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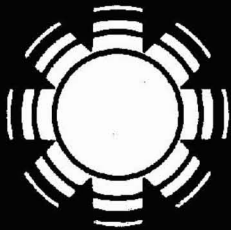
April 1987

Economic Feasibility Study of a Wood Gasification-Based Methanol Plant

A Subcontract Report

**Stone & Webster Engineering Corp.
Boston, MA**

Prepared under Subcontract No. ZX-3-03095-1



SERI

Solar Energy Research Institute

A Division of Midwest Research Institute

1617 Cole Boulevard
Golden, Colorado 80401

Operated for the

U.S. Department of Energy

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SERI Technical Monitor:
J. D. Wright

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PREFACE

At the request of the Solar Energy Research Institute (SERI), Stone & Webster Engineering Corporation (SWEC) has prepared an Economic Feasibility Study for a Wood Gasification Based Methanol Plant. The objectives of this Feasibility Study are to evaluate the current commercial potential of a small scale wood-fed methanol plant using the SERI oxygen-blown, pressurized down-draft gasifier technology and to identify areas in which further research and development are required.

The technical and economic details of this report are based on several assumptions, which are explained in detail in the body of the text. Two of the assumptions:

- ° the validity of the gasifier computer model, and the
- ° Nth plant assumption

are important to the overall concept.

The gasifier gas composition and material balance are based on a computer model of the SERI gasifier, since acceptable test data was not available at the time of this study. The model did simulate experimental data obtained at less severe operating conditions than those required for the study. Although the engineering design and cost of the facility are based on this simulation, only large discrepancies in gasifier output would result in significant changes in capital costs.

The estimated capital cost is based on the Nth plant constructed. Considering the small size and commercial nature of most of the equipment, it is reasonable to assume that N is a number between 5 and 10. The additional cost of constructing the first plant (about \$2.0 million) is discussed in the text.



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SECTION 1
INTRODUCTION AND SUMMARY

1.1 INTRODUCTION

At the request of the Solar Energy Research Institute (SERI), Stone & Webster Engineering Corporation (SWEC) has prepared an Economic Feasibility Study for a Gasification Based Methanol Plant. The objectives of this Feasibility Study are to evaluate the current commercial potential of a small scale wood-fed methanol plant using the SERI oxygen-blown, pressurized down-draft gasifier technology and to identify areas in which further research and development are required. This report presents the results of this study, including site selection, gasifier scale-up, an integrated conceptual design, a capital cost estimate, an investment analysis, an analysis of alternative designs, and recommendations for future research and development.

The full scale Biomass-Based Methanol Plant (BMP) will be capable of producing 7.68 MM gal/yr of fuel grade methanol (see Table 1-1) from green wood chips. The area adjacent to Spokane, Washington, has been selected for the site of this study. The BMP uses Lurgi technology to convert the prepared synthesis gas to methanol. It is self-sufficient in steam production, and imports only one-third of its electrical needs.

The technical information required for this study was obtained from SERI, published literature, conversations with various process and equipment suppliers, and SWEC in-house data. Dr. Tom Reed of SERI supplied the data concerning the gasifier.

1.2 SUMMARY

Methanol is currently produced commercially worldwide via the conversion of synthesis gas derived from natural gas, refinery off gas, petroleum, or



coal. The potential conversion of wood (biomass) to methanol can be performed in a manner similar to that of coal conversion. A wood to methanol conversion facility would contain unit operations (equipment) that are commercially available except for the wood gasifier. The scale of the plant (less than 10 MM gal/yr of methanol) is limited so that the potential number of sites is not unduly restricted due to the availability of biomass resources or the ability of the local market to utilize the methanol product. Therefore, a sufficient number of plants of standardized design can be built to take advantage of repetitive manufacturing and engineering, and shop fabrication techniques in order to reduce unit capital costs.

Since the gasification technology has been specified, the study focused on site selection, gasifier scale-up, process integration, research and development needs, and economic analysis. These areas may be consolidated into the following:

- Site Selection
- Technical Feasibility
- Profitability

The profitability is based on a discounted cash flow rate of return (DCFROR) analysis. The analysis includes the cost of capital, operating and maintenance costs, by-product sales, and tax credits. At a DCFROR of 15 percent, the sensitivity analysis shows a range of methanol selling prices from 81¢/gal to \$1.31/gal.

Several major assumptions made by SWEC have established the direction of the study and the base case design. The assumptions are as follows:

- The price of conventional methanol, delivered to the selected site, should be higher than BMP produced methanol.
- The product methanol will be used as an octane enhancer (with a cosolvent) or as a "neat" fuel (90 percent methanol - 10 percent gasoline).



- All unit operations except the gasifier should be commercial or nearly commercial.
- The plant will purchase raw materials and services over the fence, as opposed to self-generation, when practical; i.e., oxygen, wastewater treatment, etc.
- The maximum utilization of internally available energy through reasonable integration is a design goal.
- Maximum utilization of shop fabrication and skid construction has been specified.

The design basis for the BMP is shown in Table 1-2.

1.2.1 Site Selection

A primary assumption for this study is that the methanol produced will be used as a motor fuel product. Initially, methanol will be used as an octane enhancer and potentially as a "neat" motor fuel (90 percent methanol, 10 percent gasoline). This motor fuel methanol assumption results in a restriction of potential sites to population concentrations of more than 225,000. Other uses could potentially increase the number of sites.

