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PRODUCTION OF JET FUELS FROM COAL DERIVED LIQUIDS

VOL X -- Jet Fuels Production By-Products, Utility,
and Sulfur Emissions Control Integration Study

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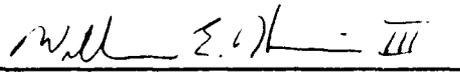
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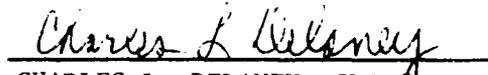
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In September 1986, the Fuels Branch of the Aero Propulsion Laboratory at Wright-Patterson Air Force Base, Ohio, commenced an investigation of the potential of jet fuel production from the liquid by-product streams produced by the gasification of lignite at the Great Plains Gasification Plant (GPGP) in Beulah, North Dakota. Funding has been provided by the Department of Energy (DOE) Pittsburgh Energy Technology Center (PETC) to administer the experimental portion of this effort. This document reports the results of the effort by Burns and Roe Services Corporation/Science Applications International Corporation (BRSC/SAIC) to evaluate the impact of integrating Jet Fuel and/or Chemical Production Facilities with the Great Plains Gasification Plant.

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FOREWORD

In September 1986, the Fuels Branch of the Aero Propulsion Laboratory at Wright-Patterson Air Force Base, Ohio commenced an investigation of the potential of production of jet fuel from the liquid by-product streams produced by the gasification of lignite at the Great Plains Gasification Plant in Beulah, North Dakota. Funding was provided to the Department of Energy (DOE) Pittsburgh Energy Technology Center (PETC) to administer the experimental portion of this effort. This report details the effort of Burns and Roe Services Corporation/Science Applications International Corporation (BRSC/SAIC), who, as a contractor of DOE (DOE Contract No. DE-AC22-87PC79338), was requested to evaluate the impact of integrating Jet Fuel and/or Chemical Production Facilities with the Great Plains Gasification Plant. DOE/PETC was funded through Military Interdepartmental Purchase Request (MIPR) FY1455-86-N0657. Mr. William E. Harrison III was the Air Force Program Manager and Mr. Gary Stiegel was the DOE/PETC Program Manager.

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1.0 Executive Summary

Burns and Roe Services Corporation/Science Applications International Corporation (BRSC/SAIC) have been requested by the Department of Energy (DOE) and Department of Defense (DOD) to evaluate the impact of integrating proposed Jet Fuels/Chemicals production facilities with the Great Plains Gasification Plant (GPGP). In addition to synthetic natural gas (SNG), the GPGP also produces three liquid by-products: naphtha, crude phenol, and tar oil. Currently, the by-product are burned in the plant's boilers and super-heaters to produce steam. However, these by-products also represent a potential source of revenue for the plant if they can be processed to pro-duce marketable products. Consequently, the Department of Energy and Department of Defense have conducted a program to evaluate the economic feasibility of producing marketable products and military jet fuels at the GPGP. Analysis of the technical and economic feasibility of producing military jet fuel (JP-4, JP-8, and JP-8X) and chemicals (phenol, cresols, and BTX) by Amoco Oil Co. and Lummus Crest, Inc. nas identified two alternates meriting further evaluation: (1) production of JP-8 subsidized by chemicals production (Amocc/LCI Case 4) and (2) production of chemicals only (Amoco/LCI Case 7).

The objectives of this evaluation are to: (1) determine the economic impact of replacement fuel and equipment costs on the economic viability of these two alternates; (2) identify replacement fuels for the GPGP boilers; (3) select a replacement fuel which minimizes SO_2 emissions; (4) determine what, if any, impact the integration of the Jet Fuels/Chemicals facility will have on utilities, waste stream cleanup, and SNG production facilities; and (5) develop a preliminary design for the equipment necessary to inte-grate these facilities.

The evaluation indicates that conversion of the GPGP by-products to JP-8 and chemicals (Alternate 1) via hydroprocessing offers a significant credit for SO_2 emission reductions provided a low-sulfur fuel is used. Depending on the fuel selected, Medora or Fryburg crudes or a blend of the two, overall plant SO_2 can be reduced emissions by 7 to 14%. SNG would be an even more

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attractive replacement fuel from this viewpoint and could reduce SO_2 emissions by 22% in Alternate 1 and 9% in Alternate 2. However, SNG sells for about 25% more than the crude oil price and thus is too valuable to be used as plant fuel. Although replacement fuel receipt, storage, and handling represent the major portion of the capital cost required for integration with the GPGP, these are of minor significance when expressed as operating costs and compared to the replacement fuel cost. Charges for the replacement fuel represent approximately 99% of the cost of integration, which is $3.05/MMBtu for Alternate 1 and $3.23/MMBtu for Alternate 2 and reflects a $17/bbl delivered cost for the replacement fuel. This is slightly above the upper limit of $3.00/MMBtu identified by Amoco/LCI as required to produce a 10% real rate of return. Consequently, any increase in replacement fuel cost will adversely affect the economic feasibility of producing jet fuel from the GPGP by-product and every effort should be made to either negotiate a lower price or locate a lower-cost fuel source. At the same time, use of low-sulfur replacement fuel offers a significant credit for SO_2 emissions reduction. Although determination of the quantitative cost advantage associated with a reduction in SO_2 emission was beyond the scope of this evaluation, any decision regarding the construction of the Jet Fuels and/or Chemicals Production Facilities will need to take this into account.

Finally, BRSC/SAIC and ANG have identified preferred integration locations for the process intermediate offgas and waste water streams. With the exception of requiring two additions, i.e., new cooling towers and gas treating facilities to remove H_2S, integration of the Jet Fuels/Chemicals Production Facilities will have only minor impacts on GPGP operations. In fact, ANG analysis has determined that most of the streams returned from the Jet Fuels Chemicals Facility to the GPGP are too small to be measured on the existing instrumentation and analysis equipment.

2.0 Introduction

At the request of the U.S. Department of Energy and the U.S. Air Force (USAF), BRSC/SAIC and ANG have reviewed and evaluated the impact of inte-

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grating proposed facilities for the production of jet fuels and/or chemicals with those of the Great Plains Gasification Plant (GPGP). Jet fuel production from the GPGP by-product naphtha, crude phenol, and tar oil is under investigation as a secure source of these jet fuels for U.S. Air Force bases in North Dakota. Analysis of the technical and economic feasibility of producing military aviation turbine fuel (JP-4, JP-8, and JP-8X) and chemicals (phenol, cresols, and BTX) has been conducted by an Amoco Oil Co./Lummus-Crest, Inc. (Amoco/LCI) team.

Amoco/LCI developed preliminary material balances, final product and process intermediate rates, utility requirements, and investments for six specific scenarios producing jet fuels either with or without coproduct chemicals production. They are as follows:

    Case 1 - Maximum JP-4 Production

    Case 2 - Profitable JP-4 Production

    Case 3 - Maximum JP-8 Production

    Case 4 - Profitable JP-8 Production

    Case 5 - Maximum JP-8X Production

    Case 6 - Profitable JP-8X Production

A seventh scenario was also evaluated - a maximum profitability case based on the production of chemicals alone. Subsequent analysis of these seven cases by Amoco/LCI has identified two cases that are particularly attractive to DOE and the USAF as profitable and technically feasible options, and are referred to in the Amoco/LCI Task 1 evaluation as Case 4-Profitable JP-8 production, and Case 7-Maximum Profitability, respectively, and are the focus of this study.

Since Amoco/LCI's evaluation dealt solely with the technical and economic evaluation of the jet fuels/chemicals production facilities, it was necessary to determine whether such facilities would still be attractive when the cost of integrating them with the GPGP was included. Therefore, an investigation into the effect of constructing and integrating a jet fuels/chemicals facility on the GPGP utilities and emissions, and overall jet

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fuels production economics was required. This study focuses on identifying and quantifying these effects by determining;

o   If new utility capacity is required.

o   Whether process intermediate streams should be treated and if so, how.

o   Whether diverting GPGP synthesis gas to H_2 production affects downstream unit operations and SNG production.

Modifications required to satisfactorily address these issues could then be designed and cost estimated to provide a basis for assigning an economic value to the effects of integrating a jet fuels/chemicals production facility. This would then permit a more thorough assessment of the impact of integrating jet fuels/chemicals facilities will have on overall process economics.

This study also considered the impact of diverting GPGP by-products (presently used as plant fuel) on operating cost and SO_2 emissions. Tar oil, crude phenol, and naphtha are currently used to fuel the plants' boilers, superheaters, and liquid waste incinerator. However, the two alternatives under consideration divert either all or a substantial portion of these by-product liquids to jet fuels and/or chemicals production. Consequently, DOE and DOD requested BRSC/SAIC include as part of this evaluation the identification of potential replacement fuels and evaluate the effects of their use on jet fuels/chemicals production costs and the overall GPGP SO_2 emission levels.

Finally, this study also reflects the impact of changes to the GPGP which have either recently been implemented or are likely to be implemented in the near future. Examples are the 1988 re-rating of GPGP SNG production capacity from 137.5 to 152.5 MMSCF/SD, identification of proposed Best Available Control Technology (BACT) to reduce current GPGP SO_2 emissions, and installation of a booster compressor to deliver SNG at pipeline

pressures of 1400-1450 psi. Since availability of plant utilities will be affected by these modifications, it is imperative that these changes be included in the study basis. Furthermore, this provides a realistic, up-to-date basis for evaluation.

3.0 Goals and Objectives

The objective of this study is to evaluate the impact of integrating a jet fuels/chemicals or chemicals production facility with the existing GPGP both from a technical as well as an economic perspective.

A second objective is to evaluate the impact of using replacement boiler fuels such as SNG and crude oil on plant economics and SO_2 emissions.

Factors which were considered in selecting a replacement fuel were cost, availability of supply, ease of use, and sulfur content. Since the proposed BACT is aimed at bringing the plant into compliance with EPA SO_2 emission regulations, the replacement fuel selection process specifically included evaluation of SO_2 emissions and the cost of any equipment required to achieve compliance.

Subobjectives of this study include:

o   Identification of utility imbalances which require the design and construction of new utility capacity.

    Identification of facilities required to treat and transfer process intermediates and waste streams (i.e., treat gas, sour water).

o   Identification of preferred return locations and conditions (temperature, pressure) for intermediate streams.

o   Development of specifications for stream contaminants such as H_2S so treatment schemes could be identified and designed.

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o   Evaluation of the impact diverting intermediate GPGP streams to the jet fuels/chemicals plant on GPGP operations.

o   Quantification of the additional costs incurred when process intermediate and waste streams from the jet fuels/chemicals production facilities are treated in the existing GPGP facilities.

This study addresses the impact of these and similar changes on GPGP and jet fuels/chemicals operations both in terms of added investment and operating costs imposed on the jet fuels facility.

4.0 Evaluation Basis

The first priority in conducting this evaluation was to develop a statement of work clearly defining the objectives of this study and the basis for the evaluations. Once developed, this statement of work was reviewed, modified, and approved by the program participants at the May 25-26, 1988 Project Review Meeting at which time it was agreed that the study would be based on a SNG production rate of 152.5 MMSCF/SD and that the liquid by-product production rates would be consistent with this SNG rate. Likewise, the base case would assume that the BACT proposed in the "Application for Major Source Modification and Permit to Operate for the Great Plains Coal Gasification Facility" (Herein referred to as "Application") (1) would be adopted and implemented prior to construction and operation of the jet fuels/ chemicals production facility. In addition, background information on by-product rates and plant SO_2 emissions would also be extracted from the "Application" to ensure that the evaluation would accurately reflect planned future operation of the GPGP.

Instead of using by-product elemental composition data (C/H/O/N/S) from the "Application", elemental compositions from the recently completed compilation and review of GPGP by-product analyses (2) were used at the request of DOD and DOE. While this results in somewhat lower tar oil and naphtha sulfur concentrations than projected for future operations as shown in Table 1

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TABLE 1 - HISTORICAL VERSUS PROJECTED FUTURE SULFUR CONTENTS OF GPGP LIQUID
BY-PRODUCTS

	Sulfur Content, Wt %		
	Historical	Amoco/LCI	Projected
Naphtha	1.2 ± 0.4	1.7	2.1
Crude Phenol	0.1 ± 0.0	0.1	0.8
Tar Oil	0.5 ± 0.1	0.4	

the overall effect on plant SO_2 emissions is quite small compared to other SO_2 sources as shown in Table 2.

Next, three specific alternates were identified for evaluation and are summarized as follows:

Alternate 1:   Profitable production of JP-8 and chemicals from liquid by-products produced at the GPGP. (Based on Amoco/LCI Case 4 - Profitable JP-8 Production.)

Alternate 2:   Production of chemicals only from the GPGP naphtha and crude phenol. Tar oil remains in the GPGP fuel pool. (Based on Amoco/LCI Case 7 - Maximum Profitability.)

Alternate 3:   Alternate 1 evaluated using SNG as the replacement fuel.

As indicated, these are based on the results of the Amoco/LCI Task 1 preliminary process design and economic screening (3,4), which indicates that these are technically feasible and economically profitable options for processing the GPGP by-products.

Naphtha, crude phenol, and tar oil production rates associated with the base GPGP SNG production rate of 152.5 MMSCF/SD were extracted from Table 4-7 of the Permit Application and are summarized in Table 3. Using this information as a starting point, hydrotreater and hydrocracker off-gas stream compositions were modified to reflect the historical elemental analysis of the naphtha, crude phenol, and tar oil. As explained in the Statement of Work, actions taken to control SO_2 emissions at the GPGP will be in accordance with the proposed BACT. Consequently, the evaluation of replacement fuels on SO_2 emissions will use data from Table 4-7 of the Permit Application as a baseline. Evaluation of the effect of incremental utility requirements on the GPGP has been based on the prior installation of a compressor to boost SNG to a pipeline pressure of 1400-1450 psig and assumes that this compressor will use an electric drive, thereby making 550 psig steam available for use elsewhere in the GPGP and the jet fuels facility. Finally, the economic

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TABLE 2 - COMPARISON OF HISTORICAL AND PROJECTED BY-PRODUCT SULFUR EMISSIONS
AT GPGP

	SO_2 Emissions, lbs/hr	
	Historical	Projected
Other GPGP Emission Sources		
Rectisol Offgas	2,450	2,450
Fuel Gas	248	248
Phosam Offgas	80	80
Subtotal	2,778	2,778
Naphtha Crude Phenol	231	411
Tar Oil	479	753
Subtotal	710	1,164
Total SO_2 Emissions	3,488	3,942

TABLE 3 - GPGP LIQUID BY-PRODUCT BASIS (Based on 152.5 MMSCFD SNG)

Naphtha	8,519 lb/hr
Crude Phenol	13,550 lb/hr
Tar Oil	47,910 lb/hr

evaluation of integrating these facilities with the GPGP is based on data extracted from studies by J.E. Sinor Consultants, Inc. (5) and Amoco/ LCI (6).

5.0 Approach

In order to assess the cost of integrating the jet fuel and/or chemical facilities with the existing GPGP, product, by-product, and process inter- mediate streams had to be identified and preferred dispositions defined. Similarly, replacement fuel demand had to be quantified in order to form a basis for evaluating candidate replacement fuels. This was particularly critical with regards to quantifying the potential SO_2 emissions contribu- tions for each of the candidates. Incremental utility demands associated with the jet fuels/chemicals facility also had to be defined so BRSC/SAIC and ANG could determine whether or not sufficient capacity to meet these demands existed at GPGP and to what extent new capacity would be required. Finally, a basis for product, by-product, and process utility costs was defined to permit assessment of the economic impact of integrating these facilities with the GPGP.

6.0 Analysis

6.1 By-Product Flowrate Definition

Using the results of Amoco/LCI's "Preliminary Analysis of Upgrad- ing Alternatives for Great Plains Liquid By-Product Streams" (7) and Task 1 "Process Design and Cost Estimate" (8) as a basis, BRSC/SAIC developed product, process intermediate, by-product, and utility flows for alternates 1 and 2. All stream rates and utili- ties were adjusted to be consistent with the naphtha, crude phenol, and tar oil rates contained in the "Application", which have been presented in Table 3. Consequently, the feed rates to the Jet Fuels/Chemicals facilities were revised as follows:

o   Rectisol Naphtha, from 725 BPSD to 709 BPSD

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o   Crude Phenol, from 936 BPSD to 871 BPSD

o   Tar Oil, from 3182 BPSD to 3232 BPSD

As indicated, the adjustments were modest and should have minimal impact upon the conclusions Amoco/LCI reached in their evaluation.

6.2 By-Product Sulfur and Nitrogen Content Definition

Somewhat more significant was the adjustment of the feedstock sulfur and nitrogen contents. As reported in the May-August 1988 Interim Progress Report (10), the program participants agreed to use nitrogen and sulfur compositions extracted from the Feed Analysis Compilation and Review, AFWAL-TR-87-2042 Vol. VI, which summarized the results from analysis conducted for the naphtha, crude phenol and tar oil during the jet fuels program. As shown in Table 4, the historical basis and the Amoco/LCI basis are not appreciably different except with regards to naphtha sulfur and tar oil nitrogen content. Nonetheless, when all three by-product streams are evaluated, the total sulfur contents in pounds per hour are virtually the same (Table 5). However, overall Amoco/LCI nitrogen content is about 28% higher than calculated from the historical basis. Should these higher nitrogen contents be realized, there would be a definite effect on either NH_3 (from hydrotreating/hydrocracking in Alternate 1) or NO_x production (from combustion in Alternate 2). Although it ultimately depends on the economics and NO_x emission regulations, hydroconversion would be the preferred option, removing the excess nitrogen as salable NH_3. The projected by-product sulfur content is of more concern since it is about 65% higher than the historical levels, primarily because the remaining GPGP coal reserves will contain more sulfur than that already processed. However, as shown in Table 2, this 65% increase in by-product sulfur increases overall plant SO_2 emissions by approximately 455 lbs/hr or 13%. Depending on the processing options chosen, this sulfur can either be

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TABLE 4 - COMPARISON OF GPGP BY-PRODUCT NITROGEN AND SULFUR CONTENTS

	Amoco/LCI Basis	Historical Basis	Projected Basis
Rectisol Naphtha			
Nitrogen	0.2 wt%	0.3 wt%	N/A
Sulfur	1.7	1.2	2.1
Crude Phenol			
Nitrogen	0.5	0.4	N/A
Sulfur	0.1	0.1	0.21
Tar Oil			
Nitrogen	0.8	0.6	N/A
Sulfur	0.4	0.5	0.8

TABLE 5 - COMPARATIVE NITROGEN AND SULFUR FLOWS FOR GPGP BY-PRODUCTS

	Amoco/LCI Basis	Historical Basis	Projected Basis
Rectisol Naphtha			
Nitrogen	17.5 lb/hr	25.6 lb/hr	N/A lb/hr
Sulfur	148.5	102.2	178.9
Crude Phenol			
Nitrogen	72.5	54.2	N/A
Sulfur	14.5	13.6	28.5
Tar Oil			
Nitrogen	381.0	287.4	N/A
Sulfur	190.4	239.6	383.2
Total Nitrogen	471.0	367.2	N/A
Total Sulfur	353.4	355.4	590.6

removed as H_2S and recovered as sulfur or offset through the use of a low-sulfur replacement fuel.

6.3 Process Intermediate and Waste Stream Definition

Following adjustment of the naphtha, crude phenol, and tar oil sulfur and nitrogen contents, the process flow diagrams for Alternate 1 and Alternate 2 were revisea to reflect the updated flow rates and compositions. These material balanced flow diagrams, shown in Figures 1 and 2, respectively, provided the basis for determining the quantities of purge gas, fuel gas, sour water, and process wastewater to be treated and processed in the GPGP. The amount of hydrogen, process water, and various utilities required by the jet fuels and/or chemical production facilities was then quantified based on these material balances.

