	18	11	17	7	6	6
10.65 07.65 Column Reflux Pump	0	0	0	17	0	0	0
10.70 Offy Water Sump Pump	0	0	0	0	0	0	0
15.10 Incinerator Forced Draft Fan	50	11	10	14	27	27	27
15.40 Utility Oil Heater	369	369	365	477	935	935	935
21.53 Utility Oil Air Cooler	5	5	5	7	13	13	13
28.31 Slurry Mix Tank Agitator	20	20	20	26	51	51	51
28.35 Catalyst Bin Feeder	0	0	0	0	0	0	0
28.45 Slurry Centrifuge	56	56	56	64	89	89	89
28.61 Slurry Tank Agitator	37	37	37	48	63	63	63
30.00 Sulfur Removal Unit	432	432	385	259	303	303	303
40.00 Refrigeration Skid Compressor	2,914	2,914	2,982	4,656	2,043	2,043	3,253
Building HVAC	30	30	30	30	30	30	30
EQUIPMENT TOTAL (HP)	18,822	20,836	18,347	17,039	15,170	16,164	17,441
EQUIPMENT TOTAL (KW)	14,036	15,537	13,682	12,706	11,319	12,053	13,006
Add 4 % for Transformer Losses, Lights, , etc	561	621	547	508	453	482	520
TOTAL - KILOWATTS	14,597	16,159	14,229	13,214	11,772	12,535	13,526

96

TABLE 5: UTILITY SUMMARY

		CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
STEAM BALANCE SUMMARY (LBS/HR)								
Stream		Op Avg	Op Avg	Op Avg	Op Avg	Op Avg	Op Avg	Op Avg
INLET STREAMS (785 PSIG, 240 F)								
402	BFW TO STEAM DRUM	77,898	80,096	101,754	98,193	112,220	112,287	112,298
432	BFW FOR WATER INJECTION	25,496	25,496	14,936	22,571	8,990	8,847	8,847
	TOTAL BFW IMPORT	103,394	105,592	116,690	120,764	121,210	121,134	121,145
STEAM USAGE BY SECTION								
420	07.73/07.77 REBOILERS	5,222	17,402	15,421	27,394	5,534	13,915	16,084
423	STEAM EXPORT	72,132	62,133	85,620	70,111	105,900	97,586	95,428
409	TOTAL STEAM AVAILABLE	77,354	70,535	101,041	97,505	111,434	111,501	111,512
OUTLET STREAMS								
401	BLOWDOWN FROM STEAM DRUMS	545	561	712	687	785	786	786
	RETURN CONDENSATE FR REBOILERS	5,222	17,402	15,421	27,394	5,534	13,915	16,084
423	STEAM EXPORT (SAT'D AT 230 PSIG)	72,132	62,133	85,620	70,111	105,900	97,586	95,428
	TOTAL RETURN STREAMS	77,899	80,096	101,753	98,192	112,219	112,287	112,298
IMPORTED STEAM TO VAPORIZE FUEL								
450	FLOW	46,274	48,089	22,373	23,379	0	0	10,336
	PRESSURE (PSIG)	230	230	230	230	0	0	230
L6	COOLING WATER SUMMARY (GPM)							
Stream								
520	07.72/07.76 CONDENSERS	100	0	0	970	0	267	383
504	21.31 METHANOL PRODUCT CONDENSER	6,370	7,865	2,908	3,952	1,079	2,012	2,011
518	21.01 - 01.10 COMPRESSOR INTERCOOLER	2,330	2,330	2,077	1,397	1,634	1,634	1,634
	21.60 - 01.60 INSTR AIR COMP COOLERS	91	91	91	91	91	91	91
	21.41 - 01.40 RED'N COMP COOLERS	44	44	43	57	111	111	111
524	21.71/07.74 LIQUID PRODUCT COOLER	244	1,265	309	328	0	0	854
528	21.72 LIQUID RECYCLE COOLER	0	0	506	0	0	0	0
	30.10 PROPANE CONDENSER	1,896	1,896	1,940	3,030	1,329	1,329	2,117
	TOTAL PROCESS C.W. USE	11,075	13,491	7,875	9,826	4,244	5,444	7,201
	TOTAL COOLING TOWER DUTY (MMBTU/HR)	111	135	79	98	43	55	72

TABLE 6: SUMMARY OF PLANT AND PRODUCTION COSTS FOR CASES 1-7

	CASE 1	CASE 2	CASE 3	CASE 4	CASE 5	CASE 6	CASE 7
PRODUCTION							
Mass Rate (TPD)	828	876	668	665	622	590	579
Energy Rate (MM BTU/hr, LHV)	635	678	618	620	645	628	602
CAPITAL COMPONENT (\$1,000 in 1992)							
Feed Compression & Gas Pretreatment	16,439	16,269	14,949	11,633	12,319	12,119	12,269
Basic Synthesis Section	23,575	24,171	17,531	22,308	19,891	20,716	19,883
Extra Synthesis Equipment	0	1,410	2,576	6,149	0	1,654	2,400
Slurry Reduction	1,273	1,260	1,270	1,494	2,706	2,662	2,536
Spent Catalyst Recovery	197	195	197	230	408	401	382
Storage Loading and Vaporization	3,003	3,018	7,795	7,692	0	0	8,569
General Utilities	1,018	1,015	1,089	1,057	1,538	1,513	1,442
Indirect Cost	17,185	17,759	17,116	18,739	14,476	15,191	17,837
Initial Catalyst Charge	1,449	1,451	1,130	1,463	1,812	1,811	2,144
TOTAL	64,139	68,548	63,653	70,763	53,150	56,067	67,462
UNIT PRODUCT COST (\$/MM BTU, LHV)							
Syngas	5.13	5.01	5.07	5.06	5.09	5.24	5.15
Steam Export	-0.13	-0.07	-0.32	-0.24	-0.52	-0.50	-0.45
Electric Power	1.05	1.10	1.06	0.97	0.81	0.91	1.00
Cooling Water	0.10	0.12	0.08	0.10	0.04	0.05	0.07
Make-up Oil	0.04	0.04	0.02	0.03	0.04	0.05	0.05
Catalyst Supply	0.29	0.27	0.29	0.38	0.72	0.76	0.77
Adsorbent	0.01	0.01	0.00	0.00	0.00	0.00	0.01
Catalyst/Oil Disposal	0.04	0.04	0.04	0.05	0.09	0.09	0.09
Adsorbent Disposal	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Operating Labor	0.09	0.09	0.10	0.10	0.09	0.10	0.10
Plant Overhead	0.19	0.18	0.19	0.19	0.19	0.20	0.20
Maintenance	0.24	0.24	0.25	0.27	0.20	0.22	0.27
Taxes and Insurance	0.24	0.24	0.25	0.27	0.20	0.22	0.27
Capital Cost	1.70	1.66	1.73	1.92	1.39	1.54	1.89
TOTAL	9.01	8.93	8.76	9.10	8.34	8.88	9.42
Unit Cost (\$/lb)	0.083	0.083	0.097	0.102	0.104	0.113	0.118