The major factor in selecting a site is to establish the delivered price of conventional methanol to the region. Other factors include the price and availability of the raw materials (wood, oxygen, power, etc), the potential for by-product sales, and the ability of the local region to utilize the product.

The cost of transporting methanol from the major domestic production sites is detailed in Section 3. SWEC has established three tiers of transportation costs for site selection.



Tier I	-	Transportation costs	less than 15¢/gal
Tier II	-	Transportation Costs	15¢-25¢/gal
Tier III	-	Transportation Costs	greater than 25¢/gal

Tier I encompasses almost the entire U.S.A. east of the Mississippi, the major western cities on the inland waterways (Kansas City, Omaha, Tulsa), and the Pacific Coast (San Diego, Los Angeles, San Francisco, Portland, Seattle). Tier II encompasses areas such as Denver, inland California (Fresno), part of inland Washington State and Oregon, the upper midwest (Minnesota, Wisconsin, eastern Dakotas), Hawaii, and coastal Alaska. Tier III includes the major western inland cities: Salt Lake City, Utah; Albuquerque, New Mexico; Spokane, Washington; Boise, Idaho; Las Vegas, Nevada; and Phoenix, Arizona.

Due to the cost of transporting conventional methanol into the area, the regions in Tier III are the most likely to be able to support a methanol plant if adequate raw materials are available. For this study, a Tier III site, Spokane, Washington, has been selected.

Additional sites outside of Tier III may be economically viable, due to special circumstances such as state incentives and the potential utilization of low cost waste feedstocks.

1.2.2 Technical Feasibility

The biomass-based methanol plant utilizes commercially available equipment to the greatest extent possible. The SERI gasifier, which has operated at a nominal throughput of 1 ton per day, is the only developmental unit in the plant design. The process integration and engineering design for the base case concentrated on scale-up of the SERI gasifier, maximizing methanol production, minimizing importation of auxiliary fuel, and the use of a skid-mounted design.

The alternative process designs investigated identified one additional area of improved plant profitability, that of reforming the methanol synthesis



purge gas for additional methanol production. The alternative designs are discussed in Section 7. Increased facility size (within the concept of a small scale plant) up to the ability of the local market to use the product and supply the resources required also improves the overall economics.

1.2.2.1 Scale-up

The analysis of the potential problems in scaling up the SERI gasifier concentrated on oxygen distribution, uniform wood distribution, and gas flow. Nonuniform gas flow would be the result of bridging of the bed (sticky wood or ash fusion) or plugging of the bed (fines generation). The maximum gasifier size was limited to what would easily fit a rail transportable skid with minimization of field installation. The present design of 8.5 ft ID and 10 ft OD, with a throughput of 1.78 MM Btu/hr-ft², was chosen as a reasonable scale-up when considering other solid fuel gasifiers, the SERI pilot plant design, and transportation limitations (see Section 4.3).

1.2.2.2 Engineering Design

The BMP design is discussed in detail in Section 5. Several energy and process integration steps have been incorporated to increase product output and reduce energy demand. The more important process design improvements are:

- The use of an indirectly heated wood drier. This allows the drier exhaust steam to be reused, via vapor compression, as the heat source for the acid gas stripper reboiler.
- The use of direct water injection to supply steam for the shift reaction. Water is injected into the raw synthesis gas for cooling and to supply the shift steam.



- The use of a Pressure Swing Adsorption (PSA) unit for increased methanol production. Approximately 75 percent of the hydrogen from the methanol purge gas is recovered and recycled to the methanol synthesis unit increasing the yield approximately 7 percent (from 72 to 77 tpd). The remaining purge gas is fed to a diesel engine generator set to produce power and steam.
- The use of a small fluidized bed combustor to burn combustible wastes, i.e., gasifier char and wood fines. This eliminates a majority of the solid waste disposal problem and provides fuel for the generation of the balance of process steam requirements and startup steam.

1.2.2.3 Skid-Mounted Design

The BMP consists of eleven sections (areas) in which maximum utilization of shop fabrication and skid mounting of equipment has been considered in order to reduce field installation. Only Section 100 - "Wood Handling and Preparation" has not been designed for skid mounting. This is due to the size and elevations of the conveyors, screens, and other equipment that make field installation more practical.

In the remaining ten sections, most of the equipment is skid-mounted except for the major towers, which are freestanding. There are a total of 46 individual skids of which 13 are vendor-supplied packages and seven are air-cooled heat exchangers placed on top of several other skids. While the total number of skids is large, many skids are of moderate size. The maximum size of the skids was based on standard rail shipment. This allows a basic envelope of 12 ft wide by 12 ft high by 80 ft long.

1.2.2.4 Research and Development Needs

The technical analysis (scale-up and process design) has identified several areas where additional data are required to firm up the conclusions drawn in this study. These requirements are detailed in Section 8. A portion of the



data relating to complete quantification of all product streams may be obtained in the current SERI pilot plant (gasifier ID = 6 in.), but the critical information relating to scale-up, mechanical integrity, and reliability requires the operation of a larger pilot facility, with a gasifier of approximately 2 ft internal diameter, designed to handle approximately two tons per hours of green wood.

1.2.3 Profitability

The investment analysis of the BMP is based on the Nth plant in a series of BMPs. The analysis is carried out using a DCFROR type analysis on a constant dollar basis (1984 dollars), 100 percent equity financing and current prices for conventional methanol (June 1984). The construction time for the Nth plant, including engineering, design, installation and a startup, is estimated to take 18 months. The total facilities investment is \$22,153,000. The economic basis is shown in Table 1-2.