Because of the hydroprocessing associated with these facilities, makeup hydrogen must be provided. Consequently, Amoco/LCI included a Pressure Swing Adsorption (PSA) unit in their design to strip hydrogen from the synthesis gas stream leaving the Rectisol Unit. The hydrogen is sent to the appropriate processing units while the remaining hydrogen-deficient synthesis gas is recompressed and sent back to the GPGP. If the remaining synthesis gas stream is large, as is the case in Alternate 1, it can be blended with the main synthesis gas stream and fed to methanation. If the stream is small, as is the case with Alternate 2, the H_2-deficient synthesis gas can be used as plant fuel and eliminate the need for recompressing the gas back to the methanator feed inlet pressure. In order to permit ANG to determine tne impact of removing hydrogen from the synthesis gas on SNG production and operation of related process facilities, a material balance was conducted around the PSA unit. These material balances are shown in Figure 3 (Alternate 1) and Figure 4 (Alternate 2) and include the composition, temperature, pressure, and flow rates for the synthesis gas, purge gas, and makeup hydrogen streams.
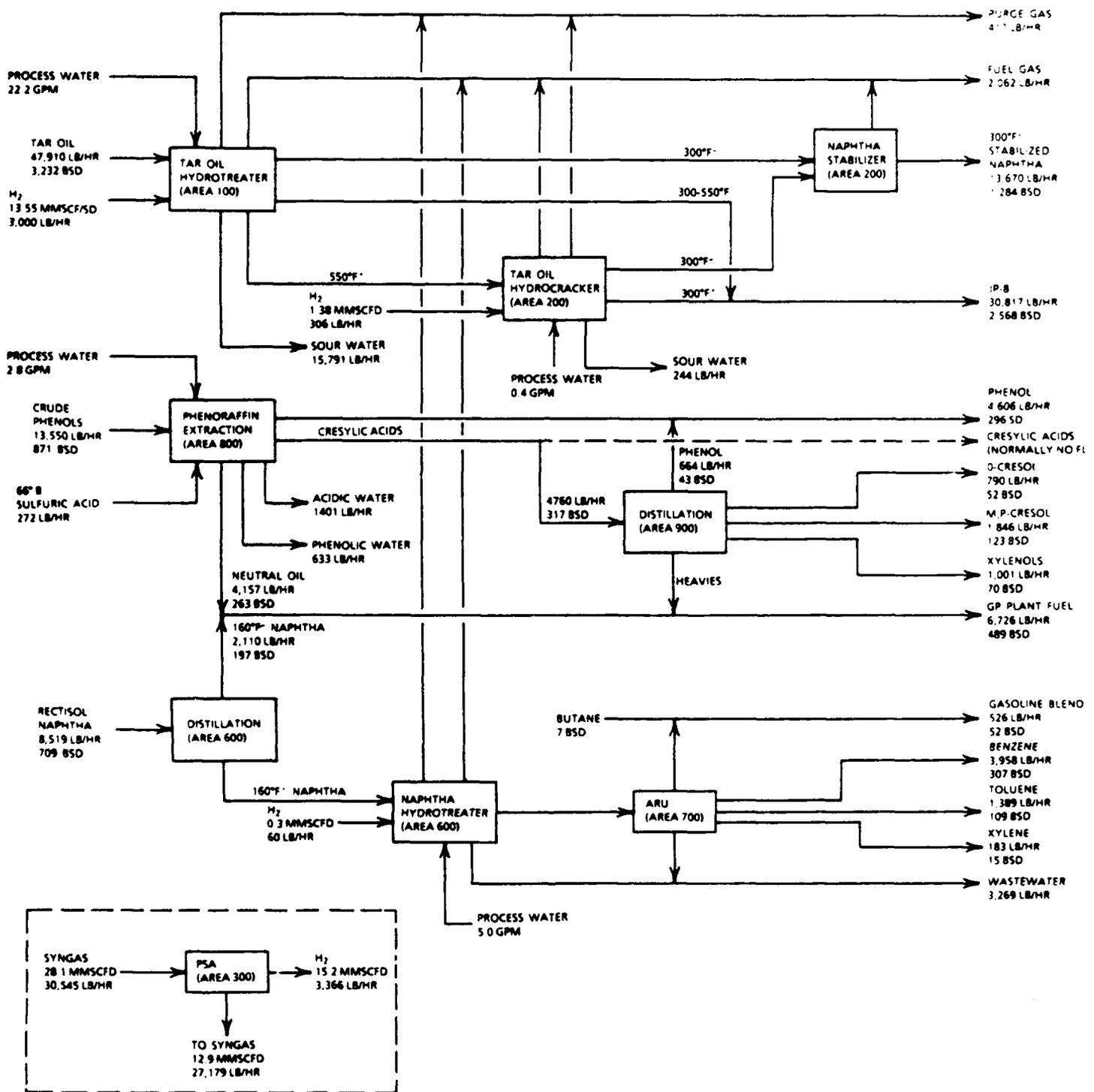
FIGURE 1. MATERIAL BALANCE FOR ALTERNATE 1 – PROFITABLE JP-8 PRODUCTION
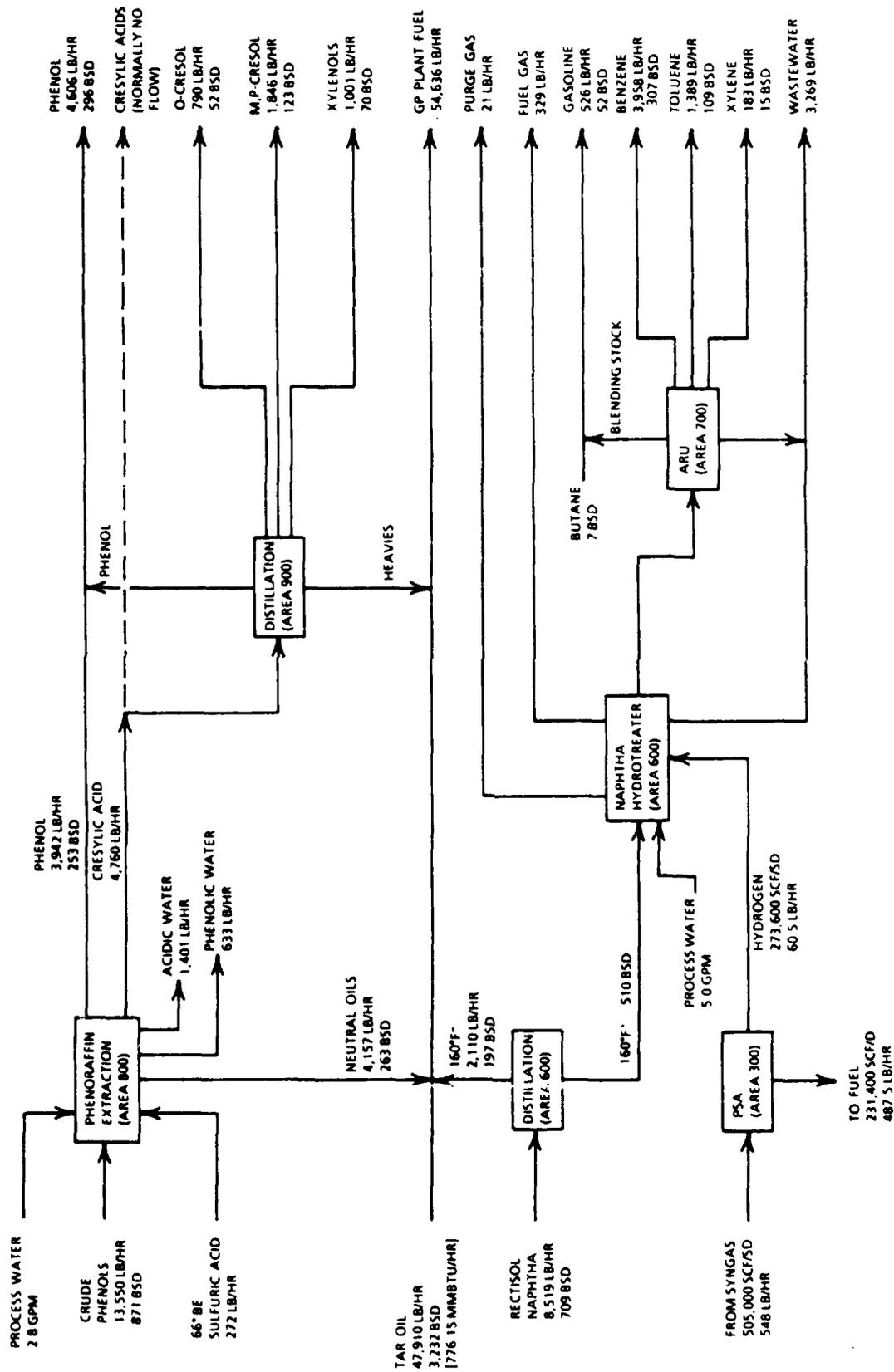
FIGURE 2. MATERIAL BALANCE FOR ALTERNATE 2 – MAXIMUM PROFIT
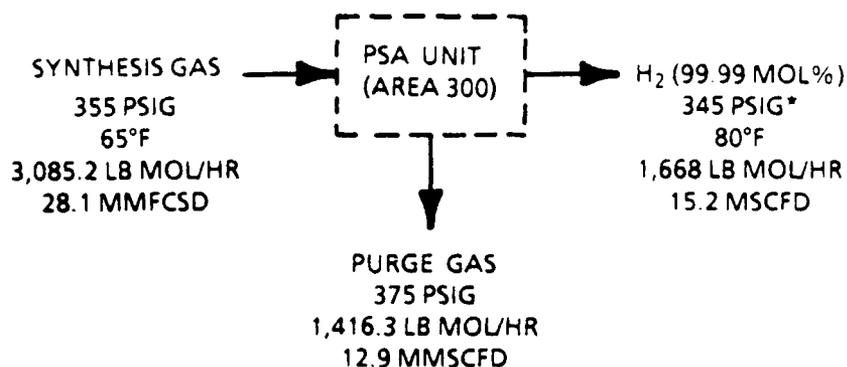
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SYNTHESIS GAS
355 PSIG
65°F
3,085.2 LB MOL/HR
28.1 MMFCSD

PSA UNIT
(AREA 300)

H₂ (99.99 MOL%)
345 PSIG*
80°F
1,668 LB MOL/HR
15.2 MSCFD

PURGE GAS
375 PSIG
1,416.3 LB MOL/HR
12.9 MMSCFD

*PURGE GAS RECOMPRESSION INCLUDED

STREAM	SYNTHESIS GAS 65°F 355 PSIG		PURGE GAS 375 PSIG		H₂ 80°F 345 PSIG	
	MOL %	M/H	MOL %	M/H	MOL %	M/H
H₂	63.19	1949.5	19.32	273.6	99.99	1668.7
CO	18.61	574.2	40.76	527.3	0.00	0.0
CO₂	1.48	45.7	3.24	45.9	0.00	0 0
CH₄	16.21	500.1	35.51	502.9	0.00	0.0
C₂H6	0.31	9.6	0.69	9.8	0.00	0 0
COS, H₂S, CS2	0.00	0.0	0.00	0.0	0.00	0.0
N₂ + Ar	0.19	3.9	0.41	0.8	0.01	0.2
H₂O	0.00	0.0	0.06	0.8	0.00	0.0
	99.99	3084.96	99.99	1416.2	100.00	1668.9

FIGURE 3. HYDROGEN RECOVERY UNIT - ALTERNATE 1 -
MATERIAL BALANCE AND STREAM COMPOSITIONS
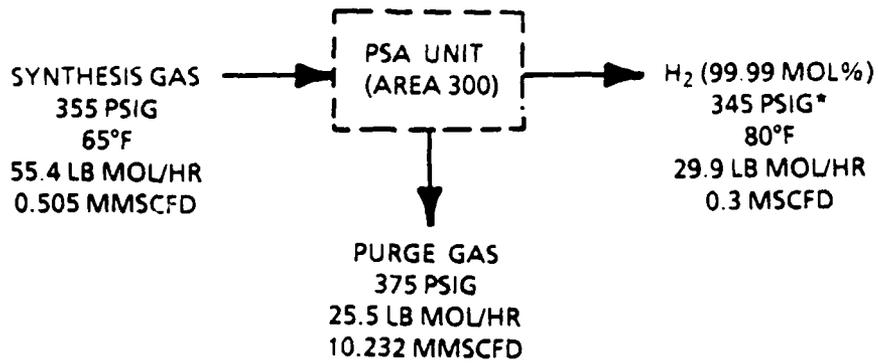
SYNTHESIS GAS → PSA UNIT (AREA 300) → H₂ (99.99 MOL%)

SYNTHESIS GAS
355 PSIG
65°F
55.4 LB MOL/HR
0.505 MMSCFD

PSA UNIT
(AREA 300)

H₂ (99.99 MOL%)
345 PSIG*
80°F
29.9 LB MOL/HR
0.3 MSCFD

PURGE GAS
375 PSIG
25.5 LB MOL/HR
10.232 MMSCFD

*PURGE GAS RECOMPRESSION INCLUDED

STREAM	SYNTHESIS GAS 65°F 355 PSIG		PURGE GAS 375 PSIG		H₂ 80°F 345 PSIG	
	MOL %	M/H	MOL %	M/H	MOL %	M/H
H₂	63.19	35.0	19.32	4.9	99.99	30.0
CO	18.61	10.3	40.76	10.4	0.00	0.0
CO₂	1.48	0.8	3.24	0.8	0.00	0.0
CH₄	16.21	9.0	35.51	9.0	0.00	0.0
C₂H6	0.31	0.2	0.69	0.2	0.00	0.0
COS, H₂S, CS2	0.00	0.0	0.00	0.0	0.00	0.0
N₂ + Ar	0.19	0.1	0.41	0.1	0.00	0.0
H₂O	0.00	0.0	0.06	0.0	0.00	0.0
	99.99	55.4	99.99	25.4	100.00	30.0

FIGURE 4. HYDROGEN RECOVERY UNIT - ALTERNATE 2 - MATERIAL BALANCE AND STREAM COMPOSITIONS

Additional information describing the finished and intermediate products is contained in Tables 6 through 9 for Alternate 1, and Tables 10 through 13 for Alternate 2. For example, Tables 6 and 10 report total stream mass and volume flow rates and the disposition of the intermediate product streams for Alternates 1 and 2 respectively. Likewise, compositions, conditions, and flowrates for the purge and fuel gas streams are presented in Tables 7 and 8 (Alternate 1) and 11 and 12 (Alternate 2). Finally, Tables 9 and 13 provide the composition, conditions, and flowrates for the process sour water and waste water streams for the respective cases. While these are not rigorous descriptions of all the possible compounds in these streams, this provides a first-pass estimate of what GPGP would have to handle. A more detailed evaluation of the process stream contaminants would likely be required if and when the decision was made to proceed with construction of these facilities.

6.4 Replacement Fuel Requirements

The impact integrating the Jet Fuels/Chemicals production facility will have on the GPGP SO_2 emissions will be highly dependent on the quantity of fuel required to replace those by-products consumed in the production of jet fuels and/or chemicals. A small, but still somewhat significant additional fuel demand will also be imposed by the Jet Fuels/Chemicals production facilities which require fuel directly, for process heaters, and indirectly, in the form of incremental steam requirements. Consequently, overall replacement fuel requirements were calculated for Alternates 1 and 2. In order to provide a basis for evaluating the effect replacement fuel selection has on overall GPGP SO_2 emissions, SO_2 emissions were adjusted for the amount of sulfur contained in the diverted by-products and the sulfur contained in the process intermediates derived from the processing of the by-products. These include the $160^{\circ}F^-$ naphtha, neutral oil/heavies from Phenoraffin extraction of the crude phenol and cresylic acid

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TABLE 6 - ALTERNATE 1 FINISHED AND INTERMEDIATE PRODUCT RATES AND
              DISPOSITIONS

FINISHED PRODUCT RATES

PRODUCT	BBL/SD	LB/HR	LB/GAL
STABILIZED NAPHTHA TO CATALYTIC REFORMING	1,210	12,882	6.08
GASOLINE BLEND			
- FROM NAPHTHA STAB.	74	788	6.08
- FROM ARU	52	526	5.78
TOTAL =	126	1,314	5.96
JP-8	2,568	30,817	6.86
PHENOL	296	4,606	8.89
o-CRESOL	52	790	8.68
m,p-CRESOL	123	1,846	8.58
XYLENOLS	70	1,001	8.17
BENZENE	307	3,958	7.37
TOLUENE	109	1,389	7.28
XYLENE	15	183	6.97
TOTAL FINISHED PRODUCT	4876	58,786	

INTERMEDIATE PRODUCT RATES

PRODUCT	BBL/SD	LB/HR	DISPOSITION
160 F- NAPHTHA	197	2,110	Boiler Fuel
NEUTRAL OIL (from PHENOL EXTRACTION)	263	4,157	Boiler Fuel
HEAVIES (from PHENOL DISTILLATION)	29	459	Boiler Fuel
PURGE GAS (from UPGRADING)	50,853 (SCF/HR)	417	Rectisol Unit feed
OFFGAS (from UPGRADING)	38,330 (SCF/HR)	2,062	Boiler Fuel (following H2S removal)
PSA PURGE GAS	537,448 (SCF/HR)	27,251	Methanation Unit feed

TABLE 7 - ALTERNATE 1 PURGE GAS STREAM DATA

STREAM		TAR OIL HYDROTREATER PURGE GAS	TAR OIL HYDROCRACKER PURGE GAS	NAPHTHA HYDROTREATER PURGE GAS	TOTAL PURGE GAS FLOWRATE
TEMPERATURE	F	120	120	120	
PRESSURE	PSIA	1890	1095	710	
FLOWRATE	LB/HR	54.9	341.3	20.4	416.6
	LB MOL/HR	16.07	116.80	0.78	133.65
	SCFH	6100.2	44323.7	297.0	50721.0
AVG MW	LB/LB MOL	3.42	2.92	26.09	3.12
COMPONENT	LB/HR				
H2O		0.3	3.2	0.2	3.7
NH3		0.1	0.0	8.4	8.5
H2		30.1	230.2	0.0	260.3
H2S		0.0	0.0	3.3	3.3
C1		12.9	11.3	0.7	24.8
C2		6.1	4.2	0.8	11.1
C3		2.9	19.9	0.6	23.4
iC4		0.5	11.6	0.0	12.1
nC4		0.1	33.6	3.3	37.0
C5+		1.9	27.3	3.2	32.4
TOTAL	LB/HR	54.9	341.3	20.4	416.6

TABLE 8 - ALTERNATE 1 FUEL GAS DATA

STREAM		TAR OIL HYDROTREATER OFFGAS	TAR OIL HYDROCRACKER OFFGAS	NAPHTHA HYDROTREATER OFFGAS	TOTAL HYDROTREATER AND HYDROCRACKER OFFGAS TO ACID GAS REMOVAL AND FUEL GAS SYSTEM			
TEMPERATURE	F	90	110	100				
PRESSURE	PSIA	40	50	60				
FLOWRATE	LB/HR	1114.1	619.3	328.7	2062.1			
	LB MOL/HR	65.94	24.05	11.15	101.1			
	SCFH	25025	9127	4231	38383			
AVG MW	LB/LB MOL	16.90	25.75	29.49	20.39			
COMPONENT		LB/HR			LB/HR	WT %	M/H	MOL %
H2O		20.7	8.6	2.3	31.8	1.5	1.8	1.7
NH3		135.3	19.6	2.7	157.5	7.6	9.3	9.1
H2		65.8	23.4	8.2	97.4	4.7	48.5	47.9
H2S		97.8	13.9	104.0	215.7	10.5	6.3	6.3
C1		107.2	6.3	1.4	114.9	5.6	7.2	7.1
C2		169.5	15.6	6.0	195.2	9.5	6.5	6.4
C3		181.9	108.9	11.8	302.6	14.7	6.9	6.8
iC4		56.8	110.8	0.8	168.4	8.2	2.8	2.8
nC4		13.9	203.8	156.5	374.1	18.1	6.2	6.2
C5+		265.3	104.2	35.0	404.5	19.6	5.6	5.6
TOTAL	LB/HR	1114.1	619.3	328.7	2062.1	100.0	101.0	99.9

TABLE 9 - ALTERNATE 1 SOUR WATER AND WASTEWATER STREAM RATES AND COMPOSITIONS

STREAM		TAR OIL HYDROTREATER SOUR WATER	TAR OIL HYDROCRACKER SOUR WATER	NAPHTHA HYDROTREATER SOUR WATER	PHENOL EXT ACID WASH WASTE WATER	PHENOL EXTRACTION WASTE WATER	ARU WASTE WATER
TEMPERATURE	F	120	120	120			
PRESSURE	PSIA	1890	1095	710			
FLOWRATE	LB/HR	15790.5	243.8	2603.5	1401.0	633.0	665.0
	LB MOL/HR	874.04	13.53	148.61	65.56	34.36	36.90
	GPM @60 F	32.04	0.49	5.55	2.55	1.26	1.33
AVG MW	LB/LB MOL	18.07	18.02	17.52	21.37	18.42	18.02
COMPONENT	LB/HR						
H2O		15358.7	243.5	2559.5	1132.0	615.9	665.0
NH3		276.1	0.2	0.1	0.0	0.0	0.0
H2		1.9	0.0	11.3	0.0	0.0	0.0
H2S		152.3	0.1	32.6	0.0	0.0	0.0
C1		0.8	0.0	0.0	0.0	0.0	0.0
C2		0.4	0.0	0.0	0.0	0.0	0.0
C3		0.2	0.0	0.0	0.0	0.0	0.0
1C4		0.0	0.0	0.0	0.0	0.0	0.0
nC4		0.0	0.0	0.0	0.0	0.0	0.0
C5+		0.1	0.0	0.0	0.0	0.0	0.0
PHENOL		0.0	0.0	0.0	0.0	17.1	0.0
H2SO4		0	0	0	269.0	0.0	0.0
TOTAL	LB/HR	15790.5	243.8	2603.5	1401.0	633.0	665.0