In order for private investors to be interested in this venture, it is assumed that the minimum acceptable after tax DCFROR is 15 percent. The maximum selling price of the fuel methanol product as shown below is \$1.16/gal in the Spokane area.

<u>Component</u>	<u>Value (¢/gal)</u>
U.S. Gulf Coast Price	40
Transportation Costs to Spokane	26
Alcohol Fuel Tax Credit	<u>50</u>
Equivalent Methanol Selling Price	116

This price is used to establish the range of DCFRORs for the base case design when considering combinations of sensitivity parameters. The corresponding methanol selling price to meet the minimum DCFROR criteria (15 percent return) is also shown below:



<u>BMP Scenario</u>	<u>DCFRR</u> (at 116¢/gal)	<u>Methanol Selling Price (¢/gal)</u> at 15 percent DCFRR
Pessimistic	11.1	131
Base Case Design	21.5	100
Optimistic	30.0	81

The extreme scenarios (see Section 6) are based on combinations of sensitivity parameters which are not likely to occur at the same time. The base design indicates that a BMP investment would attract private financing.



TABLE 1-1

METHANOL PRODUCT SPECIFICATION

<u>Characteristic</u>	<u>Method</u>	<u>Specification^a</u>
Methanol content		99.40 ^b
Water content	ASTM-D-1364	0.025 ^c
Residue of Evap (mg/100 mL)	ASTM-D-1353	1.0
Acid (as acetic acid)	ASTM-D-1613	0.002
Alkalinity (as ammonia)		0.00030
Chloride (max)	Spec. ion electrode	0.0001
Lead (g/gal)	ASTM-D-3116	0.01
Phosphorous (g/gal)	ASTM-D-3231	0.005
Sulfur	ASTM-D-2622	0.025
Iron, volatile (mg/L)	Photometry	0.02
Permanganate No. (min)	ASTM-D-1363	60 min
Acetone No. (max)	Chromatography	0.002
9-11 psi RVP Unleaded gasolines (seasonally) added for volatility (volume percent)	ASTM-D-439-89	9.0-11.0

^a weight percent except as noted

^b California Energy Commission specification is 99.85.

^c California Energy Commission specification is 0.05.



Table 1-2

BIOMASS BASED METHANOL PLANT - BASIS OF DESIGN

<u>ITEM</u>	<u>BASIS</u>
Raw Wood Feed	Whole Tree Chips - 50 percent Moisture Content - 311.4 Tons per Day
Onsite Storage	Seven (7) Days
Processed Wood Feed	Pin Chips (3/8 In. Dia. x 1½ In. L) 16 Percent Moisture Content - 173.2 Tons/day
Gasifier Pressure	150 psig
Gasifier Size	8.5 Ft Internal Diameter
Gasifier Heat Loss	2 Percent of Input HHV
Wood Carbon Conversion to Gas	98 Percent
Gasifier Throughput	1.78 MM Btu/Ft ² - Hr
Acid Gas Removal	Benfield - Hot Potassium Carbonate Followed by Zinc Oxide Guard Catalyst
Methanol Synthesis	Lurgi Low Pressure Process
Methanol Yield	77.1 Tons/day (7.68 MM gal/yr)
Energy Integration	1. All Steam Generated Internally; 2. Electricity - Part Internally Generated, Part Purchased
Solid, Liquid, Gaseous Wastes	System Designed to Meet Local/Federal Regulations



Table 1-3

BASIS FOR INVESTMENT ANALYSIS^(a)

<u>COMPONENT</u>	<u>AMOUNT</u>
Plant Life	20 years
Operating Factor	7,884 hr/yr (90 percent)
Equity/Debt	75 percent/25 percent
Minimum Return on Equity Required	15 percent DCFROR (after tax)
Interest Rate during Construction	8 percent real
Methanol Production Rate	7.68 MM gal/yr
Depreciation	ACRS (13.5%,19.8%,18.9%,18.9%,18.9%)
Tax Rate	50 percent
Investment Tax Credit	10 percent
Renewable Energy Tax Credit	10 percent
Alcohol Fuel Tax Credit	50¢/gal
Construction Time	18 months
Wood Cost	\$30/BDT
Oxygen Cost	\$52.82/ton ^(b)

(a) The investment analysis is based on a discounted cash flow rate of return (DCFROR) in constant 1984 dollars. The DCFROR analysis assumes that the ITC and RETC tax credits are used to reduce the first year capital investment. The five (5) year accelerated cost recovery system method is applied to the total depreciable plant and the DCFROR calculation is applied on a beginning of the year basis. The cost of capital (Rate of Return) and interest rates is net of inflation.

(b) The cost of oxygen is based on a 60 TPD dedicated facility operated by an independent vendor, with the methanol plant having a take-or-pay contract.



SECTION 2

FINDINGS AND RECOMMENDATIONS

2.1 FINDINGS

The economic success of a small scale, biomass-based methanol plant (BMP) is based on taking advantage of the high transportation costs for conventionally produced methanol and the Federal alcohol fuels tax credit to enable the product methanol to be sold competitively in the local market for automotive fuel use. The alcohol fuel tax credit is available since the methanol is produced from biomass. The potential selling price for fuel methanol in Spokane is \$1.16 per gallon (as defined in Section 1). Table 2-1 represents a summary of the important findings of the study for the Spokane, Washington site. The BMP is self-sufficient in fuel requirements and cogenerates approximately two-thirds of its electrical needs.