TABLE 10 - ALTERNATE 2 FINISHED AND INTERMEDIATE PRODUCT RATES

FINISHED PRODUCT RATES

PRODUCT	BBL/SD	LB/HR	LB/GAL
STABILIZED NAPHTHA TO CATALYTIC REFORMING	0	0	0.00
GASOLINE BLEND			
- FROM NAPHTHA STAB.	0	0	0.00
- FROM ARU	52	526	5.78
TOTAL =	52	526	5.78
JP-8	0	0	0.00
PHENOL	296	4,606	8.89
o-CRESOL	52	790	8.68
m,p-CRESOL	123	1,846	8.58
XYLENOLS	70	1,001	8.17
BENZENE	307	3,958	7.37
TOLUENE	109	1,389	7.28
XYLENE	15	183	6.97
TOTAL FINISHED PRODUCT	1024	14299	

INTERMEDIATE PRODUCT RATES

PRODUCT	BBL/SD	LB/HR	DISPOSITION
160 F- NAPHTHA	197	2,110	Boiler Fuel
NEUTRAL OIL (from PHENOL EXTRACTION)	263	4,157	Boiler Fuel
HEAVIES (from PHENOL DISTILLATION)	29	459	Boiler Fuel
PURGE GAS (from UPGRADING)	297 (SCF/HR)	20	Rectisol Unit feed
OFFGAS (from UPGRADING)	4,231 (SCF/HR)	329	Boiler Fuel (following H2S removal)
PSA PURGE GAS	9,999 (SCF/HR)	26	Boiler Fuel

TABLE 11 - ALTERNATE 2 PURGE GAS STREAM DATA

STREAM		TAR OIL HYDROTREATER PURGE GAS	TAR OIL HYDROCRACKER PURGE GAS	NAPHTHA HYDROTREATER PURGE GAS	TOTAL PURGE GAS FLOWRATE
TEMPERATURE	F	NO FLOW	NO FLOW	120	120
PRESSURE	PSIA	0.0	0.0	710	710
FLOWRATE	LB/HR	0.0	0.0	20.4	20.4
	LB MOL/HR	0.0	0.0	0.78	0.78
	SCFH	0.0	0.0	297.0	297.0
AVG MW	LB/LB MOL	0.0	0.0	26.09	26.09
COMPONENT	LB/HR				
H2O		0.0	0.0	0.2	0.2
NH3		0.0	0.0	8.4	8.4
H2		0.0	0.0	0.0	0.0
H2S		0.0	0.0	3.3	3.3
C1		0.0	0.0	0.7	0.7
C2		0.0	0.0	0.8	0.8
C3		0.0	0.0	0.6	0.6
1C4		0.0	0.0	0.0	0.0
nC4		0.0	0.0	3.3	3.3
C5+		0.0	0.0	3.2	3.2
TOTAL	LB/HR	0.0	0.0	20.4	20.4

TABLE 12 - ALTERNATE 2 FUEL GAS DATA

STREAM		TAR OIL HYDROTREATER OFFGAS	TAR OIL HYDROCRACKER OFFGAS	NAPHTHA HYDROTREATER OFFGAS	TOTAL HYDROTREATER AND HYDROCRACKER OFFGAS TO ACID GAS REMOVAL AND FUEL GAS SYSTEM		
TEMPERATURE	F	NO FLOW	NO FLOW	100	100		
PRESSURE	PSIA	0.0	0.0	60	60		
FLOWRATE	LB/HR	0.0	0.0	328.7	328.7		
	LB MOL/HR	0.0	0.0	11.15	11.1		
	SCFH	0.0	0.0	4231	4231		
AVG MW	LB/LB MOL	0.0	0.0	29.49	29.49		

COMPONENT	LB/HR	LB/HR	LB/HR	LB/HR	WT %	M/H	MOL %
H2O		0.0	0.0	2.3	0.7	0.1	1.2
NH3		0.0	0.0	2.7	0.8	0.2	1.4
H2		0.0	0.0	8.2	2.5	4.1	36.6
H2S		0.0	0.0	104.0	31.7	3.1	27.4
C1		0.0	0.0	1.4	0.4	0.1	0.8
C2		0.0	0.0	6.0	1.8	0.2	1.8
C3		0.0	0.0	11.8	3.6	0.3	2.4
IC4		0.0	0.0	0.8	0.2	0.0	0.1
nC4		0.0	0.0	156.5	47.6	2.6	23.4
C5+		0.0	0.0	35.0	10.7	0.5	4.4
TOTAL	LB/HR	0.0	0.0	328.7	100.0	11.1	99.4

TABLE 13 - ALTERNATE 2 SOUR WATER AND WASTEWATER STREAM DATA

STREAM		TAR OIL HYDROTREATER SOUR WATER	TAR OIL HYDROCRACKER SOUR WATER	NAPHTHA HYDROTREATER SOUR WATER	PHENOL EXT ACID WASH WASTE WATER	PHENOL EXTRACTION WASTE WATER	ARU WASTE WATER
TEMPERATURE	F	120	120	120			
PRESSURE	PSIA	1890	1095	710			
FLOWRATE	LB/HR	0	0.0	2603.5	1401.0	633.0	665.0
	LB MOL/HR	0.00	0.00	148.61	65.56	34.36	36.90
	GPM @60 F	0.00	0.00	5.55	2.55	1.26	1.33
AVG MW	LB/LB MOL	ERR	ERR	17.52	21.37	18.42	18.02
COMPONENT	LB/HR						
H2O		0.0	0.0	2559.5	1132.0	615.9	665.0
NH3		0.0	0.0	0.1	0.0	0.0	0.0
H2		0.0	0.0	11.3	0.0	0.0	0.0
H2S		0.0	0.0	32.6	0.0	0.0	0.0
C1		0.0	0.0	0.0	0.0	0.0	0.0
C2		0.0	0.0	0.0	0.0	0.0	0.0
C3		0.0	0.0	0.0	0.0	0.0	0.0
1C4		0.0	0.0	0.0	0.0	0.0	0.0
nC4		0.0	0.0	0.0	0.0	0.0	0.0
C5+		0.0	0.0	0.0	0.0	0.0	0.0
PHENOL		0.0	0.0	0.0	269.0	17.1	0.0
H2SO4		0	0	0		0.0	
TOTAL	LB/HR	0.0	0.0	2603.5	1401.0	633.0	665.0

distillation, fuel gas from the hydrotreaters and hydrocracker, and in Alternate 2, the PSA purge gas.

As discussed previously, the SO_2 emissions contributions from the GPGP by-products are based on the historical statistical maximum sulfur content of these streams. The sulfur content of the fuel gas produced in the Jet Fuel/Chemicals facility assumes 93% H_2S removal (7) via Sulfolin scrubbing in a dedicated scrubber. Based on information supplied by ANG (11), we assumed that sulfur was evenly distributed between the 160^oF^- and 160^oF^+ naphtha fractions. Since no information was available for the neutral oil/heavies we assumed that all of the sulfur in the crude phenol was concentrated in the neutral oil/heavies. Heating values (HHV) for the tar oil and the combined crude phenol/naphtha streams were taken from the "Application". Separate heating values for the naphtha and crude phenol were determined by using historical naphtha HHV data to backcalculate the heating value for the crude phenol. Higher heating values for the neutral oil, heavies, and 160^oF^- naphtha were extracted from the preliminary Amoco/LCI design.

The results of the fuel balance and SO_2 emissions adjustment are summarized for Alternates 1 and 2 in Tables 14 and 15, respectively. As shown, Alternate 1 offers a potential for a relatively large incremental SO_2 emissions credit of 745 lb/hr while Alternate 2 offers a considerably smaller credit of 204 lb/hr. Therefore, use of a low-sulfur fuel in conjunction with diversion of GPGP by-products to jet fuels and/or chemicals production presents an opportunity to reduce GPGP SO_2 emissions to levels even lower than those proposed in the "Application".

TABLE 14 - FUEL BALANCE, INCREMENTAL FUEL REQUIREMENTS, AND SO_2
CONTRIBUTIONS FOR ALTERNATE 1

		LB/HR	MMBTU/HR	LB SO2/HR
FUEL LOST TO UPGRADING				
-TAR OIL	(0.6 wt% S)	47910	(776.1)	(575)
-CRUDE PHENOL	(0.1 wt% S)	13550.0	(186.8)	(27)
-RECTISOL NAPHTHA	(1.6 wt% S)	8519.0	(144.2)	(273)
ADDITIONAL FUEL DEMAND (Eff=85%)				
-PROCESS HEATERS			(7.4)	
-IMPORT STEAM		62140	(73.1)	
GROSS FUEL REQ'T			(1,187.6)	
FUEL PRODUCED IN UPGRADING				
-160 F- NAPHTHA	(1.6 wt% S)	2110.0	36.7	68
-NEUTRAL OIL	(0.3 wt% S)	4616.0	69.2	28
-FUEL GAS	(0.84 wt% S)	2062.1	37.9	35
-PSA PURGE	(0 wt% S)	0.0	0.0	0
TOTAL ADD'L FUEL			143.8	
TOTAL NET FUEL MAKEUP REQUIREMENT			(1,043.8)	
NET SO2 EMISSIONS				(745)

TABLE 15 - FUEL BALANCE, INCREMENTAL FUEL REQUIREMENTS, AND SO₂ CONTRIBUTIONS FOR ALTERNATE 2

	LB/HR	MMBTU/HR	LB SO2/HR
FUEL LOST TO UPGRADING			
-TAR OIL (0.6 wt% S)	0	0.0	0
-CRUDE PHENOL (0.1 wt% S)	13550.0	(186.8)	(27)
-RECTISOL NAPHTHA (1.6 wt% S)	8519.0	(144.2)	(273)
ADDITIONAL FUEL DEMAND (Eff=85%)			
-PROCESS HEATERS		0.0	
-IMPORT STEAM	69250	(81.5)	
GROSS FUEL REQ'T		(412.5)	
FUEL PRODUCED IN UPGRADING			
-160 F- .NAPHTHA (1.6 wt% S)	2110.0	36.7	68
-NEUTRAL OIL (0.3 wt% S)	4157.0	62.4	25
-FUEL GAS (0.84 wt% S)	196.0	3.5	3
-PSA PURGE (0 wt% S)	488.0	5.4	0
TOTAL ADD'L FUEL		108.0	
TOTAL NET FUEL MAKEUP REQUIREMENT		(304.5)	
NET SO2 EMISSIONS			(204)

6.5 Jet Fuels/Chemical Production Facility Investment, Operating Costs, and Utility Consumption

The investments, operating costs, and utilities consumptions for the Jet Fuel and Chemical Production Facilities are taken from the preliminary Amoco/LCI design and have been adjusted for the naphtha, crude phenol, and tar oil rates used in this study. Process unit investments developed in the Amoco/LCI design were prorated based on their respective feedrates using an 0.7 exponent. Although the proration exponents for each area may be somewhat different than 0.7, the differences in unit throughput are so small that fine-tuning was not judged to be required. The unit investments can be more accurately determined following completion of the detailed design. These revised investments are presented for Alternate 1 and Alternate 2 in Tables 16 and 17, respectively. Using the by-product flow rate basis previously discussed results in about a 0.5% investment increase for Alternate 1 and a 0.2% decrease for Alternate 2, changes which are insignificant compared to those which may occur as the project moves into detailed design stage.

Utilities consumptions and operating costs were similarly adjusted for the revised by-product flow rates, and are presented for Alternates 1 and 2 in Tables 18 and 19, respectively. Since these are assumed to be linearly dependent on the unit feed rates, the changes are directly proportional to the differences between the Amoco/LCI basis and the basis used for this study. Overall, the net effect of these changes will have only a very minor effect on the economics of producing jet fuels and chemicals.

Values for the GPGP liquid by-products and the Jet Fuels/Chemicals Facility products and co-products have been extracted from the Amoco/LCI study and are presented in Table 20.

TABLE 16 - INVESTMENT BASIS FOR ALTERNATE 1 - PROFITABLE JP-8 PRODUCTION

AREA	PROCESS DESCRIPTION	T.I.C.
		(1000$)
AREA 100	TAR OIL HYDROTREATER	$20.778
20U	TAR OIL HYDROCRACKER	10.049
300	PSA & RECOMPRESSION	8.182
400	OSBL	9.351
500	CATALYST HANDLING	1.290
600	NAPHTHA DISTILLATION & HYDROTREATING	4.545
700	AROMATICS RECOVERY UNIT (ARU)	9.231
800	PHENOL EXTRACTION	11.792
900	CRESYLIC ACID DISTILLATION	4.641
	SUBTOTAL	$79.859
700	SOLVENT INVENTORY	100
	TOTAL	$79.959

Note :

Total Installed Costs (T.I.C.) include labor, material, subcontracts, indirects, engineering costs, and contigencies.

The T.I.C. does not include:
Spare Parts
Start-Up
Insurance & Taxes
Permits
Process Licensing Royalties

Contigencies have been applied as follows:
20%   to Areas 100 thru 700
30%   to Areas 800 and 900

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TABLE 17 - INVESTMENT BASIS FOR ALTERNATE 2 - MAXIMUM PROFIT

AREA	PROCESS DESCRIPTION	T.I.C.
		(1000$)
AREA 100	TAR OIL HYDROTREATER	0
200	TAR OIL HYDROCRACKER	0
300	PSA & RECOMPRESSION	510
400	OSBL	5,909
500	CATALYST HANDLING	0
600	NAPHTHA DISTILLATION & HYDROTREATING	4,545
700	AROMATICS RECOVERY UNIT (ARU)	9,231
800	PHENOL EXTRACTION	11,792
900	CRESYLIC ACID DISTILLATION	4,641
	SUBTOTAL	$36,628
700	SOLVENT INVENTORY	100
	TOTAL	$36,728

Note :

Total Installed Costs (T.I.C.) include labor, material, subcontracts,
indirects, engineering costs, and contigencies.

The T.I.C. does not include:
    Spare Parts
    Start-Up
    Insurance & Taxes
    Permits
    Process Licensing Royalties

Contigencies have been applied as follows:
    20% to Areas 100 thru 700
    30% to Areas 800 and 900

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TABLE 18 - OPERATING REQUIREMENTS FOR ALTERNATE 1 - PROFITABLE JP-8 PRODUCTION

CATALYST AND CHEMICALS

	QUANTITY USED	COST	OPERATING COST
NAPHTHA HYDROTREATER	0.021 lb/bbl	3.00 $/lb	46 $/SD
TAR OIL HYDROTREATER	0.3 lb/bbl	3.00 $/lb	2,909
TAR OIL HYDROCRACKER	0.0095 lb/bbl	6.00 $/lb	39
INHIBITORS	50 ppm	10.00 $/gal	52
ARU SOLVENT (SULFOLANE)	24 lb/day	2.10 $/lb	50
SULFURIC ACID	7100 lb/day	0.04 $/lb	285
			3,381 $/SD

MAINTENANCE SUPPLY COSTS   4,010 $/SD

MANPOWER REQUIREMENTS

SHIFT PERSONNEL	60
SUPERVISOR + ADMIN.	6
QC TECHNICIAN	2
MAINTENANCE	7
OTHER (STORES or JANITORIAL)	1
	76

UTILITY REQUIREMENTS

POWER	6,204 KW	
COOLING WATER	5,885 GPM	(Assumes 30 F Rise)
PROCESS WATER	30.4 GPM	
STEAM IMPORT/(EXPORT)		
HP(750 F, 550 psig)	55,005 LB/HR	
MP(337 F, 100 psig)	7,135 LB/HR	
LP(296 F, 50 psig)	(5,460)LB/HR	
LP BFW	17,400 LB/HR	(Assumes 2% Blowdown)
CONDENSATE RETURN	70,680 LB/HR	
SNG EQUIVALENT OF SYN GAS AND PURGE GAS	3.7 MMSCFSD	(Removed in PSA Unit)
REPLACEMENT FUEL	-1043.8 MMBTU/HR	(Includes fuel required to generate process steam as well as process heater fuel. Assumes 85% efficiency.)

TABLE 19 - OPERATING REQUIREMENTS FOR ALTERNATE 2 - MAXIMUM PROFITABILITY

CATALYST AND CHEMICALS

	QUANTITY USED	COST	OPERATING COST
NAPHTHA HYDROTREATER	0.021 lb/bbl	3.00 $/lb	46 $/SD
TAR OIL HYDROTREATER	0 lb/bbl	3.00 $/lb	0
TAR OIL HYDROCRACKER	0 lb/bbl	6.00 $/lb	0
INHIBITORS	0 ppm	10.00 $/gal	0
ARU SOLVENT (SULFOLANE)	24 lb/day	2.10 $/lb	50
SULFURIC ACID	7100 lb/day	0.04 $/lb	285
			381 $/SD

MAINTENANCE SUPPLY COSTS $1,836 $/SD

MANPOWER REQUIREMENTS

SHIFT PERSONNEL	44
SUPERVISOR + ADMIN.	10
QC TECHNICIAN	2
MAINTENANCE	8
OTHER (STORES or JANITORIAL)	2
	66

UTILITY REQUIREMENTS

POWER	501 KW	
COOLING WATER	3,818 GPM	(Assumes 30 F Rise)
PROCESS WATER	7.8 GPM	
STEAM IMPORT/(EXPORT)		
HP(750 F, 550 psig)	55,005 LB/HR	
MP(337 F, 100 psig)	14,245 LB/HR	
LP(296 F, 50 psig)	(5,485)LB/HR	
LP BFW	7,031 LB/HR	(Assumes 2% Blowdown)
CONDENSATE RETURN	70,655 LB/HR	
SNG EQUIVALENT OF SVN GAS AND PURGE GAS	3.7 MMSCFSD	(Removed in PSA Unit)
REPLACEMENT FUEL	-304.5 MMBTU/HR	(Includes fuel required to generate process steam as well as process heater fuel. Assumes 85% efficiency)

TABLE 20 - PRODUCT PRICING STRUCTURE (AMOCO/LCI BASIS)

STREAM	S.G. 20 deg	DENSITY lb/gal	HHV Btu/Lb	FUEL VALUE $/MMBtu	PRICE $/Bbl	PRICE $/Lb
RECTISOL NAPHTHA	0.827	6.89	20,396	2.15	12.69	0.044
CRUDE PHENOL	1.065	8.87	13,022	2.15	10.43	0.028
TAR OIL	1.026	8.55	16,930	2.15	13.07	0.036
GPGP SYNGAS FOR H2	1.23 $/MSCF H2 EXTRACTED			2.47		
NATURAL GAS				2.15	13.57	
LPG/Propane	0.508	4.23		2.15	7.57	0.043
i-BUTANE	0.563	4.69		4.98	19.11	0.097
n-BUTANE	0.583	4.86		2.95	11.76	0.058
UNLEADED GASOLINE	0.767	6.39			23.35	0.087
UNLEADED PREMIUM	0.767	6.39			26.29	0.098
SWEETENED GPGP NAPHTHA					25.45	
REFORMER FEED	0.743	6.19			24.61	0.095
HYDROTREATED GPGP NAPHTHA	0.744	6.20			30.00	0.115
JP-8	0.807	6.73			21.84	0.077
BENZENE	0.885	7.37	18,375	8.44	48.00	0.155
TOLUENE	0.872	7.26	18,359	6.79	38.00	0.125
XYLENE	0.866	7.21	18,398	8.80	49.00	0.162
PHENOL	1.080	9.00	13,948	15.17	80.00	0.212
o-CRESOL	1.046	8.71	14,708	33.83	182.00	0.498
m.p-CRESOL	1.035	8.62	14,711	37.36	199.00	0.550
XYLENOLS	1.029	8.57	14,828	32.04	171.00	0.475
CRESYLIC ACIDS	1.005	8.37	14,709	25.91	134.00	0.381
2,4-,2,5-XYLENOLS	1.024	8.53				
3,5-XYLENOL	1.022	8.51				
GPGP FUEL POOL				2.15		
SULFUR						0.032

7.0 Process Integration

Once the intermediate and by-product return streams produced in the Jet Fuels and Chemicals Production Facilities were quantified, it then became possible to identify the preferred locations for these streams to be returned to the GPGP. Although Amoco/LCI developed preliminary recommendation regarding the integration locations, the actual practicality of returning these streams to the suggested locations had to be determined. Furthermore, the impact these returning streams have on the existing GPGP facilities also had to be assessed, particularly with regards to utility and catalyst/chemicals consumption. In addition, the Jet Fuels/Chemical Production Facility also places an incremental demand on existing GPGP utilities since it requires fuel, power, steam, cooling water, and boiler feed water. Although these utility consumptions are small relative to the overall GPGP, many of the utility facilities are either at or close to their operating limits. Consequently, the question of whether the small additional loads will dictate the construction of additional capacity was also addressed.

Therefore, BRSC/SAIC requested ANG to address these issues and to determine if and where new capacity is required. ANG also evaluated and defined the preferred integration locations for the streams returning to the GPGP. Although ANG was able to directly address most of the integration issues, Amoco/LCI provided some assistance in defining the types of contaminants that may be present in some streams, particularly in the crude phenol processing areas. This allowed ANG to more clearly determine whether streams needed additional treatment before return to GPGP or whether they should be rerouted to other units for disposal.