The capital costs for the BMP are detailed in Table 2-2 and the annual operating costs are shown in Table 2-3. The primary operating cost components are shown to be wood, oxygen and labor. The investment analysis is based on the DCFROR method, as outlined in Section 6. The minimum required DCFROR (after tax) necessary to attract private investors has been assumed to be 15 percent. Figure 2-1 shows the variation of DCFROR with the potential selling price of methanol from a BMP. In the Spokane area, this minimum return can be obtained with a selling price (100¢/gal) that is less (by 16¢/gal) than the potential maximum methanol selling price (of 116¢/gal). Although in this analysis the alcohol fuels tax credit is required to attract private investment, any increase in the domestic methanol price directly offsets the need to depend on the tax credit on a one for one basis.

The differential between the potential selling price for methanol and the required methanol selling price is significant. This 16¢/gal reduction enables regions with lower transportation cost to become potentially viable



sites for a BMP. Most sites in Tier II (transportation costs 15-25¢/gal) become acceptable as long as raw material prices and labor costs are not substantially different from the Spokane (base case) location. This increase in the number of potential sites substantiates the assumption of a sufficient number of BMPs to justify the economics associated with the skid design and Nth plant criteria.

Two additional areas of improvement of the plant economics which can expand the geographic viability of BMPs or reduce the dependence on the alcohol fuels tax credit (or increase the attractiveness of the investment) are:

- Reforming the methanol synthesis purge gas and recycling it for increased methanol production.
- Financing a portion of the plant cost via the assumption of debt.

The net effect of the added process operation of reforming the purge gas (see Section 7) is to reduce the BMPs production costs by 6¢/gal. This process improvement should be added to the base case in future evaluations of the BMP concept.

Replacement of equity with debt reduces the methanol selling price required for a 15 percent DCFROR. The precise level of the replacement must be determined by the financial community. Assuming a debt level of 25 percent at a real interest rate of 8 percent results in a reduction of the required methanol selling price of 14¢/gal. Table 2-4 shows the summation of these improvements, which result in a required methanol selling price of 80¢/gal.

2.2 RECOMMENDATIONS

Since the technology to convert synthesis gas to methanol is well established, the main technical questions raised in this study relate to the performance of the SERI gasifier, which converts wood to synthesis gas. Additional pilot plant testing, both at the current scale and at a larger scale, is required to confirm gasifier product yields, unit capacity, and operational integrity.



This additional data confirmation may result in changes to the design basis which cause a change in the base case DCFROR and make the facility more or less dependent on the alcohol fuel tax credit.

The analysis of the BMP has indicated (with the assumptions made) that the concept of a small scale biomass-based methanol plant can be an attractive investment, once the few remaining technical issues are resolved.

Therefore, it is recommended that further testing be conducted on the current SERI pilot plant to answer the questions enumerated in this study (Section 8). With positive results from this small scale testing, a larger unit should be built to answer the questions concerning scale-up and operational integrity.



Table 2-1
 METHANOL FROM BIOMASS
 SPOKANE, WASHINGTON BASE CASE

<u>COMPONENT</u>	<u>PARAMETER</u>
Wood Consumed	311.4 tons per day (at 50% moisture)
Oxygen Consumed	58.2 tons per day
Methanol Produced	77.1 tons per day (7.68 MM gal/yr)
Capital Cost	\$22,153,000
Operating Costs	58.94¢/gal
Required Methanol Selling Price ^(a)	1.00/gal

(a) at 15% DCFROR (after tax), 100% equity, constant 1984 dollars.



Table 2-2

BIOMASS BASED METHANOL PLANT

CAPITAL COST - BASE CASE

<u>COMPONENT</u>	<u>COST</u>
Material and Equipment	\$11,354,900
Field Labor (including Contractor's Fee)	4,610,000
Freight and Tariffs	<u>189,100</u>
Total Base Cost	16,154,000
Land Cost	30,000
Engineering and Construction Management	1,870,000
Contingency and Process Development Allowance	<u>1,700,000</u>
Total Installed Cost	19,754,000
Initial Catalysts and Chemicals	207,000
Startup Expenses (2 months)	751,000
Interest During Construction	1,156,000
Working Capital	<u>285,000</u>
Total Facilities Investment	\$22,153,000



Table 2-3
BMP ANNUAL OPERATING COSTS SUMMARY^(a)

<u>COMPONENT</u>	<u>COST (\$)</u>	<u>METHANOL COMPONENT COST ¢/GAL</u>
Wood (\$30/ton)	1,534,424	19.98
Oxygen (\$52.82/ton)	1,009,443	13.14
Electricity (4.07¢/kWh)	176,894	2.30
Catalysts and Chemicals	236,567	3.08
Miscellaneous	99,798	1.30
Labor	720,000	9.38
Maintenance	507,000	6.60
Insurance and Taxes	<u>242,310</u>	<u>3.16</u>
	4,526,436	58.94