Based on ANG's review (12,13) and discussions with Amoco/LCI, block diagrams were developed summarizing the general integration scheme for the GPGP and are presented in Figures 5 and 6 for Alternates 1 and 2, respectively. A more specific description of the integration locations and required return conditions defined by ANG is discussed in Sections 7.1 and 7.2.
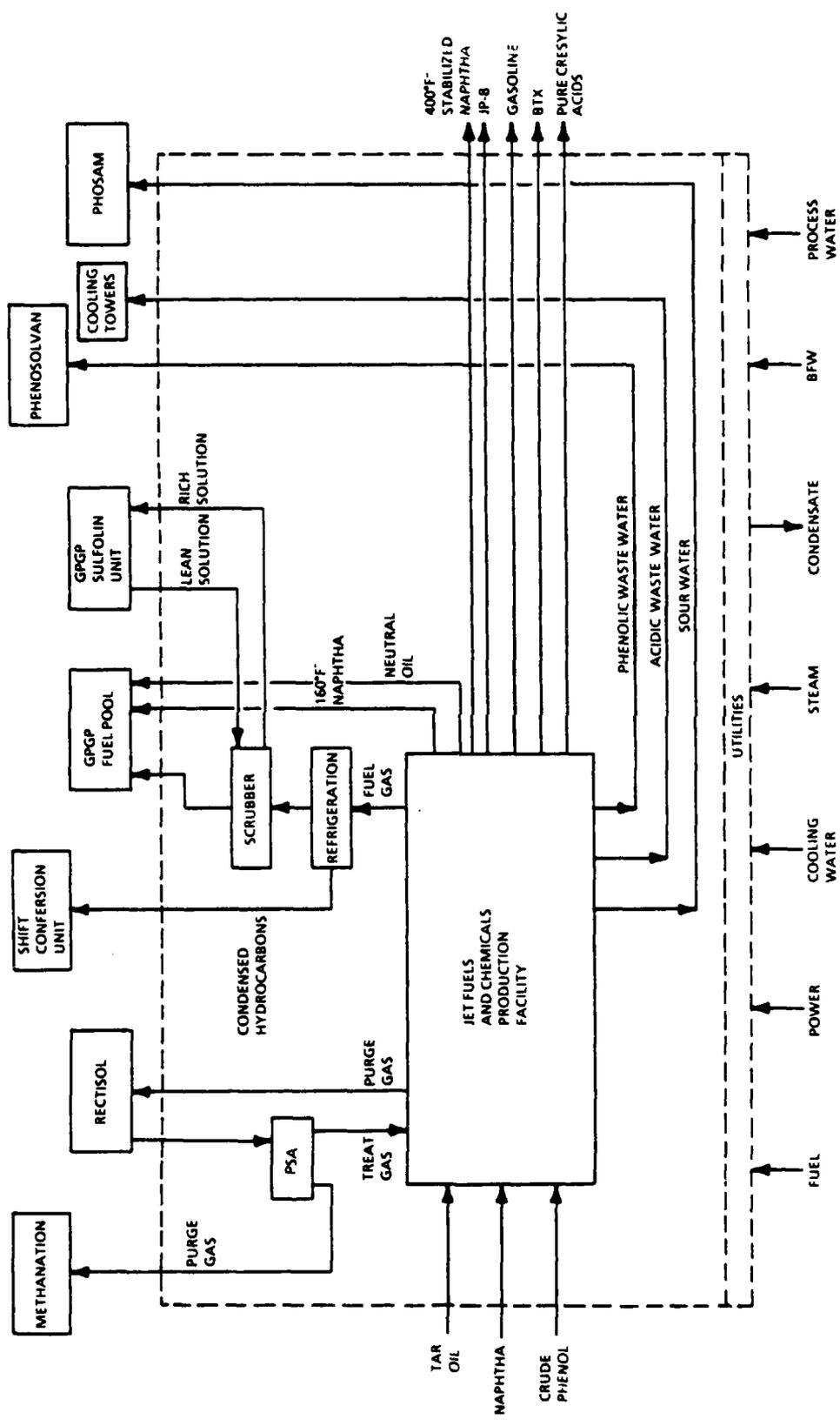
FIGURE 5. ALTERNATE 1 – INTEGRATION WITH GPGP FACILITIES
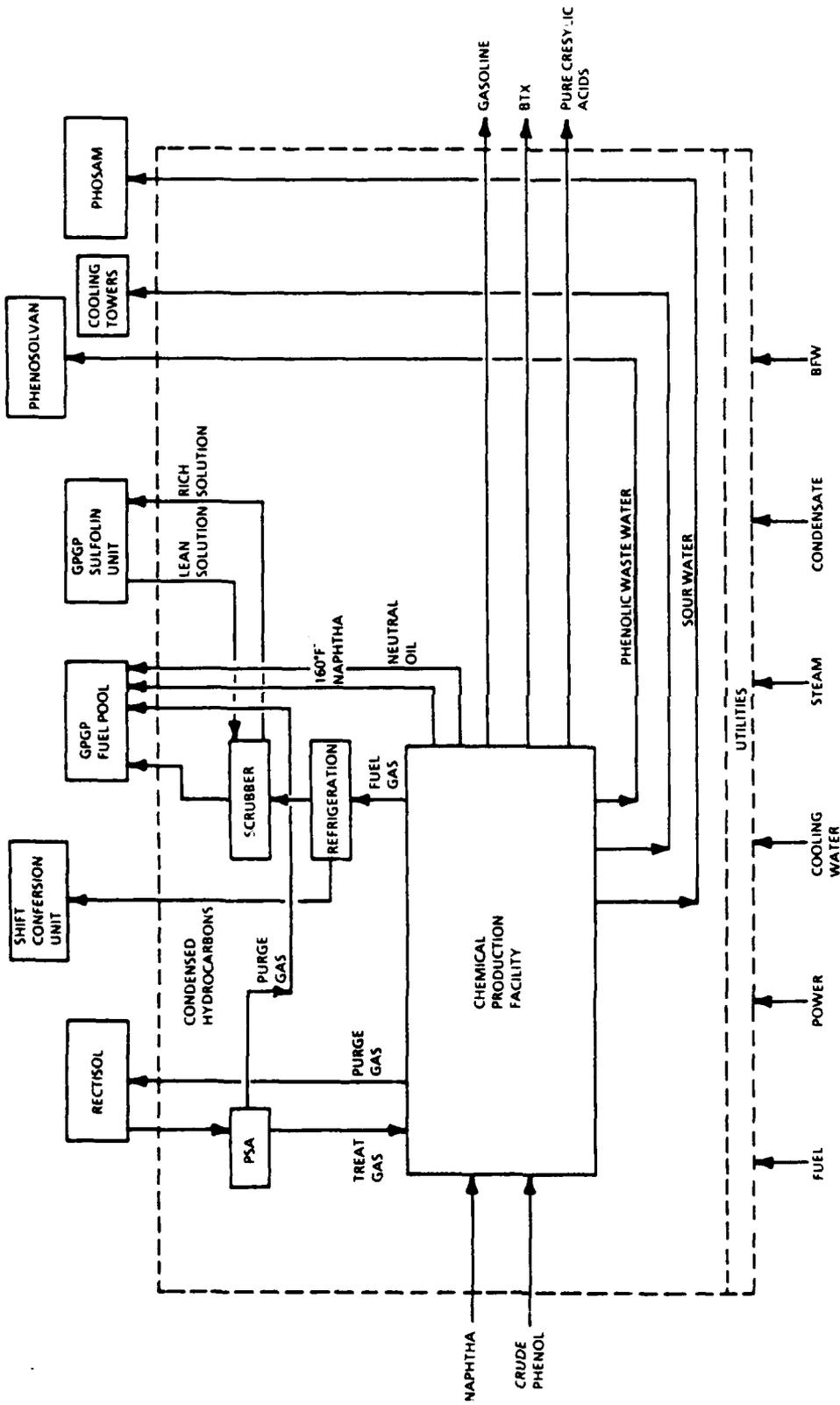
FIGURE 6. ALTERNATE 2 – INTEGRATION WITH GPGP FACILITIES

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7.1 Purge Gas and Off Gas Streams

o   Purge-gas from the Naphtha and Tar Oil Hydrotreaters and the Tar Oil Hydrocracker is to be sent to the inlet of the Rectisol Unit at a minimum inlet pressure of 450 psig and 95°F maximum temperature.

o   In Alternate 1, the PSA Unit purge gas will be blended into the feed to the Methanation Unit at a minimum pressure of 375 psig and at compressor discharge temperature. In Alternate 2 the PSA purge gas will be sent directly to fuel at 35 psig and 200°F maximum temperature owing to its low flowrate.

o   Fuel gas streams from Areas 100, 200, and 600 will be sent to the GPGP fuel system after treating to remove H_2S. These fuel gas streams will be sent to a refrigeration unit to remove condensible hydrocarbons before scrubbing with Sulfolin solution to remove H_2S. Following refrigeration and scrubbing, the return temperature should be ~100°F and the pressure shall be ~35 psig. While the refrigeration unit and scrubber will have to be built as part of the Jet Fuel/Chemicals production facilities, the GPGP will provide lean Sulfolin solution and regenerate the rich solution.

o   Condensate from the refrigeration unit will be injected into the Shift Conversion Unit feed at a minimum pressure of 450 psig.

7.2 Process Wastewater Streams

o   Waste water from the tar acid wash section of the Cresylic Acid Recovery Section will be sent to the GPGP cooling towers. This stream should have a pressure that is suffi-cient to enter the CW return header which has a normal

operating pressure of 30-40 psig and taking into considera-
tion that this line is about 25 feet above ground.

o   Phenolic waste water from the Crude Phenol column can be
     delivered to the Phenosolvan Unit at 70 psig and preferably
     at ambient temperature.

o   Waste water from the Aromatics Recovery Unit (Area 700) will
     be blended with the gas liquor stream and processed in the
     Phosam Unit along with sour water from the Naphtha and Tar
     Oil Hydrotreaters and the Tar Oil Hydrocracker at a minimum
     pressure of 90 psig. Any temperature is acceptable.

7.3 Impact of Jet Fuel Facility Intermediate, Waste, and Utility Streams

The evaluation also assessed the effect returning purge-gas, off
gas, and waste water streams would have on existing unit
operations.   Operating costs were given particular attention,
considering if and how much incremental catalyst and chemical
consumptions and utility requirements would change.   The effect of
synthesis gas diversion to the PSA Hydrogen Recovery Unit on SNG
production was similarly assessed.   As a result of this
evaluation, we reached following conclusions:

o   Flow rates of returning streams are so small relative to
     normal process circulation that the overall impact will be
     negligible.   In fact, these rates are so low that unit
     instrumentation could not truly indicate a real difference.

o   GPGP has no spare cooling water capacity, and therefore,
     additional cooling water will have to be provided from a new
     unit.

o   Assuming the SNG booster compressor is installed as stated in the integration basis, 550 psig steam can be produced in the existing boilers. Lower pressure steam may be directly available or may have to be let down from the HP header.

o   LP boiler feedwater will be available for the quantities required.

o   Condensate must be returned at a maximum temperature of $100^\circ F$ since existing condensate coolers are already at capacity. The remainder of the condensate system can process the quantities produced by the Jet Fuels Facility.

To further facilitate the integration of utility connections from the Jet Fuels/Chemicals Production Facilities with the existing GPGP systems, and to supplement information contained in the GPGP Public Design Report (14), additional information was requested from ANG and LCI. LCI subsequently provided design basis information for the original Steam Generation, Oily Waste Water Treatment, and Cooling Water systems while ANG provided similar information for the Electrical System.

ANG provided additional feedback on the changes to the original design of the cooling tower, particularly regarding the adaptation of the GPGP cooling towers to utilize contaminated waste water streams. Although ANG had concluded that there was no spare cooling water capacity available, they also advised BPSC/SAIC that the existing cooling water system would be able to accommodate incremental Stripper Gas Liquor (SGL) from the Phosam Unit (15). The additional SGL would be produced by Phosam unit processing of the waste water streams derived during Tar Oil and Naphtha Hydrotreater, Tar Oil Hydrocracker, and Aromatics Recovery Unit (ARU) operation. Processing and treatment of the incremental SGL in the existing cooling towers has two benefits: first, it avoids the need to install a dedicated sour water treatment unit, and

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second, it permits the new cooling tower capacity to be based on the use of clean make-up water following standard design procedures. Therefore, BRSC/SAIC proceeded to design the required cooling water facilities as a clean system, using softened water from the GPGP rather than cold lime softened water, which according to ANG, is not as readily available throughout the plant.

ANG stated that they had "no concerns about handling minor quantities of streams for or from the jet fuel plant additions" regarding the steam generation and oily water systems.

8.0 SO_2 Emissions Control Evaluation

There are basically two general approaches to controlling the main stack SO_2 emissions, e.g.:

1. Use replacement fuel which has sulfur levels below or equivalent to those in the tar oil, phenols/naphtha streams currently being fired in the Riley boilers. Or,

2. Desulfurize the flue gases from the Riley boilers when firing with high sulfur oils or lignite. Options in this category include commercial wet and spray dryer desulfurization (FGD) systems and the use of emerging technologies such as furnace or in-duct injection of SO_2 sorbents.

Each of these are discussed below.

8.1 Replacement Boiler Fuels Identification and Characterization

Since the GPGP currently utilizes the naphtha, crude phenol, and tar oil by-products as plant fuel, diversion of any or all of these streams to jet fuels and/or chemicals production requires that a suitable makeup fuel be identified. Furthermore, Amoco/LCI's preliminary economic evaluation indicates the fuel should cost no more than $3.00/MMBtu for Alternate 1 to be

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profitable (16). The Amoco/LCI economic evaluation indicates that Alternate 2 (Case 7) is far less sensitive to fuel price because it requires a much smaller quantity of makeup fuel.

Consequently, BRSC/SAIC identified and characterized several alternative fuels in order to evaluate their impact on boiler performance, sulfur emissions, and process economics. Based on recommendations from program participants as well as BRSC/SAIC's own investigations, several candidate replacement fuels were identified. They are:

1. Medora/Fryburg Crude Blena - available at Amoco's Mandan, North Dakota refinery.

2. Decanted Oil (DCO) - also available from Amoco's Mandan refinery.

3. Industrial Fuel Oil - available from Koch Refining Company's Rosemount, Minnesota refinery.

4. Medora Crude - Owned by Amerada-Hess, available in North Dakota.

5. Fryburg Crude - Owned by Amerada-Hess, available in North Dakota.

In addition, SNG was also evaluated as a replacement fuel in Alternate 3. Lignite did not receive consideration because of special technical and economic limitations presented in Appendix A.

Properties of these potential replacement fuels have been obtained or estimated for use in this study. Information provided by Amoco for DCO, Mandan and Fryburg crudes and Koch Refining Co. for the industrial fuel oil is provided in Table 21. Since combustion data for the Medora crude, the Fryburg crude, and the Decant Oil (DCO) were not available, Amoco R&D recommended we use the API Data Book Correlation (17). The resulting calculated gross heating values and sulfur contents are also summarized in Table 21. As shown, Medora and Fryburg crudes have almost identical heating

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TABLE 21 - PROPERTIES OF POTENTIAL LIQUID FUEL REPLACEMENTS

Property	Mandan[1] DCO	Medora[1] Crude	Fryburg[1] Crude	Koch[2] Industrial Fuel Oil	Medora/ Fryburg[3] Blend
API Gravity	11.1	36.7	36.8	-2.3	36.7
RVP, psi	0.0	3.1	0.8	---	1.9
Ash, wt%	0.01	0.0	0.002	---	0.001
B.S.&W., wt%	0.0	---	---	1.0 Max	---
Sulfur, wt%	1.7	0.47	0.32	1.5 Max	0.40
Nitrogen, wt%	0.09	0.02	0.02	---	0.02
Pour Point, °F	50	0	65	50 Max	---
Metals, ppm					
Nickel	0	2	3	---	2.5
Vanadium	0	1	1	---	1
Iron	10	15	13	---	14
Conradson Carbon, wt%	---	2.1	2.9	---	2.5
Salt, ppb	0	6	25	---	16
Viscosity, SUS					
at 68°F	---	---	85	---	---
at 100°F	680	43	52	700-1100	---
at 122°F	---	34	44	---	---
at 210°F	55	---	---	---	---
Boiling Range, vol%					
500°F minus	1	42	35	---	---
500-1000°F	89	41	51	---	---
1000°F plus	10	17	14	---	---
IBP, °F	450	80	80	---	---
Gross Heat of Combustion, Btu/gal	149,300	136,800	136,900	153,000	136,850
Current Price, $/B (FOB)	8.40	---14.00---		14.70	14.00
Maximum Volume, B/D	500	--	--	--	--

[1] Reference 18.

[2] Reference 19..

[3] Calculated based on 50/50 blend of Medora and Fryburg crudes.

values and thus, the overall heating value of the blend should not vary much from those shown in the table. The most important factor is the high sulfur content of the DCO, which is four to six times that of the Medora or Fryburg crudes. Furthermore, Amoco has indicated that a maximum of only 500 Bbl/Day of the DCO would be available compared to a maximum requirement of almost 4000 Bbl/day. Consequently, the high sulfur content and limited availability of the DCO outweigh the advantage of its low cost, which was about $8.50/Bbl compared to $14.00/Bbl (FOB refinery) for the Medora and Fryburg crudes. Delivered costs for the candidate fuels are also shown in Table 22.

In addition, ANG has supplied information on the sulfur content of the Rectisol naphtha as a function of boiling point since both alternates utilize the $160^{\circ}F^-$ naphtha as fuel. Therefore, it is important to know how the sulfur is distributed between the $160^{\circ}F^-$ and $160^{\circ}F^+$ cuts in order to determine what sulfur emissions control steps are required. These data should be used cautiously since they was taken in 1985 and may not be completely representative of operations with the naphtha stripper which ANG recommissioned earlier this year. Nonetheless, the ANG data indicates that the $160^{\circ}F^-$ fraction will contain about the same concentration as the $160^{\circ}F^+$ material. Therefore, in the absence of more recent data, this study assumes that the sulfur concentration of the $160^{\circ}F^-$ and $160^{\circ}F^+$ fraction are the same.

8.2 Replacement Fuel SO_2 Emissions Evaluation Basis

Under normal GPGP operations, the tar oil, naphtha, and crude phenol are used to fuel the Riley boilers, the superheaters, and the liquid waste incinerator. Table 23 (excerpted from the "Application" presents the total main stack emissions for normal fuel firing rates. Under these conditions and based on the projected by-product sulfur contents, the SO_2 emisisons from the main stack total 3,942 lbs/hr.

TABLE 22 - CALCULATED HEATING VALUES AND SULFUR CONTENT OF REPLACEMENT FUELS CANDIDATE

Fuel	HHV, Btu/gal	Sulfur, lb/MMBtu	Delivered Cost $/bbl
Mandan DCO	149,300	0.94	~ 9.25[1]
Medora Crude	136,800	0.24	~17.00[2]
Fryburg Crude	136,900	0.16	~17.00[2]
Medora/Fryburg Blend	136,850	0.20	~17.00[2]
Koch Industrial Fuel Oil	153,000	0.89	~18.60[3]
SNG	1,008 Btu/SCF	nil	3.70 $/MMBtu

[1] Based on transportation charge of 0.01 $/bbl/mile for a tank truck within North Dakota and 73 mile distance between Mandan and Beulah per J.E. Sinor report.

[2] Based on posted price of $12.90-$14.00/bbl plus $3.00/bbl charge for pipeline to Beulah.

[3] Includes $3.86/bbl transportation charge to take IFO from Minneapolis, MN to Beulah, ND. Based on J.E. Sinor report.

TABLE 23 - GREAT PLAINS MAIN STACK EMISSION SUMMARY NORMAL FUEL FIRING RATE

Fuel	Firing Rate SCFH/pph	Heating Value Btu/cf-Btu/lb	Heat Input MMBtu/hr	%S	NOx Factor lb/MMBtu	Particulate Factor lb/MMBtu	SO2 Contribution lb/hr	NOx Contribution lb/hr	Particulate Contribution lb/hr
Boilers									
Rectisol Off-Gas	9.05×10^6	50	452.5	1,605 ppmv[1]	0.20	0.10	2,450	91	45.2
Fuel Gas	1.02×10^6	310	316.2	1,440 ppmv[2]	0.20	0.10	248	63	31.6
Phosam Off-Gas	2.19×10^5	45	9.9	2,169 ppmv[3]	0.20	0.10	80	2	1.0
Tar Oil	26,012	16,200	421.4	0.80%	0.65	0.10	416	274	42.1
Naphtha/Phenol	22,069	15,000	331.0	0.93%	0.40	0.10	411	132	33.1
			1,531.0				3,605	562	153.0
Superheaters									
Tar Oil	17,651	16,200	286.0	0.80%	0.75	0.10	282	215	28.6
Liquid Waste Incinerator									
Tar Oil	4,247	16,200	68.8	0.80%			55[4]	50[4]	25[4]
			Main Stack Total				3,942	827	207

[1] 1,029 ppmv H₂S + 576 ppmv Org. S. (93.0% H₂S Removal Efficiency, 0% Org. S. Removal Efficiency).