(a) Methanol production rate is 7.68×10^6 gal/yr

Table 2-4

IMPROVEMENTS OVER THE BASE CASE

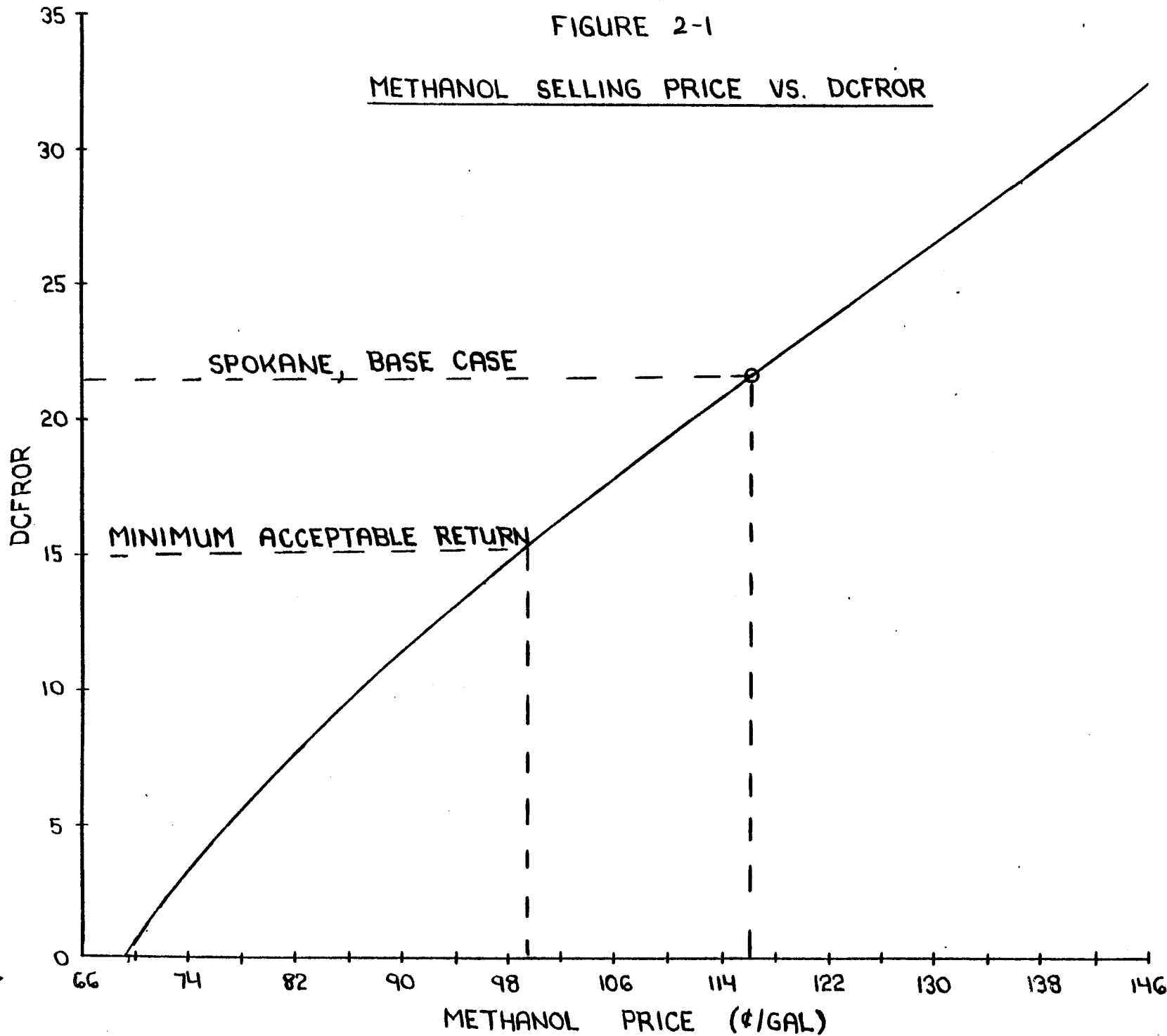
<u>COMPONENT</u>	<u>REQUIRED METHANOL SELLING PRICE (¢/GAL)</u>
Spokane Site @ 15% DCFROR	100
Reforming Purge Gas	-6
Debt Assumption (a)	<u>-14</u>
New Required Selling Price @ 15% DCFROR and 25% Debt	80

(a) 25% debt at 8% real interest rate



FIGURE 2-1

METHANOL SELLING PRICE VS. DCFROR



2-8

STONE & WEBSTER



SECTION 3

SITE SELECTION

The selection of the site for the biomass (feedstock) methanol plant (BMP) is based on (1) a probable market for the methanol product and (2) the production of methanol at a price that can compete with the price of conventional methanol transported to the site. Specific state tax incentives are not considered in the site selection. The number of acceptable sites would be increased with sufficient incentives. This variable is addressed in the economic analysis (Section 6.6).

3.1 MARKET DEFINITION AND SITE LOCATION FACTORS

In order to assure a market for the methanol, metropolitan areas with a population in excess of 225,000 people have been identified. Only these areas have been given further consideration as potential BMP sites. Populations of this size would be able to utilize all of the methanol produced, assuming that methanol has fully penetrated the fuel market as an octane enhancer (10 percent blend). Although use as an octane enhancer is the most immediate market for fuel methanol, the ultimate goal is widespread use of neat methanol as a motor fuel. It should be noted that as the acceptance of neat methanol as a fuel grows, the population requirement for a viable site decreases, increasing the number of potential sites for small-scale methanol from biomass facilities. This is due to the substantial difference in methanol concentration between the fuels, i.e., 10 percent methanol assumed when used as an octane enhancer as opposed to near 90 percent purity methanol in "neat" methanol. Additional locations for captive industrial uses are possible but this consideration is outside the scope of the size definition. Some current methanol concentrations in motor fuels are:



<u>Blend</u>	<u>Vol % MeOH</u>
Petrocoal	12, maximum
G&G Alcohol Fuels, Inc.	10
Oxinol 50 (Arco)	4.75

Given the areas that could support a methanol production facility, it is necessary to determine the likelihood that methanol produced in the BMP could compete in a local methanol market. If methanol produced from natural gas could be obtained at a lower price than from an BMP, the BMP would not be economic. Location factors that influence the economics of the BMP include:

- transportation costs from existing methanol plants to potential sites (markets)
- availability and/or cost of oxygen at potential sites, and
- wood availability and costs at potential sites.

3.2 TRANSPORTATION COSTS FOR CONVENTIONAL METHANOL

Because BMP methanol is competing with conventionally produced methanol, it is necessary to locate the BMP at some point where shipping charges substantially increase the market price of conventional methanol from natural gas. The locations of existing conventional plants were identified (Gulf Coast and Delaware City) and shipping costs to potential market areas from the nearest methanol plant were determined. This was accomplished by obtaining actual tanker, barge, or rail transportation costs between key points. Shipping charges for intermediate locations were then prorated from these values on a cost per unit-mile basis for the appropriate method of transport.



Table 3-1 summarizes the actual shipping costs and specifies the quantities upon which the rate, used to determine these costs, is based as well as the means of shipping. These data show that transportation charges are not solely a function of distance. For example, Gulf Coast methanol can be shipped to the West Coast by rail or by water through the Panama Canal. The large capacity of a parcel tanker allows a shipping cost of 13.3¢/gal to Los Angeles, whereas by rail, the transportation charge is 16.8¢/gal, even though the rail distance is less. Rail rates can be inconsistent because they depend on traffic density and competition. For example, consider the rail charges to Denver and Albuquerque from Texas City. Although Albuquerque is closer to Texas City than Denver is, the cost to Denver is less than half that to Albuquerque (15.2¢/gal vs 33.1¢/gal).

These actual rates along with the interpolated charges are summarized in Figure 3-1, a cost contour map of the United States showing estimated shipping costs in cents per gallon. Based on a Gulf Coast (fob) price of 45¢/gal for methanol, and assuming a cost of 60-70¢/gal for BMP production, three tiers of potential sites were established:

- Tier I <15¢/gal transportation cost
- Tier II 15-25¢/gal transportation cost
- Tier III >25¢/gal transportation cost

Tier I is subdivided in Figure 3-1 into three cost ranges for better definition. If transportation costs fall into Tier I, delivered conventional methanol costs would be 45-60¢/gal. In these areas, BMP methanol is probably not economic because its cost is above that of natural gas based methanol. For shipping costs of 15-25¢/gal(Tier II), corresponding methanol costs would be 60-70¢/gal. Since this is the same range that was expected for BMP methanol, the current tax incentives may make these areas economic. At transportation costs of greater than 25¢/gal, conventional methanol prices exceed the estimated net cost of BMP methanol, and BMP methanol is economically attractive.



As can be seen in Figure 3-1, low transportation costs probably disqualify all points east of the Mississippi River as well as major western cities on the inland waterways as potential sites. Due to the extensive barge and rail systems, all eastern and midwestern population centers fall into the 0-15¢/gal transportation cost range.

On the West Coast, BMP facilities located in coastal cities, i.e., San Diego, Los Angeles, San Francisco, Portland, and Seattle, would probably be uneconomic since they have transportation costs of less than 15¢/gal from domestic methanol sources. Foreign methanol is assumed to be sold at the domestic landed price. Rail costs obtained for the adjacent population centers indicate that most would fall into the 15-20¢/gal market. That is, BMPs located in Eugene, Sacramento, Fresno, Bakersfield, or San Bernardino, with tax incentives, may be economic. Spokane, however, is sufficiently distant from Seattle to have a total methanol shipping cost of 26¢/gal. Therefore, Spokane is potentially an economically viable site.

Because the interior western U.S. is so sparsely populated, point to point transportation costs were determined for 200,000+ population centers only. Cost bands were not determined for this portion of the country because of (1) insufficient local populations and (2) inability to prorate rail costs because of the large variety of shipping costs that result from the different frequencies of use of the rail lines. Denver falls into the 15-25¢/gal range, indicating once again possible economic feasibility with proper incentives. Both Salt Lake City and Albuquerque have shipping costs greater than 25¢/gal, and therefore may be attractive sites. The other qualifying population centers, Phoenix and Tucson, were eliminated because they lack a nearby adequate wood feedstock source.

In summary, the entire eastern half of the United States, as well as the major western coastal cities, are probably uneconomical locations due to the low cost of transporting conventional methanol. Denver and population centers adjacent to the western coastal cities, such as Eugene, Sacramento, Fresno, Bakersfield, and San Bernardino, may be viable locations for BMPs



with tax incentives. Hawaii and coastal Alaska also fall into this category. Spokane, Salt Lake City, and Albuquerque, at first estimate, are the most potentially attractive sites due to sufficiently high shipping costs and the availability of wood resources.

3.3 OXYGEN AVAILABILITY AND COST

Because the SERI gasifier that is the basis for this evaluation must be oxygen blown to produce a suitable gas for methanol synthesis, the availability and cost of oxygen are important economic siting factors. Oxygen can be supplied "over-the-fence" from either an existing oxygen plant or a new dedicated facility (about 60 tpd are required) or be shipped to a site from an existing plant. Unless shipment is by an existing pipeline, the last option is more expensive than building a new, small dedicated facility. The extra energy cost required to liquefy oxygen for truck or rail transport plus the transportation costs are greater than the savings due to the economies and efficiencies of scale obtainable in a larger (1000 tpd plus) plant. Since all existing oxygen pipelines are in areas where transportation costs for conventional methanol are low (Indiana, Texas, and Louisiana), only gaseous oxygen purchase from an existing plant or construction of a new facility are viable options for an BMP.