[2] 1,170 ppmv H₂S + 270 ppmv Org. S. (70.0% H₂S Removal Efficiency, 0% Org. S. Removal Efficiency).

[3] 1,729 ppmv H₂S + 330 ppmv Org. S. (93.0% H₂S Removal Efficiency, 0% Org. S. Removal Efficiency).

[4] Same as original permitted. (Some SO₂ removal is provided by LWI quench tank and scrubber).

Using the tar oil, naphtha, and crude phenol as feedstock for the Jet Fuel Production Facility alters the SO_2 emission balance and route. In this case, the equivalent SO_2 content in the feedstock streams are sent as sulfur-bearing organic compounds through the processing units. In the case of the tar oil and the 160°F+ cut of the naphtha, the sulfur is converted to H_2S in the hydrotreaters and hydrocracker and removed with the offgas streams, ultimately ending up in the Sulfolin units where 93% of the equivalent SO_2 is removed as elemental sulfur.

Subtracting the equivalent feedstock SO_2 content of 1164 lb/hr (Table 2) used as the BACT basis, the net SO_2 contribution of the Rectisol and Phosam off-gas streams and fuel gas is 2,778 lb/hr. Based on historical rather than projected by-product sulfur contents, the equivalent SO_2 content of these by-products is 875 lb/hr and thus, the maximum potential GPGP SO_2 emissions would be (875 + 2,778) or 3,653 lb/hr. This is the basis against which replacement fuel SO_2 emissions impact is assessed.

In order to accurately assess the overall impact of Jet Fuels/Chemical Production Facility integration on the SO_2 emissions, the use of process intermediate streams as process fuel must also be added in along with incremental fuel requirements. Since in Alternate 2, the tar oil is not upgraded and utilized as fuel it, too, must be accounted for in the emissions calculation. This was done during basis development and is summarized in Table 24. As indicated, the maximum reduction in SO_2 emissions would be 745 lb/hr for Alternate 1 and 204 lb/hr for Alternate 2 (Case 7). Expressed in other terms, the base SO_2 emissions level, for evaluating replacement fuels in Alternate 1 (Case 4) is 2,908 lb/hr, and 3,449 lb/hr in Alternate 2. Adding in the SO_2 emissions from the Thermal Oxidizer (7 lb/hr) and the startup flare (340 lb/hr) increases the overall GPGP SO_2 emissions from 3,653 lb/hr to 4,000 lb/hr, the Alternate 1 (Case 4) base from 2,908 lb/hr to 3,255 lb/hr, and the Alternate 2 (Case 7) base from 3,449 lb/hr to 3,796 lb/hr.

TABLE 24 - POTENTIAL MAXIMUM SO$_2$ EMISSIONS REDUCTION SUMMARY

	Alternate 1		Alternate 2	
	MMBtu/hr	LBS SO$_2$/hr	MMBtu/hr	LBS SO$_2$/hr
Fuel Diverted to Upgrading	1107.1	(875)	331.0	(300)
Jet Fuels/Chemicals Facility Fuel Requirement	80.5	---	81.5	---
Overall Fuel Requirement	1187.6		412.5	
Fuel from Jet Fuels/Chemicals Production Facility	143.8	130	108.0	96
Net Fuel Demand	1043.8		304.5	
Net SO$_2$ Emissions Reduction		(745)		(204)

8.3 Replacement Fuel Impact on SO_2 Emissions

Following establishment of a consistent basis for evaluating and comparing GPGP SO_2 emissions, the candidate replacement fuels were evaluated. As discussed previously these candidates are:

o   Medora crude

o   Fryburg crude

o   Medora/Fryburg crude blend

o   Mandan Refinery Decanted Oil (DCO)

o   Industrial Fuel Oil

o   SNG

Since the Mandan and Fryburg crudes are very similar, and data on the blend of these two were unavailable, the blend was evaluated on the basis of a 50/50 mixture of the two crudes. In addition, the Mandan DCO was dropped from the evaluation due to its' high sulfur content and limited availability.

The impact of replacing the GPGP naphtha, phenol, and tar oil by-products with the candidate replacement fuels was evaluated with regards to their potential effect on total GPGP SO_2 emissions and is summarized in Table 25. Not surprisingly, SNG would be the most attractive fuel for use as replacement fuel since it is essentially sulfur-free and would reduce overall GPGP SO_2 emission by 19% in Alternate 1 (Case 4), and 5% in Alternate 2 (Case 7). However, SNG at its' current selling price of $3.70 to $3.80 per MMBtu is too valuable to be consumed internally as fuel. Therefore, the next most attractive replace-ment fuel would be either the Fryburg or Medora crudes or a blend of both. In Alternate 1 (Case 4), use of these crudes or the crude blend as replacement fuel would reduce overall GPGP SO_2 emissions by 6%-10% as compared to the BACT base case. The reduction in SO_2 emissions for Alternate 2 (Case 7) is virtually insignificant (only 1%-3%) owing to the use of the high sulfur (0.6 wt%) tar oil as fuel rather than process feedstock. Furthermore, use of the tar oil reduces the requirement for lower   sulfur   (0.3 wt%-0.5 wt%)   Fryburg   and   Medora   crudes   from

TABLE 25 - REPLACEMENT FUEL IMPACT ON GPGP SO₂ EMISSIONS

| | ----Total SO₂ Contributions---- | | | ---Total SO₂ Emissions--- | | |
	From Offgases and Misc Sources(2) lb/hr	From Byproducts and Intermediates lb/hr	From Replacement Fuels lb/hr	lb/hr	% of Base	Overall(3) Lb SO₂ /MMBtu
Base Case(1)	3125	875	0	4000	Base	1.96
Alternate 1						
- w/Medora Crude	3125	130	501	3756	94%	1.77
- w/Fryburg Crude	3125	130	345	3600	90%	1.69
- w/Industrial Fuel Oil	3125	130	1440	4695	117%	2.21
- w/SNG	3125	130	0	3255	81%	1.53
Alternate 2						
- w/Medora Crude	3125	671	146	3942	99%	1.85
- w/Fryburg Crude	3125	671	100	3896	97%	1.83
- w/Industrial Fuel Oil	3125	671	420	4216	105%	1.98
- w/SNG	3125	671	0	3796	95%	1.79

NOTES:

(1) Base SO₂ emissions are based on historical S content from naphtha, phenol, and tar oil.

(2) Includes Rectisol OG (2450 lb/hr), fuel gas (248 lb/hr), Phosam OG (80 lb/hr), expansion vessel vent gas to the startup flare (340 lb/hr), and fuel gas to the thermal oxidizer (7 lb/hr).

(3) Includes waste fuels (i.e., Rectisol and Phosam Offgases).

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1043.8 MMBtu/hr to 304.5 MMBtu/hr and minimizes the impact of using low-sulfur replacement fuels. Finally, Industrial Fuel Oil contains too much sulfur to be considered as a viable replacement fuel. Use of the Industrial Fuel Oil would actually increase overall GPGP SO_2 emissions by 17% in Alternate 1 (Case 4) and 5% in Alternate 2 (Case 7).

8.4 Desulfurization of Flue Gases from the Riley Boilers

This general subject was completely covered in Sections 3 and 4 of the Sulfur Recovery Technical Committee Report which was excerpted in their entirety in Appendix A. It is this information that should be adjusted herein for the Flue Gas Desulfurization (FGD) cases using replacement fuels. These adjustments include corrections for capacity and performance reductions owing to the use of liquid fuels rather than coal or lignite which the Wyodak plant uses. (The Wyodak facility was used as the model in the Sulfur Recovery Technical Committee Report.) However, based upon the information presented in the Committee report, we concluded that the utilization of an FGD system is not warranted because of cost and retrofit considerations. Consequently, the most appropriate solution to the control of SO_2 emissions from the main stack in the event of jet fuels production is to procure replacement fuels with low sulfur contents providing the cost parameters of the Amoco study are met.

9.0 Design and Cost of Required Process Integration Facilities

The use of crude or fuel oil would require the installation of unloading facility, storage tanks with dikes, a pumping station, and heat traced piping from the tanks to the burners. No modifications to the burner guns are expected. A gas refrigeration system will also be installed to chill the hydrotreater and hydrocracker offgas streams to remove condensible hydrocarbons. This is required before scrubbing with Sulfolin solution in a dedicated absorber to remove H_2S in order to avoid fouling problems. Offgas refrigeration is also proposed to be implemented at GPGP for the Phosam offgas and lock gas as part of the BACT and is discussed more fully in Appendix B.

The use of SNG from the plant would require the installation of piping from the source to the burner front and the change-out of the oil atomizers for gas burners. Installation of a Sulfolin gas treatment system would still be required to desulfurize the hydrotreater and hydrocracker offgas streams prior to their use as boiler fuel.

For the SNG case, ANG has indicated that sufficient capacity exists to regenerate the Sulfolin solution in their equipment, but that a separate chiller, absorption column, knock-out drums, and pumps will be required to treat the offgas from the Jet Fuels/Chemicals Production Facility.

9.1 Facility Design and Investment Estimate

9.1.1 Replacement Fuel Handling

For the case where oil is used as replacement fuel, we assumed that the oil would be received in railroad tank cars at the GPGP plant. Equipment required for handling the oil would be, at a minimum, a tank car unloading station, transfer pumps, an elevated and heated pipe line transferring the oil to a storage tank located adjacent to the Riley Boilers with leak detection equipment, and metering pumps feeding steam atomized burners at the Riley Boiler base. The basic equipment is shown in Figure 7, while Table 26 provides information on the design of these facilities.

9.1.2 Gas Refrigeration/Desulfurization

As discussed previously, the offgas produced in the Jet Fuel/Chemicals Production Facility must be refrigerated and then desulfurized before it can be utilized as boiler fuel. The equipment required to accomplish this would be, at minimum, a unit to chill the gas to $40^{\circ}F$, a knock-out drum and condensate pump, a trim gas reheater, and a packed absorber with transfer pumps for the lean and rich Sulfolin solution. The basic flow plan depicting the equipment layout is presented in Figure 8, while Table 27 provides some basic design parameters for these facilities. As described in Table 27, the equipment is quite small, even though all three gas streams are combined for
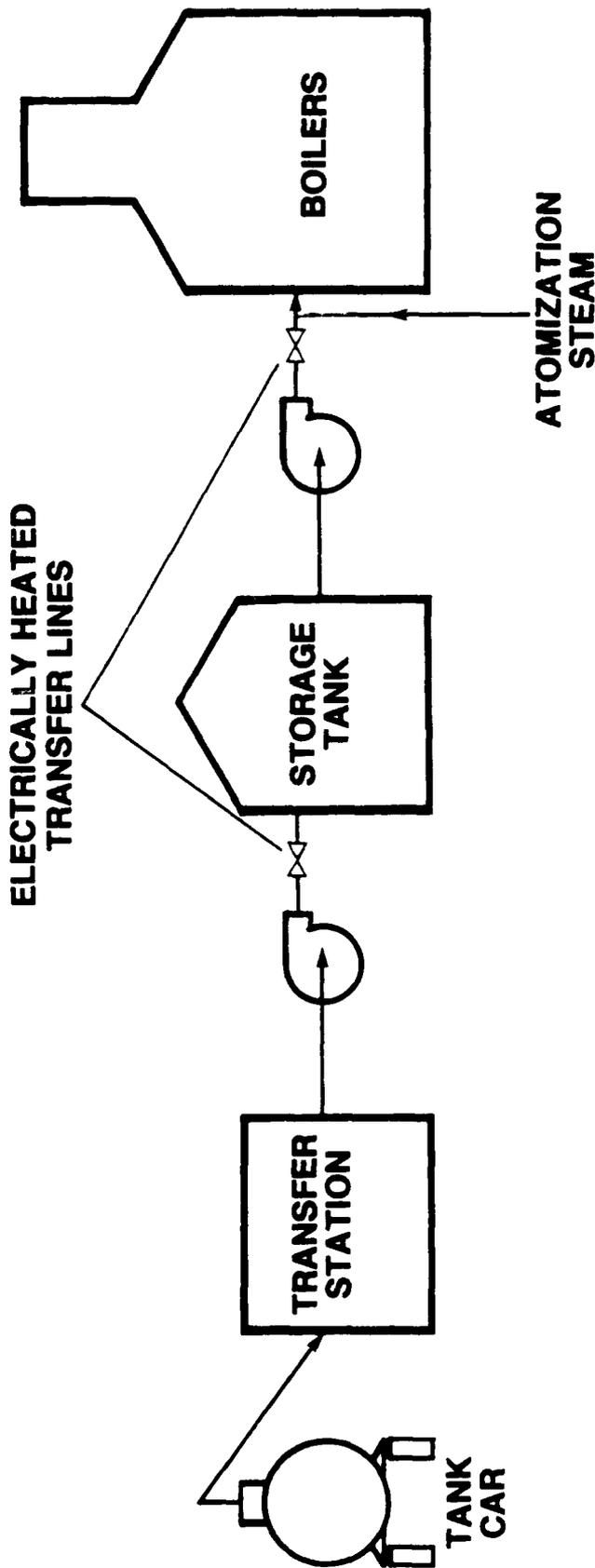
FIGURE 7. HANDLING AND STORAGE FACILITIES FOR REPLACEMENT FUEL OIL

TABLE 26 - DESIGN PARAMETERS FOR FUEL HANDLING AND STORAGE FACILITES

|	
Transfer Station	10' x 20' Prefab Building
Oil Pumps	1000 GPM - 2 required
Heated Transfer Pipe	Electric Heated 12" Diameter
Storage Tank	57' Dia x 40' High; 750,000 gallons, 2 required
Metering Pumps	2 required
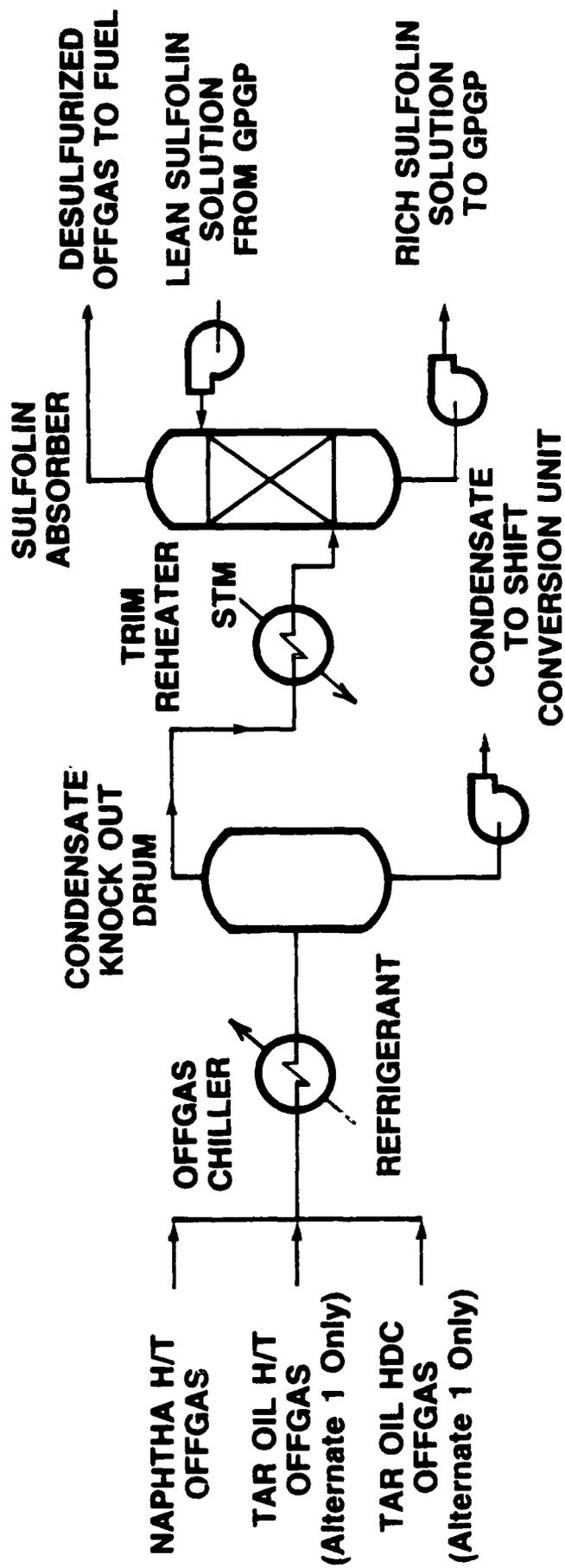
FIGURE 8.   OFFGAS REFRIGERATION AND DESULFURIZATION FACILITIES

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TABLE 27 - BASIC DESIGN PARAMETERS FOR ALTERNATE 1 OFF-GAS REFRIGERATION AND
DESULFURIZATION FACILITIES

1. Off-gas Chiller

No. Required		1
Duty	Btu/hr	56,750
Flowrate	lb/hr	2,062
	ACFS	4.2
Area	ft^2	~250
Tin/Tout	°F	90/40
ΔP	psia	2

2. Condensate K.O. Drum

No. Required		1
Vapor Flowrate	ACFS	4.0
Liquid	lb/hr	425 (1.36 gpm)
Diameter		2'-3"
Height		3'

3. Off-gas Trim Heater

No. Required	1
Flowrate (inlet)	4.0 ACFS
Tin/Tout	40/70
Area ft^2	120
Duty	~35,000 Btu/hr

4. Sulfolin Absorber

No. Required	1
Flowrate	4.7 ACFS
Tin/Pin	70°F/34 psia
Diameter	2 ft
Height	50 ft

processing together as is the case in Alternate 1. In Alternate 2, only the naphtha is hydrotreated and the offgas flow from the naphtha hydrotreater is only about one-tenth of that produced and processed in Alternate 1 (11.2 M/H vs. 101.1 M/H). However, this stream contains approximately 104 lb/hr of H_2S which could ultimately represent up to 196 lb/hr of SO_2. Consequently, this stream must be treated before sending it to the fuel system. The basic design parameters to treat this stream are presented in Table 28 and this system conceivably could be fabricated as a skid-mounted unit. Such a determination should be performed during a detailed design of these facilities.

9.1.3 Cooling Towers

Although the cooling towers are not required as a direct result of the inclusion of SO_2 emissions control related facilities, they will be needed to supply the Jet and Fuels/Chemicals Production Facility cooling water demand. Consequently, costs were developed for the cooling towers as a part of this study.

9.2 Overall Investment Requirements

Based on the design information developed, capital costs (in 1988 dollars) were estimated for the handling of two fuels and are presented for each of the two alternates in Table 29. The major capital cost associated with the use of an alternate liquid fuel is for the facilities required to unload, transfer, and store the fuel at the site. This represents over 80% of the total investment requirements of Alternate 1 (Case 4) and 95% of the investment required for Alternate 2 (Case 7). Futhermore, the cost of installing fuel receipt and handling facilities is largely insensitive to the amount of fuel supplied to the plant. In any event, the investment required to integrate these facilities with the GPGP is relatively small compared to the approximately $80 million required for the Alternate 1 (Case 4) Jet Fuels/ Chemicals Production Facility and $36.7 million required for the Alternate 2 (Case 7) Chemicals Production Facility.

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TABLE 28 - BASIC DESIGN PARAMETERS FOR ALTERNATE 2 OFF-GAS REFRIGERATION AND DESULFURIZATION FACILITIES

1. Off-gas Chiller

No. Required		1
Duty	Btu/hr	9,000
Flowrate	lb/hr	328.7
	ACFS	0.31
Area	ft^2	~250
Tin/Tout	°F	100/40
ΔP	psia	2

2. Condensate K.O. Drum

No. Required		1
Vapor Flowrate	ACFS	0.3
Liquid	lb/hr	35
Diameter		8"
Height		3'

3. Off-gas Trim Heater

No. Required	1
Flowrate (inlet)	3.0 ACFS
Tin/Tout	40/70
Area ft^2	9
Duty	~3,000 Btu/hr

4. Sulfolin Absorber

No. Required	1
Flowrate	0.3 ACFS
Tin/Pin	70°F/34 psia
Diameter	8" (assume 1 fps)
Height	30 ft

TABLE 29 - CAPITAL COST SUMMARY

	Alternate 1	Alternate 2
With Oil as Replacement Fuel		
Oil unloading, transfer lines and pumps, tanks, etc.	2,500,000	2,250,000
Sulfolin treatment (chiller, packed tower, pumps, piping, etc.), gas burners	285,000	125,000
SUBTOTAL	2,785,000	2,375,000
Cooling Towers	250,000	185,000
TOTAL	3,035,000	2,560,000

	Alternate 1	Alternate 2
Alternate 3 - SNG as Replacement Fuel		
Gas Burners, Piping & Valves	150,000	150,000
Sulfolin Treatment	285,000	125,000
SUBTOTAL	435,000	275,000
Cooling Towers	250,000	185,000
TOTAL	685,000	460,000

9.3 Operating Costs

Annual operating costs were determined for the Fryburg/Medora crude mixture
and for SNG from the GPGP for both alternatives. The assumptions that went
into developing the operating costs are presented in Table 30, while
Table 31 presents the results of the analysis. Table 32 presents the same
analysis adjusted to include the effect of installing the cooling towers has
on fuel cost.