These two options were considered for Spokane, Albuquerque, and Salt Lake City. Only Salt Lake City has existing oxygen plants. Two of these oxygen plants produce only liquid oxygen, which is more expensive than gaseous oxygen from a new, dedicated plant. The other two plants (owned by Union Carbide) produce a total of about 800 tpd of 99.5 percent oxygen at 50 psig but have no excess capacity for use by an BMP. Therefore, for any of the three sites, a new dedicated plant is required, whether the BMP owner owns the oxygen plant or purchases oxygen "over-the-fence." Purchased oxygen from a new dedicated plant will be used in the design basis. The economics and efficiencies of scale achievable in a large plant (1000 tpd plus) are not available at these sites, and the oxygen price will reflect this.



Oxygen production via air separation is efficient but energy intensive. Depending on plant size and delivery pressure, approximately 240-400 kWh/ton of oxygen are required. Since power costs typically represent about 40-50 percent of the cost of oxygen (at 4-5¢/kWh), local power costs (power and demand charges) can significantly affect methanol production costs. A 1¢/kWh change in electricity cost will cause about a 1¢/gal change in methanol cost.

3.4 WOOD AVAILABILITY

When potential BMP sites were evaluated with respect to wood supply, three major considerations were identified: (1) the availability of a sufficient supply of wood (the BMP requires 311.4 tpd of 50 percent moisture wood), (2) the location of this wood within a reasonable transportation radius from the BMP, and (3) the existence of an adequate wood harvesting season.

Timber Management Offices of the U.S. Forest Service were contacted for the Salt Lake City, Albuquerque, and Spokane areas. Information received indicates that, of the three potential locations, Spokane most satisfactorily meets the criteria.

Spokane has several options for wood supply. To the north, the Colville National Forest has an annual potential yield of 120 MM board ft (approximately 660 wet tpd), consisting largely of Douglas firs. This wood is harvested anywhere from 40 to 150 miles from Spokane, with most available at 100 miles. Currently, the Colville National Forest does not enter into long-term contracts. Usual contract length is 3-5 years. East of Spokane, the Panhandle National Forests, in northern and northwestern Idaho, have a potential yield of 265 MM board ft/yr (approximately 1452 wet tpd) of saw timber. Twenty to 30 MM board ft/yr (110-164 wet tpd) of this is dead wood and is inexpensive, \$0.60/100 ft³ (\$0.25/wet ton in the forests). Additionally, an estimated 1120 wet tpd of recoverable wood residue is generated in the Panhandle National Forests, with 427 wet tpd specifically from the Coeur d'Alene area. Coeur d'Alene, Idaho is approximately 30 miles



east of Spokane. The residue is sold by competitive bidding. The wood from the Panhandle National Forest is available at an average of 50-60 miles from Spokane. Some is located as close as 20-30 miles. The farthest is about 100 miles.

Not all forest land in the Spokane area belongs to the National Forest Service. State and private lands are also potential suppliers of wood. Chipped wood is available commercially in Coeur d'Alene at \$35-\$55 a bone dry ton, depending on cleanliness. Typical moisture content is 40-45 percent water. A 2 month stockpile would be necessary for operation in Spokane. There are generally two, 6 week periods each year, one in the spring and the other in the fall, in which it is too wet to harvest wood.

Although the Salt Lake City area seems to have sufficient wood to supply the BMP, restraints placed on this supply by the National Forest Service reduce the potential of supply to levels below that required by the BMP. Transporting distances from the available wood to a BMP near Salt Lake are unacceptably long, and the amount of stockpiling required by a restricted harvesting season is prohibitive.

For Salt Lake City, the potential wood supply is from three National Forests: the Ashley in the northeastern corner of the state; the Uinta which forms a semicircle around Provo, approximately 50 miles southeast of Salt Lake City; and the Wasatch-Cache in northeastern and north-central Utah. Together this represents an annual potential yield of 74 MM board ft (405 wet tpd) of Lodge Pole pine and Douglas fir. Of this, however, only 31 MM board feet (170 wet tpd) can be contracted for. The remainder of the wood is located in areas that have developmental constraints or are physically inaccessible. An undetermined quantity of slash is also produced. Currently, there is heavy competition for this residue for personal fuel use. There is no readily available wood within a 75 mile radius of Salt Lake City. Land in this area is recreational or has watershed constraints. Realistic transporting distances are 150-200 miles.



Finally, for the Salt Lake City area, the length of the harvesting season is prohibitively short. Because the wood is located in mountainous regions (6,000-10,000 ft elevation), access is limited to approximately 6 months per year. Roads do not open until June or July and become impassable in the winter. This would require provision for a 6 month stockpile of wood at the BMP.

Albuquerque, New Mexico has problems similar to Salt Lake City - limited wood supply, long transporting distances, and a short harvesting season. Three National Forests are potential suppliers of wood for an BMP located in Albuquerque. The Cibola is located approximately 75 miles west of Albuquerque and has an allowable cut of 5-6 MM board ft/year (27-33 wet tpd). Approximately 60 miles to the north is the Santa Fe National Forest which has a potential yield of 40 MM board ft/year (219 wet tpd) and in Taos, approximately 150 miles from Albuquerque is the Carson which has an allowable cut of 30 MM board ft/yr (164 wet tpd). This represents saw timber which currently is utilized by the mills. Additionally, these forests produce a total of 88.5 MM board ft/yr (485 wet tpd) of slash which is used as firewood. This firewood is sold or distributed free of charge to the general population. New Mexico has a significant population that depends on this wood not only for home heating, but also as fuel for cooking. Although, theoretically, there is sufficient wood in the Albuquerque area to fuel an BMP, this wood is a scarce resource and is fully utilized.