In both tables, the capital charges and other operating costs are relatively
insignificant compared to the purchase cost of the replacement fuels.
Consequently, the price of the replacement fuel will control the overall
cost of integrating the Jet Fuels/Chemicals Facilities with the GPGP. From
the table, the use of oil as the replacement fuel is the most cost-effective
and just meets the cost criteria of the Amoco/Lummus Crest report. If a
delivered price of $16/bbl could be negotiated for the Medora, Fryburg, or
Medora/Fryburg crudes, then the use of these crudes would meet the afore-
mentioned cost criteria. Similarly, if the cost of the fuel SNG could be
reduced to under $3.00/MMSCF, then it, too, would meet the fuel cost
criteria.

TABLE 30 - OPERATING COST CALCULATION BASIS

Assumptions:

1. Capital charges at 16.13%/year

2. Power at 5¢/kWh

   Oil - 200 kW
   Gas Compression - 100 kW

3. Steam at $0.0025/lb

4. Oil at $17/bbl and $15/bbl delivered

5. SNG at $3.79/MMBtu and
   $2.59/MMBtu (HHV ~ 1045 Btu/cu ft)

TABLE 31 - ANNUAL OPERATING COSTS - REPLACEMENT FUEL CASES AT 90% OPERATING FACTOR,
EXCLUDING COOLING TOWERS

Replacement Fuel	Oil at 15$/bbl	Oil at 17$/bbl	SNG at 3.79$/MMSCF	SNG at 2.59$/MMSCF
Alternate 1				
Capital Charges	$ 449,220	$ 449,220	$ 70,165	$ 70,165
Power	87,600	87,600	10,000	10,000
Water/Steam	120,000	120,000	-----	-----
Labor	40,000	40,000	-----	-----
Fuel	21,478,437	24,342,229	3i,192,108	21,315,907
Annual Cost	$22,175,257	$25,039,049	$31,272,273	$21,296,144
Cost/MMBtu	$ 2.69	$ 3.04	$ 3.80	$ 2.60
Alternate 2				
Capital Charges	$ 383,088	$ 383,088	$ 44,358	$ 44,358
Power	32,200	32,200	3,000	3,000
Water/Steam	34,222	34,222	-----	-----
Labor	40,000	40,000	-----	-----
Fuel	6,265,144	7,100,497	9,098,750	6,217,756
Annual Cost	$ 6,754,654	$ 7,590,007	$ 9,145,928	$ 6,265,114
Cost/MMBtu	$ 2.81	$ 3.16	$ 3.81	$ 2.61

TABLE 32 - ANNUAL OPERATING COSTS - REPLACEMENT FUEL CASES AT 90% OPERATING FACTOR, INCLUDING COOLING TOWERS

Replacement Fuel	Oil at 15$/bbl	Oil at 17$/bbl	SNG at 3.79$/MMSCF	SNG at 2.59$/MMSCF
Alternate 1				
Capital Charges	$ 489,546	$ 489,546	$ 110,491	$ 110,491
Power	221,844	221,844	99,515	99,515
Water/Steam	124,384	124,384	2,923	2,923
Labor	40,000	40,000	-----	-----
Fuel	21,478,437	24,342,229	31,192,108	21,315,979
Annual Cost	$22,354,211	$25,218,002	$31,405,036	$21,528,907
Cost/MMBtu	$ 2.72	$ 3.06	$ 3.82	$ 2.62
Alternate 2				
Capital Charges	$ 412,928	$ 412,928	$ 74,198	$ 74,198
Power	166,444	166,444	89,515	89,515
Water/Steam	38,606	38,606	2,923	2,923
Labor	40,000	40,000	-----	-----
Fuel	6,265,144	7,100,497	9,098,570	6,217,756
Annual Cost	$ 6,923,122	$ 7,758,475	$ 9,265,206	$ 6,384,392
Cost/MMBtu	$ 2.88	$ 3.23	$ 3.86	$ 2.66

10.0 Conclusions

Review of the impact integrating the proposed Jet Fuels and Chemicals Production Facilities will have on GPGP SO_2 emissions and the jet fuels/ chemicals production costs indicates:

1. Hydroprocessing of the GPGP naphtha and tar oil combined with their replacement with a low-sulfur crude oil (such as a Medora/Fryburg blend) and desulfurization of the hydroprocessing offgas can reduce GPGP SO_2 emissions by 7 to 14%.

2. Use of SNG as replacement fuel can significantly reduce SO_2 emissions in Alternate 1 (Case 4) (22%) and Alternate 2 (Case 7) (9%) but is too expensive at $3.79/MMBtu to attain a 10% return.

3. Replacement fuel receipt, storage, and handling facilities represent the bulk of the capital requirement for integrating these facilities.

4. Replacement fuel costs overwhelm the effects of capital charges and other operating costs.

5. Medora/Fryburg crude at $16/bbl will be marginally acceptable relative to Amoco/LCI's guideline that replacement fuel should cost no more than $2.50 to $3.00/MMBtu for the project to attain a 10% rate of return.

6. With the exception of cooling water, integration of utilities, process intermediate, and process waste streams will have minimal impact on the GPGP operation and SNG production as proposed in the Permit Application.

Consequently, integration of either a Jet Fuels/Chemicals or Chemicals only production facility will be marginally attractive if based only on replacement fuel/integration cost. However, there is a fairly significant potential to reduce the GPGP SO_2 emissions through the use of a low-sulfur replacement fuel. Determination of the quantitative cost advantage

associated with the reduction in SO_2 emissions was not within the scope of this study but should be evaluated if interest warrants.

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APPENDIX A

EXCERPT FROM BACT FLUE GAS DESULFURIZATION EVALUATION

3 BOILER STACK EMISSION CONTROL WITH
FLUE GAS DESULFURIZATION (5100 AREA)

3.1 SPRAY-DRYER FLUE GAS DESULFURIZATION

In the lime spray-dryer flue gas desulfurization (FGD) process, the hot flue gas typically enters a cylindrical, conical-bottom spray-dryer vessel. Within this vessel, an atomized slurry of lime and recycled solids meets the boiler flue-gas stream. The SO_2 in the flue gas reacts with the alkaline lime to form a mixture of calcium sulfite ($CaSO_3$), calcium sulfate ($CaSO_4$), and ash. The presence of alkali in the ash may provide considerable removal benefits in a typical coal-fired boiler, but there will be no ash contribution for this application. A fabric filter (baghouse) typically is used for particulate removal. These particulates (at 2% moisture) are partitioned between a recycle portion and landfill, with the recycled powder being used for slurry preparation. The water in the slurry vaporizes and is emitted with the scrubbed gas. By way of contrast, wet lime/limestone FGD systems carry out the same set of reactions with SO_2 in the flue gas, but the scrubbing liquor does not dry out, and the solids are later recovered as $CaSO_4$ in forced-oxidation reaction tanks. Hence, one principal tradeoff between these two systems is that the spray-dryer FGD system will require a fabric filter, while a wet FGD system will require reaction and thickening tanks. Both the H_2S and the organic sulfur in the fuel burned in the Riley boilers are converted to SO_2, which can be removed by FGD.

3.2 GP APPLICATION

If an FGD system were to be installed at the GP facility, it would be advantageous to limit desulfurization to the flue gas from the boilers, because the superheater flue gas accounts for 10% of the total flue gas volume but less than 3% of main stack SO_2 emissions. As pointed out in Sec. 2, it would also be preferable to continue operation of the Sulfolin system and to chill the Phosam and fuel gas streams. However, the substantial expenditures required to effect further improvement in Sulfolin performance could not be justified.

The block diagram for the retrofit of this system appears in Fig. 3.1. The FGD system would have a maximum gas flow rate of approximately 628×10^3 SCFM. This rate is equivalent to a high-sulfur, coal-fired boiler of 250-MW capacity.[1] The ductwork from the three Riley boilers to the stack would be diverted to a set of three spray-dryers and ten fabric filter compartments (eight operating and two on standby); a new stack would also be included (see Sec. 3.3.4). The spent sorbent is pneumatically conveyed to a holding silo at the facility boundary. From here, dustless loaders fill dump trucks that take the spent sorbent to the permanent ash disposal area. It is envisioned that with the Sulfolin system in operation, H_2S removal efficiencies of 85-99% could be achieved. The addition of FGD could reduce H_2S and organic sulfur emissions in the Sulfolin off gas to a level of 1340 lb/hr SO_2.

Any change in the performance of the Sulfolin system would have a direct impact on FGD operating costs (i.e., sorbent utilization) as well as on approximately 50% of the
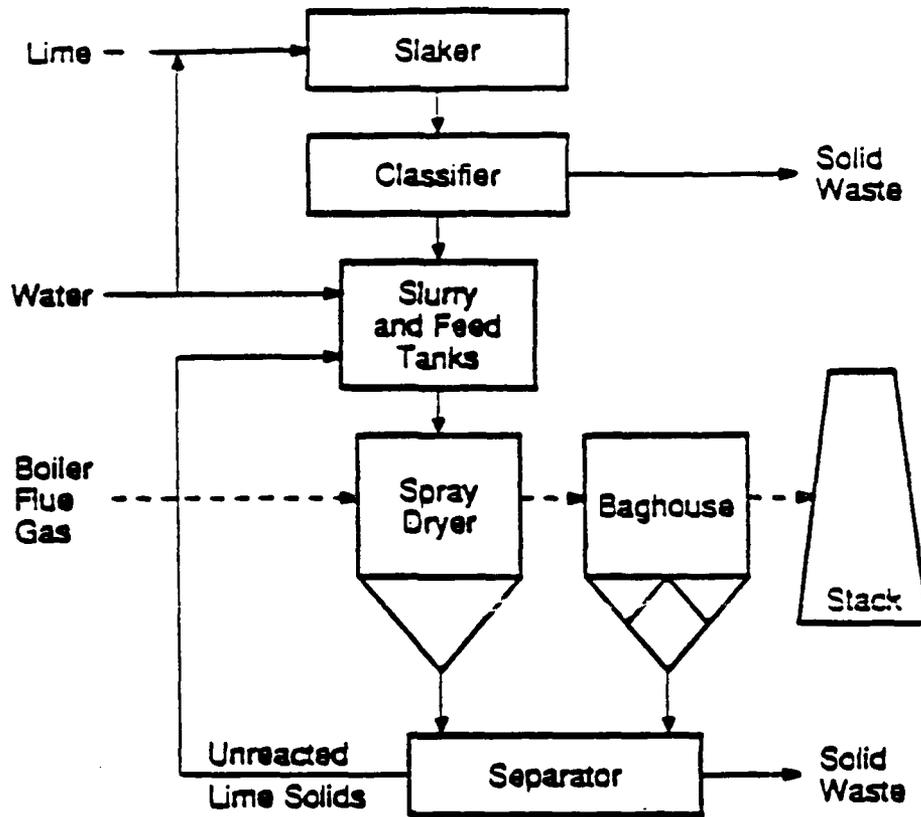
FIGURE 3.1 Retrofit Flue Gas Desulfurization System for Boiler Stack Emissions

capital costs for FGD equipment. This is because the sorbent handling equipment is sized to match the SO_2 load, whereas the flue-gas equipment is sized based on gas flow rates, which are essentially unchanged by Sulfolin performance.

3.3 FEASIBILITY

3.3.1 Advantages

Advantages of a spray-dryer FGD system:

- It is a commercially proven control technology.

- It produces dry powder suitable for environmentally acceptable landfilling or use as a bottom-ash disposal pit stabilizer.

- There is no need for reaction tanks or for a forced oxidation system, as in the wet lime/limestone FGD.

- Water consumption is lower than for other commercial FGD systems.

3.3.2 Limitations

Limitations of a spray-dryer FGD system:

- The high carbon dioxide (CO_2) content of the flue gas would result in poorer sorbent utilization than found with the same-capacity FGD system for a coal-fired boiler.

- Significant flows of finely sized solid materials (both pebble lime and spent sorbent) must be maintained.

- There would be significant retrofit problems in the limited space around the Riley boilers, which may give conventional wet FGD an advantage.

- A minimal cost strategy for tying in the system may require the installation of a new stack.

- Future recovery of CO_2 as a product would not be possible.

3.3.3 Commercial Experience

Spray-dryer FGD technology has been considered principally in view of the leading role that power plants located near Beulah, North Dakota, have played in their

development. The first two commercial systems were located near Beulah at Coyote 1 (440 MW, February 1981) and Antelope Valley 1 (440 MW, May 1983). Basin Electric continued its commitment to this technology with the deployment of Antelope Valley 2 (440 MW) in May 1985. All three of these systems lie within sight of the GP facility. Currently, there are 13 other systems in operation or under contract at electric utilities in this country. These represent an additional total scrubbed capacity of 4,300 MW. Wet lime/limestone FGD systems are represented by 138 utility-scale systems representing a total scrubbed capacity of 58,128 MW. Neither of these technologies has operated on a flue gas with as high a CO_2 content as that from GP. This would translate into somewhat higher sorbent costs for these systems than would be expected in a utility FGD application with a similar flue gas concentration of SO_2.

3.3.4 Retrofit Considerations

Although a retrofit of a spray-dryer FGD system is possible, it clearly would not be a simple or straightforward operation. Some units, normally near grade level, would have to be built on supports above the ground. Joy Manufacturing, a vendor of spray-dryer FGD systems, has retrofitted the lignite-fired boilers at the Wyodak facility in Wyoming. The retrofit difficulty at GP clearly would be worse than at Wyodak, and this difficulty would likely increase costs over Wyodak by nearly 20%. Joy pointed out that the retrofit of the Wyodak facility cost $28 million, not including poured foundations. Wyodak already had a particulate control system, which Joy estimated should cost around $10 million. A good estimate for the cost of foundations would be 20-25% of facility costs. Wyodak is a good model because it is a 330-MW facility, compared to the approximate 250-MW equivalent capacity for the Riley boilers.

A set of three spray dryers would have to be built directly north of the boiler house in the open area above the service road. The ducting connections and bends are both tight and not optimal. Additionally, the existing induced-draft (ID) fans are probably not sufficient for this increased service, and this assumption does not include the contribution from the superheater (about 10% of flow).

The area to the northwest of the GP boiler house was originally open, but some maintenance buildings have since been located in this area. These would have to be relocated or demolished prior to construction. It is in this area that the fabric filters would have to be installed. Ten separate compartments are envisioned. It will be necessary to elevate these fabric filters, and the location would be very tight. An area bordered by the Kaiser battery plot line, the ducting, and the road is about 150 ft × 150 ft. The projected fabric filter area of 150 ft × 100 ft would take up 66% of this area and leave only marginal clearance. A crane would have to be assembled inside the pipe racks circling the $100 area to build the FGD system. Either the outlets from the fabric filters would be ducted on a common manifold to the existing stack, or a minimum cost strategy would require construction of a new stack. Because fabric filters are sized on an air-to-cloth ratio, the physical size of the fabric filters would be similar to those for a 250-MW coal-fired system. No room is available for storing spent sorbent; it would have to be pneumatically conveyed to a location west of the 1300 area. While the severity of

the bends in the flue-gas piping has not been established, the area available is clearly not optimal for this type of system.

The sorbent preparation area (lime-slaking) could be located in an elevated area to the south of the boiler house, possibly in the area above the electrical equipment. This raises safety concerns, since the slaking equipment tends to be a high-maintenance item and spills frequently occur with this type of equipment. A better option would be to locate the fresh lime storage, spent sorbent storage, lime slaking, and recycle sorbent-fresh lime mix tanks west of the 1800 area. This presents no exceptional retrofit challenge, although it has a negative impact on the operating economics because of the need to move large amounts of material between the 5100 area and the 1800 area.

3.3.5 Impact on SNG Production

The ducting from the Riley boilers to the stack would have provisions for bypass in the event of FGD system unavailability. Consequently, no impact on SNG production is envisioned with this control option.

3.3.6 Impact on Sulfur Dioxide Emissions

The spray-dryer/fabric filter system has been designed to treat all of the boiler flue gas and to achieve 90% SO_2 removal. Under these conditions, main stack SO_2 emissions in the base case (160 MMSCFD rate with 1.7%-S coal, 78% Sulfolin H_2S removal) could be reduced to 1169 lb/h; or to 1045 lb/h with chilling of the Phosam and fuel gas streams. The permitted SO_2 emission rate of 1340 lb/h could be reached with SO_2 removal efficiencies of 88.4% and 86.6%, respectively, as shown in Table 3.1.

3.4 ECONOMICS

The economics are consistent with the assumptions in Sec. 1.8. The uncontrolled SO_2 emissions from the Riley boilers are taken from the Table 1.3 for conditions under which the H_2S sulfur control in the 4000 area varies between 80% and 99% for 1.7% sulfur coal and 160-MMSCFD production. The stack emission rate has been set at 1,340 lb/h SO_2.

3.4.1 Capital Costs

The capital costs are based on the installed retrofit costs for the WYODAK plant (see Sec. 3.3.4). These costs have been roughly estimated at $53 million (first quarter, 1987) for high-sulfur coal at 160 MMSCFD. Stearns-Rogers projected costs for a "green-field" system at 1,000-MW power plant burning low-sulfur coal in a recent EPRI study.[2] The costs of the particulate control system were split between those required for flash-ash control and spent sorbent collection in the Stearns-Rogers study. Thus, these costs were reexamined using a peer-reviewed computer code.[3] The overall capital costs was

TABLE 3.1 Sulfur Dioxide Removal Rates and Capital Costs
for Spray-Dryer/Fabric Filter FGD Systems for 1.7%-S Coal
at 160 MMSCFD

System	SO_2 Removal (%)	SO_2 Inlet (lb/h)	SO_2 Removed (lb/h)	Main Stack SO_2 (lb/h)	Capital Cost (10^6)
A[a]	88.4	8731	7687	1340	62.0
B[b]	86.6	7491	6447	1340	64.6

[a] 78% H_2S removal from Rectisol off gas.

[b] 78% H_2S removal from Rectisol and Phosam off gas, 70%
H_2S removal from fuel gas. System cost includes cost
of chilling Phosam and fuel gas streams.

adjusted for size (the 5100 area is about 250-MW equivalent) using a 0.66 factor and a 1.3 retrofit factor. With this adjustment, the total installed cost comes to $65 million (first quarter, 1987). Taking the average of these two figures yields a total plant capital cost of $59 million (first quarter, 1987), and an annualized plant capital cost for the case of 160 MMSCFD and high-sulfur coal of about $9.7 million/yr. Capital costs snown in Table 3.1 include a process contingency to allow for the higher than normal (30%) CO_2 content of the flue gas.

3.4.2 Fixed Costs

The fixed costs should be 3% of total capital; the figures are in Table 3.2.

3.4.3 Maintenance Costs

These would be 5% of total capital (see Table 3.2). Following EPRI's lead in the Stearns-Rogers study, the costs of periodic bag replacements will be a variable operating cost.

3.4.4 Variable Operating Costs

The variable operating costs for this system would include lime, water, solids disposal, fabric filter replacement, and power costs. The spray-dryer/fabric filter FGD system would be designed for 95% availability;[4] typically, a coal-fired boiler in this application would recycle the spent sorbent to maintain the Ca:S ratio at 1.6:1. However, based on the observed calcium carbonate ($CaCO_3$) formation from 15% CO_2 flue gas,[5] the high CO_2 content of the GP flue gas will raise this to 1.8:1. The chemical cost data will match those used in the EPRI Technical Assessment Guidelines discussed in Sec. 1.8, with one exception. A recent EPRI study by ICF[6] has conclud- hat disposal costs for spray-dryer retrofit should be $17.40/ton as opposed to ·   u0/ton used in the guidelines. The growing significance of this issue has been noted by other studies.[7,8] The costs for the specific case of low-sulfur (1%) coal at a 160.0-MMSCFD production level and Sulfolin operating at 90% efficiency are given in Table 3.3. A summary of variable generating costs is found in Table 3.2.

3.4.5 Summary of Economics

A summary of the annual operating costs for the spray-dryer/fabric-filter systems is found in Table 3.2. This includes capital, fixed maintenance, and variable operating costs, designed so that the stack emissions are 1,340 lb/h of SO_2.

TABLE 3.2 Annual Costs for Spray-Dryer/Fabric-Filter FGD System (4000 Area) for High-Sulfur (1.7%) Coal at 160 MMSCFD

System[a]	Annual Capital (10^6/yr)	Fixed Cost[b] (10^6/yr)	Maintenance Cost (10^6/yr)	Variable Operating Costs (10^6/yr)	Total Annual Cost (10^6/yr)
A	9.3	1.9	3.1	13.0	27.3
B	9.7	1.9	3.2	11.8	26.6[c]

[a]See Table 3.1.

[b]The fixed costs include operating labor, administration, and support.

[c]Includes costs associated with chilling the Phosam and fuel gas streams.

TABLE 3.3 FGD Costs for Low-Sulfur (1%) Coal, Production Level 160 MMSCFD, Sulfolin Operating Efficiency, 90%

	Rate (lb/h)	Unit Cost	Annual Cost (10^6/yr)
SO_2 removed	1,354	—	—
Lime	2,820	$65/ton	1.26
Water	111,880	$0.60/1000 gal	0.11
Spent solids	5,590	$17.40/ton	0.38
Fabric filters	—	—	0.75
Power	7,500 kW	$0.05/kWh	2.96
Total			5.46

3.5 REFERENCES FOR SECTION 3

1. Keeth, R.J., J.E. Miranda, J.B. Riesdorf, and R.W. Scheck, *Economic Evaluation of FGD Systems, Vol. 1: Throwaway FGD Processes, High- and Low-Sulfur Coal*, Electric Power Research Institute Report EPRI-3342, Vol. 1 (Dec. 1983).

2. Keeth, R.J., op. cit., pp. 11-17.

3. Swietlik, C., and P. Farber, *Users Guide for PARTIC II (Version II): A Computer Code for Analyzing the Performance of Particulate Control Devices*, Argonne National Laboratory Report ANL/ECT-TM-6 (Sept. 1986).

4. Doctor, R.D., *Utility Flue-Gas Desulfurization: Innovations and System Availability*, Argonne National Laboratory Report ANL/ECT-11 (March 1982).

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7. Burnett, T.A., *Technical Review of Dry FGD Systems and Economic Evaluation of Spray Dryer FGD Systems*, U.S. Environmental Protection Agency Report EPA-600/7-81-014, TVA EDT-127, p. xxxii (Feb. 1982).

8. Bechtel Group, Inc., *Design of Advanced Fossil Fuel Systems (DAFFS): Pulverized Coal-Fired Power Plant with a Lime Slurry Spray Dryer/Fabric Filter Flue Gas Desulfurization System*, Argonne National Laboratory Report ANL/FE-83-12, pp. 6-11 (June 1983).

4 BOILER SULFUR DIOXIDE EMISSIONS CONTROLS
— ADDITIONAL OPTIONS (5100 AREA)

4.1 ALTERNATE COMMERCIAL GAS DESULFURIZATION SYSTEMS

The Electric Power Research Institute (EPRI) maintains an active program for reviewing the economics and performance of FGD systems that might be considered as alternatives to the spray-dryer filter system reviewed in Sec. 3. Stearns Catalytic Corporation is the contractor for this program, and the most recent volume of their results presents a summary of the levelized operating costs for commercially available nonregenerable FGD processes.[1] Thirteen of the technologies surveyed fell within ±12.5% of the median value (the spray-dryer/fabric-filter system was at the median value). From this is may be concluded that no FGD system shows a distinct advantage, and the costs projected in Sec. 3 will be typical of those that could be anticipated from any of the commercially available FGD technologies.

4.2 DEVELOPMENTAL FLUE GAS DESULFURIZATION

Direct injection of calcium-based sorbents into coal-fired steam boilers and the downstream ductwork has been investigated to develop retrofittable FGD systems with low capital costs and moderate SO_2 removal capability. These were previously reviewed for application at the GP facility and not recommended. Four basic variations of sorbent injection have been identified: solid sorbent may be injected directly into either the lower or the upper regions of the boiler, or the sorbent is injected into the ductwork downstream of the air heater as either a solid or a slurry. In all of the variations, flue gas humidification at tail-end temperatures (below 600°F) is essential for effective SO_2 removal and sorbent utilization. The importance of flue gas humidification for enhancing calcium sorbent and SO_2 reactions has been observed in spray dryers. Fabric filters treating a flue gas humidified to within a 20°F approach to adiabatic saturation removed substantial amounts of the remaining SO_2.

The probability of successfully applying dry sorbent in-duct injection is low in terms of achieving adequate SO_2 removal, for the following reasons.

- Testing in the United States has been limited to coal-fired boilers at bench-scale or pilot-plant levels, and numerous uncertainties regarding design features and the effects on boiler performance remain.

- The potential SO_2 removal efficiencies are not well established, and the effects of high CO_2 flue gas concentrations on removal have not yet been investigated.

- Capital and operating costs would be high, due to the need to install a baghouse and a new main stack and to use large amounts of expensive sorbents in a boiler that operates almost constantly.

Applying sorbent injection technology at GP represents a high-risk, high-cost alternative. Numerous technical uncertainties remain and would require significant research prior to serious consideration. These technical uncertainties include dispersion effectiveness of the solid sorbent, numerous issues regarding the appropriate injection equipment (Riley stoker), SO_2 removal efficiency and gas compatibility (Riley stoker), short residence time, and the lack of data from commercial applications comparable to the GP facility.

The following sections present technical descriptions of the sorbent injection variations, the SO_2 reductions achievable, advantages and limitations, developmental status, probability of success, and estimated costs (if available). The probability of success must take into account the likelihood of attaining reliable operation, permissible SO_2 emission levels, and levelized costs that do not jeopardize sale of the facility.

With regard to the first condition, it is important to note that (1) none of the variations has yet been demonstrated at commercial scale in the United States; (2) each would require the addition of a full-scale (250-MWe) particulate collection device; and (3) most of the existing data apply to coal-fired boilers with different temperature profiles, and lower CO_2 and SO_2 flue gas concentrations than are present in the 5100 area boilers and superheaters. Consequently, extensive investigation would be required before any of the sorbent injection processes could be implemented at GP with reasonable assurance of reliable operation. With respect to the second condition, the maximum SO_2 emission level that would be permitted by North Dakota remains to be determined. Currently the facility has been allocated 2640 lb/h SO_2. The New Source Performance Standards (NSPS) for oil-fired steam boilers is 0.8 lb/MM Btu; for coal-fired boilers, it is 1.2 lb/MM Btu.

The SO_2 emission reduction required to comply with the 2640 lb/h allocation varies with the gas production rate and the sulfur content of the coal. It ranges from 50% (at a production rate of 148 MMSCFD with the current source of coal) to 75% (at a production rate of 160 MMSCFD using coal with the maximum expected sulfur content).

4.2.1 Lower Boiler Sorbent Injection

In this process, sorbents such as limestone, dolomite, or their hydroxides are injected directly into the lower boiler. On exposure to the high temperatures, the sorbents are rapidly converted to reactive lime particles that combine with SO_2 to form $CaSO_4$. The $CaSO_4$ and unreacted lime are swept out of the furnace by the flue gas and collected, along with fly ash, in the particulate control device. The limited data available indicate that SO_2 reduction is limited to about 50% at practical Ca/S ratios. The main advantage of lower boiler injection is the possibility of using limestone, the lowest-cost sorbent.

Major issues in applying this process at Great Plains include the unique boiler characteristics; the possible need for modifications to the three boilers, causing disruption of production; potential problems with boiler convective section fouling; and solid waste handling and disposal. Unreacted sorbent occurs as quicklime (CaO) in the

collected solids and may require special waste handling to accommodate high pH, heat of reaction with water, and cementitious behavior if disposed of with coal ash.

A number of research and development projects on lower boiler injection are being sponsored by EPRI, EPA, DOE, and industrial organizations. The programs cover a wide range of laboratory and field test conditions with various sorbents, coal types, and injection system designs.

Although a number of power stations in Europe are operating boiler injection systems for emission compliance, the unique design features of their boilers do .ot permit direct extrapolation of their results to U.S. boilers nor to flue gas with moderate to high SO_2 concentrations. However, two utility scale demonstration projects in the United States are scheduled to start up in mid-1987: (a) 60-MW tangential-fired prototype at Whitewater Valley and (b) 105-MW wall-fired demonstration unit at Edgewater No. 4 in Ohio.

It is clear that this approach does not achieve adeq ate SO_2 removal and that the probability of success is very low. Furthermore, serious technical questions remain to be answered.

4.2.2 Upper Boiler Sorbent Injection

In this process, powdered hydrated lime or half-calcined dolomite is injected into the upper part of the boiler, near the economizer inlet, at around $1,000°F$. The sorbent must be mixed rapidly with the flue gas since most of the SO_2 capture occurs during the short time that the sorbent and flue gas flow through the steep temperature gradients of the economizer. Supplemental SO_2 capture may occur in the duct between the air preheater and the boiler and in the particulate collection device. Because sorbent reactivity appears to slowly decrease at typical duct temperatures, performance may be improved by the use of hydration techniques currently being developed to increase surface area and porosity and decrease particle size. With the best experimental sorbent produced so far, 70% removal of SO_2 was achieved at a Ca/S ratio of 2 when burning coal with close to 2% sulfur.

The advantages of upper boiler injection include the opportunity to benefit from the use of the highly reactive sorbents now being developed. the lack of effect on heat transfer in the lower boiler zone and in the superheater and reheater tube banks, and the virtual absence of reactive CaO in the waste. The limitations include potential fouling problems in the economizer and preheater, the possible need for boiler modifications, and the absence of information from large-scale tests.

The only known development work on this process is the EPRI-sponsored, laboratory testing on a 10,000-Btu/h isothermal bench-scale reactor at KVB, Inc. Recent pilot-scale tests have confirmed the bench-scale results.

The probability of success in applying upper boiler injection at GP is low in view of the immaturity of the technology and the adverse effects of high CO_2 flue gas concentrations on SO_2 removal.

4.2.3 In-Duct Sorbent Injection

In this process, powdered hydrated lime is injected into the flue gas duct upstream from a particulate control device. SO_2 capture takes place within the duct and in the particulate control device, as on a fabric filter dust cake. Humidification with steam is necessary since SO_2 removal requires high relative humidity in the flue gas. The humidity, however, must be carefully controlled to avoid condensation in the duct(s) and particulate control devices. This is done by maintaining the temperature $10°$ to $50°F$ above the flue gas adiabatic saturation temperature. Sorbent recycle and the concurrent injection of chemical additives are being investigated to improve sorbent utilization. With the use of recycle, steam, and fabric filter collection, a 70% SO_2 removal at a Ca/S ratio of 2 has been achieved in small-scale tests.

The duct injection process is being intensively investigated by both industrial and government R&D organizations. Laboratory and pilot plant tests have produced promising results. Large-scale applications of the process are also being evaluated and supported by the DOE Clean Coal Program. Routine industrial applications in the United States have not yet been demonstrated on a large scale, but a variation of the concept is being demonstrated at a 250-MW coal-fired plant in Finland. The process injects limestone into the upper furnace cavity, followed by humidification of the particulate-laden stream after the air heater. At a Ca/S ratio of 3.8 and an approach to adiabatic saturation within 10 to $35°F$, SO_2 reductions ranged from about 30 to 70%. The ducts were modified to increase the gas residence time.[2]

The installation of an in-duct control process would be complicated, because the flue gas streams with significant quantities of SO_2 are delivered to the stack by two ducts, one of which receives input from three upstream ducts. Since a particle control device (baghouse) would also be required to collect reacted sorbent, a significant rerouting of ducts would be required; however, special provisions (such as increased residence time) could be incorporated as an added advantage of an in-duct control method.

Additional design flexibility may be possible since almost all of the SO_2 is produced by the boilers. If sufficiently high SO_2 removal can be attained on only boiler flue gas, the need for treating superheater flue gas may be eliminated, giving design options such as gas bypass for reheating and allowing operation close to adiabatic saturation, which enhances SO_2 removal while protecting the downstream baghouse.

The advantages of in-duct removal processes:

- They are suitable for space-limited retrofit applications.

- Sulfur dioxide control is focused on one or two gas streams at a process location prior to release to the environment; thus, it would also represent an emission control backstop to other plant control processes.

- The flue-gas stream contains little ash from fuel combustion; thus, recycling of partially reacted sorbent may be possible, which would result in increased sorbent utilization and reduced operating cost.

- The waste for disposal is dry.

The limitations of an in-duct sulfur dioxide removal processes:

- Large-scale applications have not been demonstrated.

- A baghouse is required for particulate control.

- At present, uncertainties exist in design and placement of injection equipment, ultimate SO_2 removal, sorbent utilization values (reagent selection and preparation techniques), and possible sorbent deposition in ducts.

- A possible negative effect exists from high CO_2 levels (compared to coal combustion) in the flue-gas stream, which may reduce SO_2 removal and sorbent utilization.

- Fouling problems could occur, and boiler modifications in the furnace injection approach might be needed.

- Only a short residence time is available for SO_2 absorption.

4.2.4 Slurry Injection

In this approach, a rotary (dual fluid) slurry atomizer is installed in the ductwork upstream from a particulate control device. A lime slurry is atomized into the duct and removes SO_2 as it dries. The concept is similar to commercial spray-drying FGD systems, except the residence time for drying provided by the existing ductwork is only 1-2 s, compared with the 10-12 s in typical spray dryers. As in the conventional spray-drying FGD systems, additional SO_2 removal occurs in the baghouse. In early EPRI tests, 50% SO_2 removal was achieved at a Ca/S ratio of 1.5 and a 40°F approach to saturation. Sorbent recycle may lead tr improved removal and utilization.

The advantages of in-duct spray drying for FGD:

- Capital and installation costs are less that those associated with tail-end spray drying systems.

- It has good sorbent utilization with moderate SO_2 removal.

- The waste is dry and does not require special handling.

The major limitations:

- There is a high potential for wet solids deposition in the ductwork.

- No proof-of-concept data currently exist.

Developmental efforts on in-duct spray drying techniques are being conducted by DOE, EPRI, EPA, and industrial organizations. DOE is the main sponsor of this technology, with two pilot-plant projects (5 and 12 MW) scheduled for startup late in 1986.

The probability of success for applying in-duct spray drying is relatively low because the potential impacts of solids deposition, i.e., reduced production and inadequate SO_2 removal, remain.

4.2.5 Sorbent Injection Economics

Estimated costs of lower boiler sorbent injection have been reported by EPRI for a 1,000-MW coal-fired power plant with two 500-MW boilers and a capacity factor of 65%.[3] Since the flue gas flow to the stack from the three 5100 area boilers and two superheaters, including the waste gas, is nominally equivalent to that from a 250-MWe coal-fired power plant, appropriate scaling factors for costs associated with the boiler and baghouse must be applied to develop estimated costs for implementation at Great Plains. The reported capital costs of $70/kW for a baghouse, $25/kW for boiler modifications, and $45/kW for ancillaries correspond to a total cost of $70 million for a 500 MW_e unit, and scale to a capital cost of $41.3 million for a system treating the GP boiler flue gas at a flow rate equivalent to 225 MW_e. Variable operating costs for 50% SO_2 removal from boiler flue gas containing 7491 lb/h SO_2 (see Table 3.1) are $11.3 million/yr, not including capital charges. These costs are based on 50% utilization of hydrated lime at CA/S = 2 and correspond to an incremental SNG cost of $0.23 per decatherm. Main stack SO_2 emissions for the system are 4042 lb/h, and the emission reduction is 6869 lb/h, or 27,500 ton/yr. With the inclusion of capital charges and fixed operating costs, calculated on the basis of EPRI guidelines, the annual operating costs are $20.8 million, and the cost-effectiveness of emission reduction is $756/ton SO_2 removed.

A summary of estimated cost factors for in-duct sorbent injection was developed by Morgantown Energy Technology Center (METC) in Ref. 2. These economics should be considered preliminary.

4.3 REFERENCES FOR SECTION 4

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APPENDIX B

EXCERPT FROM BACT OFFGAS REFRIGERATION EVALUATION

20 PHOSAM OFF GAS REFRIGERATION

20.1 SYSTEM DESCRIPTION

The option of refrigerating the lock gas (or fuel gas) for the Lurgi gasifiers has previously been proposed by ANG. Certain feedstreams (phosam off gas, the naphtha stripper overhead, and lock gas streams) created operating problems (e.g., too light a sulfur froth and very fine sulfur particles) for the Stretford unit, which has now been converted to Sulfolin. Such problems have been attributed to organic trace compounds. Cooling the feedstreams to ambient or refrigerated temperatures is expected to condense and wash out these compounds. Sulfur pilot-plant work demonstrates that this pretreatment step, in which the streams are cooled to 40°F, improves the composition of these streams sufficiently to achieve acceptable processing through the Sulfolin unit. Although the elimination of contaminants via cooling and refrigeration seems to be a viable approach, the original plant design was for Phosam waste gas to go to the Stretford Unit at 110°F. The problems in the Phosam Unit reboiler and subsequent change to spargers resulted in an elevated temperature of 220°F. An alternate method to solve the problem would be to install reboilers with the proper design and metallurgy and thus achieve cooling of the Phosam stream at 110°F. However, evaluation of this alternative is beyond the scope of this report.

20.2 PHOSAM AND FUEL GAS COOLING — CASE 1

The Phosam stream is cooled to 110°F in an exchanger (for comparison both air- and water-cooled exchangers have been cost estimated). Most of the water ammonia and a part of the organic species are condensed and sent to the liquid waste incinerator (LWI), or to Gas Liquor Separation. The gaseous stream enters a chiller and is cooled to 40°F by ammonia from a dedicated refrigeration unit. The cooled Phosam stream is introduced to a knock-out pot, then steam heated to 110°F before entering the Stretford/Sulfolin unit.

The lock gas, which makes up the fuel gas in the plant, consists of a low-pressure and high-pressure stream as the lock gas from the gasifiers is let down in two steps. The low-pressure lock gas is compressed in the existing compressor to 28 psig and cooled to 110°F before the two streams are joined. The common stream is cooled to 40°F with refrigerant. The cooled stream is introduced to a knock-out pot, then heated to 110°F and piped to the fuel gas venturi and absorber of the Stretford/Sulfolin unit.

The condensates are treated in the LWI. Because the Phosam stream contains appreciable NH_3 (about 16.5 T/D), the NH_3 could be recovered and marketed with that obtained in the Phosam unit. However, NH_3 recovery is not included in this evaluation.

The overall cost of lock gas and Phosam gas chilling may be reduced by using a cold methanol slip stream from Rectisol in place of a refrigeration unit.

20.3 FEASIBILITY — CASE 1

20.3.1 Advantages

Refrigeration is simple and no unproven processes or equipment are used. Experiments at GP have proven that eliminating the contaminants will eliminate emulsion formation and foaming and improve the sulfur quality.

20.3.2 Limitations

The approach does not affect the sulfur plugging tendency in the absorbers, and the sulfur reduction will depend on the H_2S removal efficiency of the Stretford/Sulfolin process.

20.3.3 Commercial Experience

All the equipment has been used in commercial processes, although the process scheme for this particular application has only pilot scale experience.

20.3.4 Impact on SNG Production

During installation of this system, a minor reduction in SNG production may occur with no significant impact on overall rates.

20.3.5 Impact on Sulfur Dioxide Emissions

The Phosam gas at a 160 MMSCFD plant SNG production rate with high sulfur coal (1.7% by wt) feedstock and without any H_2S removal, contributes 951.8 lb/h SO_2 to the total plant emission; under similar conditions the fuel gas contributes an additional 737.5 lb/h of SO_2. Of the Phosam emission 16.7 lb/h of SO_2 is from organic sulfur and for the fuel gas 47.8 lb/h comes from organic sulfur. It is doubtful that refrigeration will eliminate even part of the organic sulfur; therefore, only the H_2S-derived sulfur will have the potential to be reduced in the Sulfolin/Stretford Unit. If 90 percent of the H_2S-derived sulfur in these streams is eliminated via Sulfolin processing, the Phosam stream sulfur emission would decrease by 841.7 lb/h of SO_2 and the fuel gas sulfur emission would decrease by 620 lb/h of SO_2.

20.4 PHOSAM CHILLING — CASE 2A, CASE 2B

The Phosam off gas cooling and chilling can be achieved with two different approaches. The current high superstill overhead temperature and vapor load are caused by the substitution of a shell and tube reboiler with live steam injection. Last year, Fluor completed a design for additional cooling of the overhead system (see Fig. 20.1).
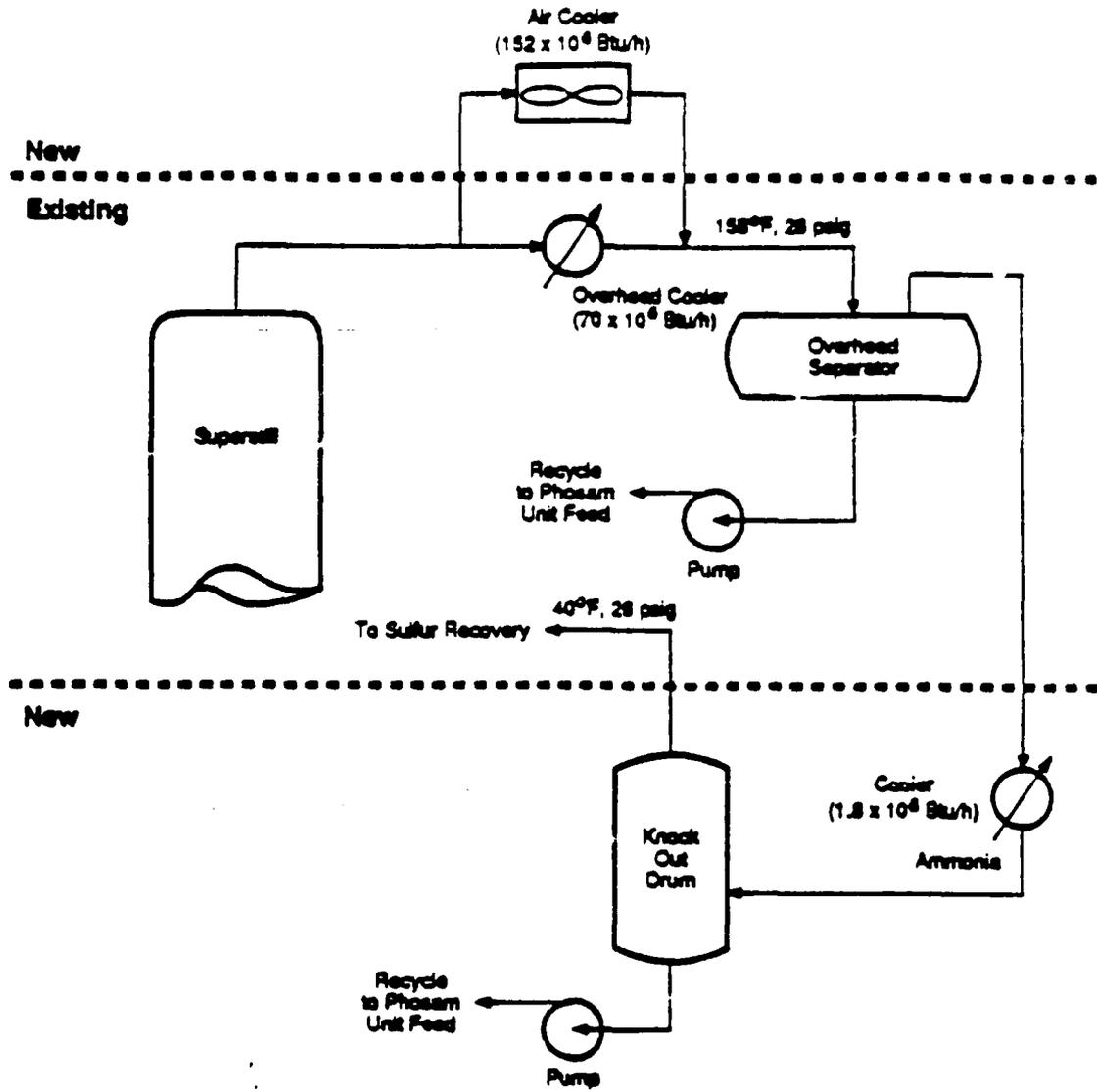
FIGURE 20.1 Phosam Chilling — Case 2A

Alternatively, the ammonia stripper reboiler can be replaced by one with an improved design and upgraded materials of construction. In this case, the overhead system additions will be reduced in scope (see Fig. 20.2). It appears as if the latter approach will be lower in cost. However, additional technical and economic evaluations are in progress, and the economics will be based on the first approach. The alternative arrangement will only be used if further study confirms the technical soundness and the economics are favorable.

The following descriptions explain the two possible arrangements in more detail.

20.4.1 Case 2A — Additional Cooling Plus Refrigeration

The basis for the Phosam overhead cooling system is the 160 MMSCFD Debottlenecking Study completed by Fluor in early 1987 (Fig. 20.1). Fluor was assisted in this study by UEC, the designer of the Phosam Unit. UEC calculated a new heat and material balance for the increased unit throughput, taking into account the present use of a steam sparger in the ammonia stripper rather than a reboiler.

The use of steam sparging in the ammonia stripper results in an increased water load in the absorber section of the superstill. In order to keep the absorber section within its mechanical limitations (flooding) and still drive the extra water overhead, the required process conditions (pressure and temperature) at the top of the superstill had to be raised.

Since the extra water from steam sparging is drive overhead, the required heat duty in the overhead condenser system is increased substantially. The combined duty of the parallel air cooler and existing cooling water exchanger is approximately 222 MMBtu/h, where the overhead stream is cooled to $158°F$ and the majority of the water is condensed.

The condensed liq is separated from the gas via the existing overhead separator. The gas from the separator is cooled further to $40°F$ via an ammonia chiller. The estimated duty is approximately 1.8 MMBtu/h. The condensed liquid from the chiller is knocked out via a knockout drum before the noncondensable gases are sent on to the Sulfolin Unit for treating.

20.4.2 Case 2B — Ammonia Stripper Reboiler Addition Plus Refrigeration

Phosam overhead cooling requirements can be substantially reduced by using an ammonia stripper reboiler in lieu of the live stream sparger (Fig. 20.2). This lowers the required heat duty in the overhead condensers from 222 MMBtu/h to approximately 80 MMBtu/h. Installing a reboiler on the ammonia stripper reduces the amount of water being introduced into the absorber section by 150,000 lb/h. This allows the superstill absorber section to operate at a lower pressure and temperature than the live steam sparger case.

At the present 150 MMSCFD SNG rate, it appears that the existing overhead condenser would be able to cool the superstill overhead to $140°F$. To support a
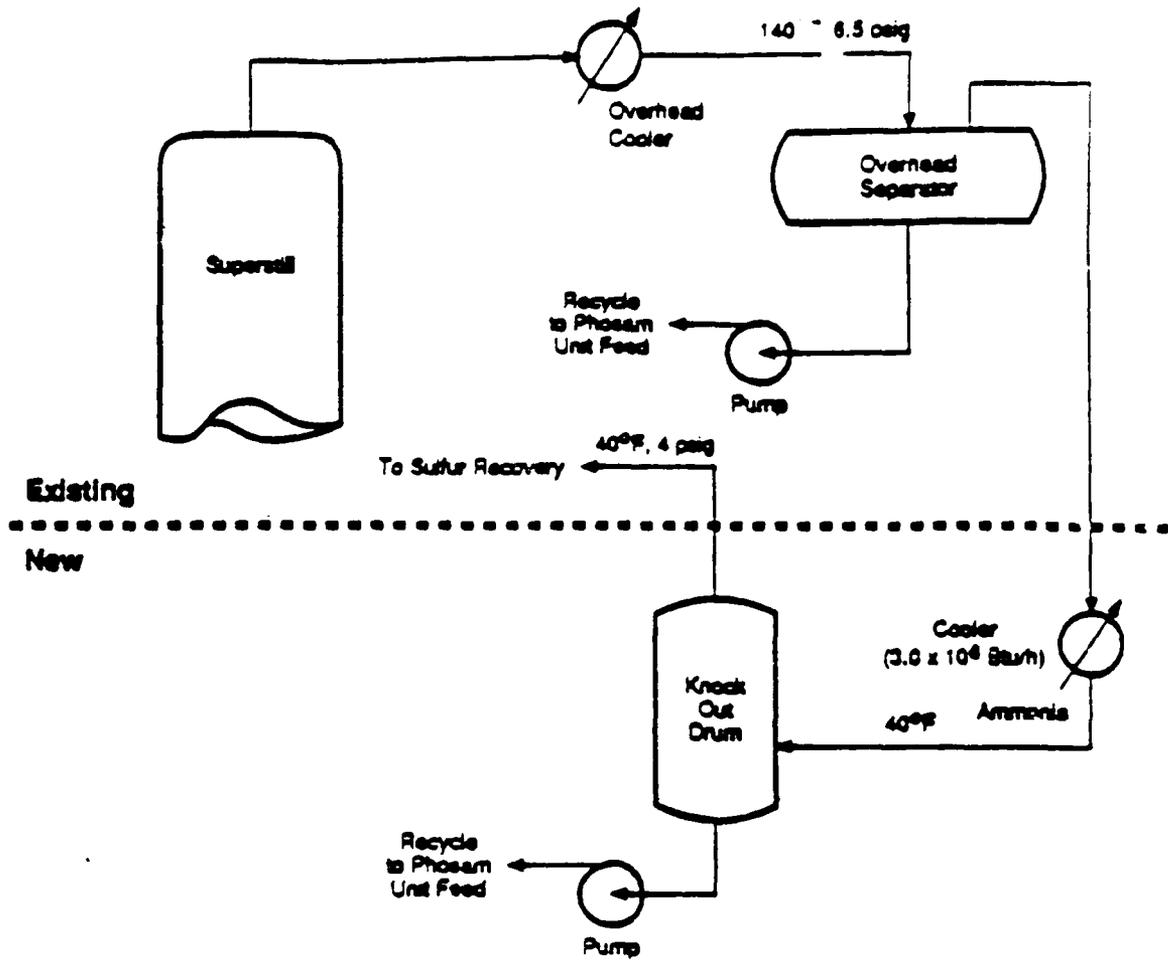
FIGURE 20.2 Phosam Chilling — Case 2B

160 MMSCFD SNG rate, additional exchanger capacity may be required; however, this can only be verified by performance tests with the ammonia stripper reboiler in service.

The condensed liquid would then be separated from the gas via the existing overhead separator. The gas from the separator is cooled further to 40°F by the ammonia chiller. The estimated duty is approximately 1.75 MMBtu/h. The condensed liquid from the chiller is knocked out via a knockout drum before the noncondensable gases are sent on to the Sulfolin Unit for treating.

20.5 FEASIBILITY — PHOSAM COOLING CASES 2A, 2B

20.5.1 Environmental Benefits

Chilling the Phosam stream to 40°F will permit processing in the Sulfolin Unit. Current removal efficiency in the Sulfolin Unit is predicted to be 78 percent at the 160 MMSCFD plant capacity. This indicates that SO_2 emissions could be reduced by 747 lb/h. Organic sulfur species will not be reduced by this approach; however, it should be noted that the Phosam off-gas contributes only 17 lb/h of SO_2 from organic sulfur.

20.5.2 Investment Cost

The first arrangement, providing additional overhead condensing capacity, has been estimated to cost $2.5 MM. The alternative arrangement should be less and will improve the economics of this option to permit processing Phosam overhead gas in the Sulfolin Unit.

20.5.3 Operating Cost

Again, the first arrangement has higher utilities requirements to operate the fans for the air coolers. The other arrangement could reduce the operating costs by 30 percent.

20.5.4 Pros and Cons

This option offers SO_2 reduction at similar cost per pound of SO_2 as fuel gas chilling. This is a simple method to make the Phosam off-gas acceptable for feed to the Sulfolin Unit. It would be low cost and has inherent benefits to other plant unit operations, e.g., proper cooling of this overhead stream permits recovery of IPE which, at times, has been present in excessive quantities. In case of adding a reboiler, it will also improve the plant water balance and operation of the Phosam Unit. There is some uncertainty regarding Sulfolin decanter operation since the pilot plant cannot simulate continuous melter operation.

20.5.5 Resulting Emissions If Fuel Gas and Phosam Gas Are Treated in Sulfolin

The cost for removal of hydrogen sulfide from fuel gas and Phosam off gas by chilling both streams to 40°F is very attractive compared to other options.

20.6 ECONOMICS

20.6.1 Phosam and Fuel Gas Cooling — Case 1

The economics appear are presented in Table 20.1.

20.6.2 Phosam Cooling — Cases 2A, 2B

The economics of this option area as attractive as pretreating the fuel gas. As shown in Table 20.2, the net operating costs for Case 2A are only $0.13 million per year. The operating costs for Case 2B are expected to be about 30% lower.

Pilot plant operation with the blended feedstreams of Phosam, fuel gas, and Rectisol waste gas has shown acceptable performance and solution behavior. Therefore, risks are considered low. As indicated in the description, the two arrangements regarding the Phosam stripper reboiler versus superstill overhead are being evaluated for reasons other than SO_2 reductions. Subject to these findings, the most attractive arrangement will be used for implementation.

TABLE 20.1 Economics of Phosam Off
Gas Cooling — Case 1

Item	Cost
Lock gas refrigeration	
Total plant cost[a]	$1,000,000[b]
Operating costs (annual)	
Annualized capital cost	$ 150,000
Fixed operating cost	$ 30,000
Maintenance materials	$ 50,000
Variable operating cost	$ 197,000
Total annual cost	$ 427,000
Phosam gas refrigeration	
Total plant cost[a]	$ 789,000[c]
Operating costs	
Annualized capital cost	$ 118,000
Fixed operating cost	$ 24,000
Maintenance materials	$ 40,000
Variable operating cost	$ 718,000[d]
Total annual cost	$ 900,000

[a] $ first quarter 1987.

[b] Using a cold methanol slip stream from
Rectisol for chilling would reduce this
cost to $745,000.

[c] Using a cold methanol slip stream from
Rectisol for chilling would reduce this
cost to $591,000. Using an air cooler
in place of a water cooler would
increase this cost to $1,352,000.

[d] Includes a cost for liquid waste
disposal.

TABLE 20.2 Economics of Phosam Off
Gas Chilling: Case 2A

Item	Cost
Capital Cost	$2,500,000
Operating Costs	
180 HP Fan	46,000
915 HP Refrigeration	50,000
30 GPM Cooling Water	2,000
Maintenance	100,000
Subtotal	198,000
Sulfur Credit	67,000
Net Operating Cost	131,000

23 LOCK GAS REFRIGERATION

23.1 PROCESS DESCRIPTION

The option of refrigerating the lock gas (or fuel gas) from the Lurgi gasifiers was previously examined. Certain feedstreams (Phosam off gas, the naphtha stripper overhead, and lock gas streams) created problems with the operation of the original Stretford unit. The difficulty was attributed to organic trace compounds. Cooling of the feedstreams to ambient or refrigerated temperatures is expected to condense and wash out these compounds. This pretreatment step may improve the composition of the problem streams sufficiently to achieve acceptable processing through the Sulfolin unit.

23.2 GP APPLICATION

The flow diagram in Fig. 23.1 shows the arrangement for chilling the fuel gas. Condensate from the knockout pot will be returned to the gas/liquor separation area.

Pretreated fuel gas can now be processed in the existing Sulfolin unit. In order to achieve a reasonable removal efficiency, solution pumps from both existing Sulfolin trains will be operated. Both venturis will be operated in parallel. Solution from the fuel gas absorber will be returned to both trains using restriction orifices for near-equal solution flow distribution.

23.3 FEASIBILITY

23.3.1 Advantages and Limitations

Chilling fuel gas has proven to improve pilot plant Sulfolin operation, resulting in solution behavior similar to operation with Rectisol waste gas. The major unknown at this time is the impact that residual trace compounds may have on the melter operation.

23.3.2 Impact on Sulfur Dioxide Emissions

Based on evaluation of 1985 plant data, it was determined that the fuel gas venturis, if operated in parallel, could achieve a 70 percent removal efficiency (see Table 23.1). This represents a 493 lb/h reduction in SO_2 emissions.

No reduction in organic sulfur species is assumed while fuel gas is processed in the Sulfolin unit. Analysis of current Sulfolin feed and effluent streams did not indicate a noticeable reduction in organic sulfur species.
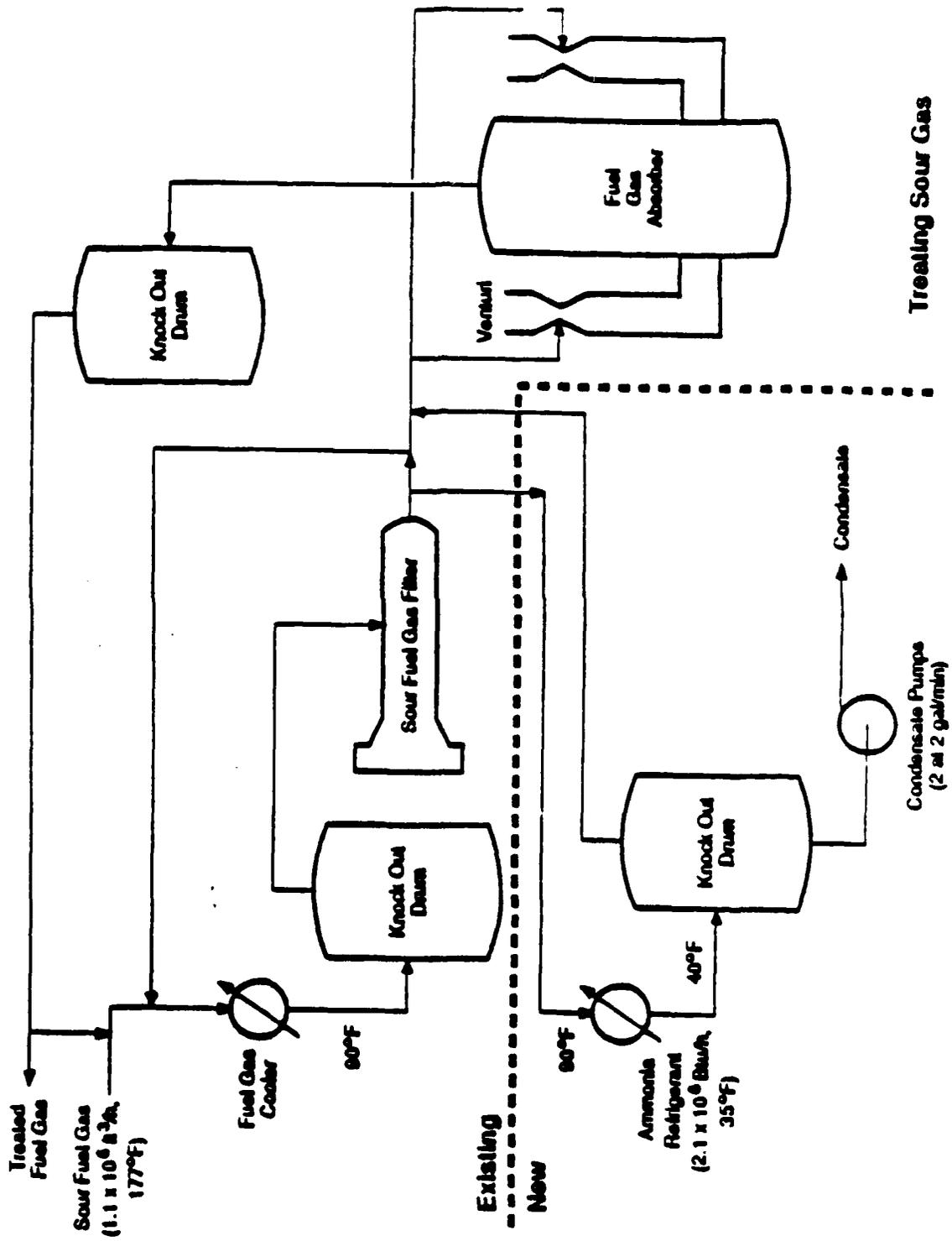
FIGURE 23.1 Flow Diagram for Chilling Fuel Gas

TABLE 23.1 SO$_2$ Emissions with Lock Gas
Refrigeration

Parameter	Untreated Fuel Gas	Lock-Gas Refrigeration
Flow (MMSCFH)	1.1	1.1
Removal Efficiency (%)	-	70[a]
H$_2$S in (ppm)	3,900	3,900
H$_2$S out (ppm)	3,900	1,170
Org. S in, out (ppm)	270	270
SO$_2$ Emissions (lb/h)	774	267

[a]Assumed efficiency of H$_2$S removal in the Sulfolin
fuel gas unit using a packed absorber.

23.4 ECONOMICS

23.4.1 Capital Costs

The cost for such a refrigeration system with associated equipment has been estimated at $1.0 MM. Due to the insignificant changes to the current Sulfolin fuel gas system, no specific allowance has been included. A review of cost required to install larger pumps and associated piping in each train indicates a potential cost of $0.3 MM. It is recommended to utilize the current arrangement and replace pumps, etc., only if such equipment must be replaced due to wear and tear.

23.4.2 Operating Cost

Utility costs are incurred for the refrigeration unit and Sulfolin circulation pumps. Table 23.2 provides these utilities ass options. In addition, a 4 percent allowance is added as a percentage from the required investment cost to cover fixed costs, maintenance materials, and labor.

23.4.3 Summary for Economics

The economic analysis indicates that SO_2 can be reduced at reasonable cost. It costs about 3.4 cents/lb of SO_2 and adds 0.3 cents per MMBtu of SNG produced. The risk regarding melter operation is considered to be reasonable compared to the other alternative of compressing fuel gas to system pressure for processing through the gas cooling and Rectisol units. Parallel venturi operation should achieve sufficient H_2S removal for fuel gas.

23.5 RECOMMENDATIONS

It is recommended that GP pretreat the fuel gas by chilling it to 40°F at the 20 psig operating pressure before sending the fuel gas to the Sulfolin unit.

TABLE 23.2 Economics for Lock Gas Refrigeration

Capital cost	$1,000,000
Annual operating costs	
Refrigeration	49,000
Cooling water	11,000
Solution pumps	77,000
Maintenance	40,000
Subtotal	177,000
- Sulfur credit	(44,000)
Net cost	$133,000