Finally, the wood, which is largely Ponderosa pine with some Douglas fir, is located at elevations of 7,500 ft and greater. A reasonable harvesting season, therefore, is 7 months a year. This would require a minimum stockpile of 5 months supply.

3.5 SELECTED SITE

Because of a ready wood supply and relatively high methanol transportation costs, Spokane, Washington, has been selected as the site for the BMP.



Climatic and resource details affecting plant design are discussed in Section 5.1. Environmental restrictions on the design imposed by governmental jurisdictions are discussed in Section 4.5.



TABLE 3-1
METHANOL TRANSPORTATION RATE SUMMARY
 (November 1983)

VIA WATER

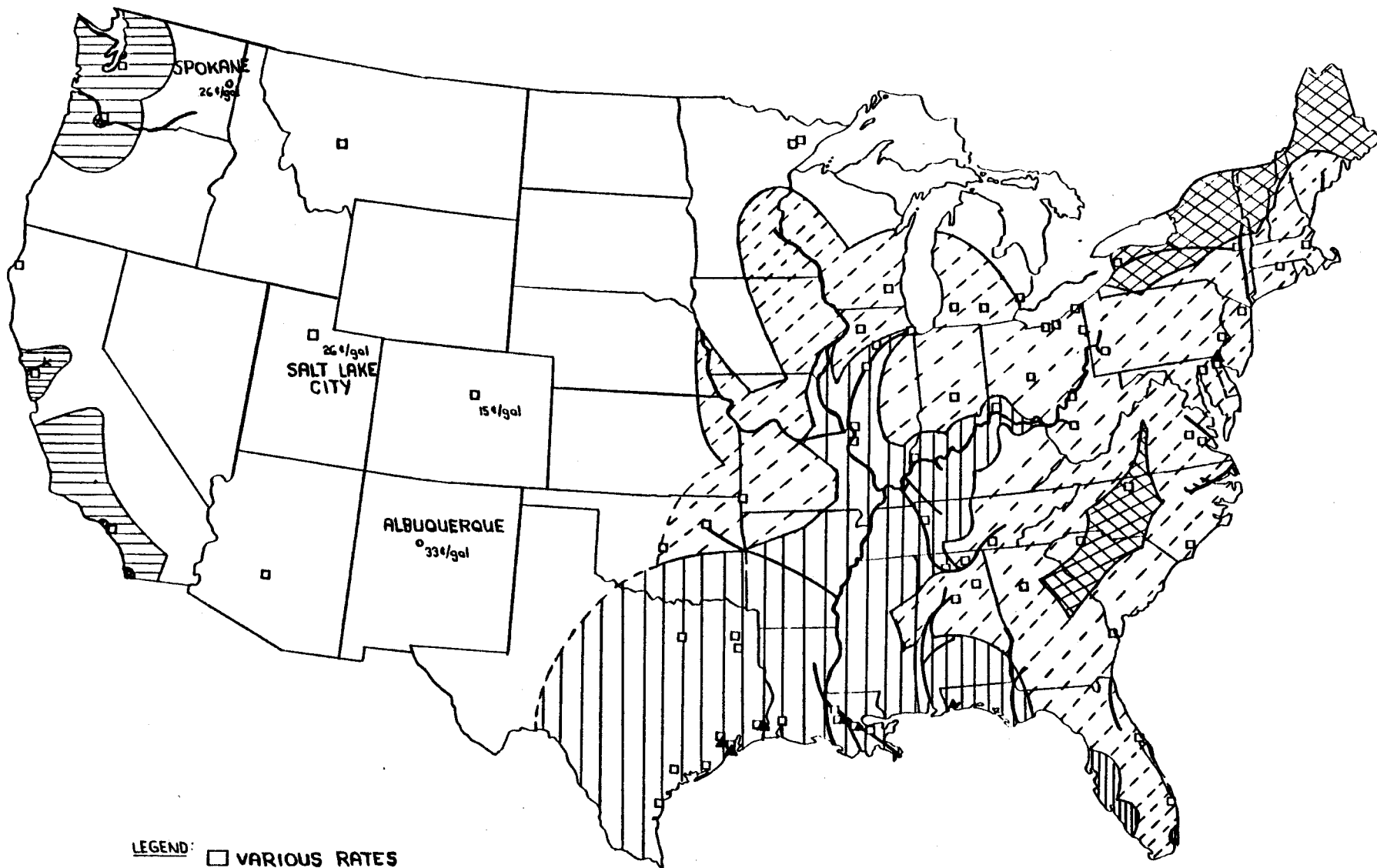
<u>FROM</u>	<u>TO</u>	<u>RATE, ¢/GAL</u>	<u>RATE BASIS, GAL</u>	<u>VESSEL</u>	<u>NOTES</u>
Texas City	Honolulu	19.8			
Texas City	Los Angeles	13.3	840,000 min	Parcel Tanker	
Texas City	Seattle	14.5	840,000 min	Parcel Tanker	
Texas City	Portland	14.5	840,000 min	Parcel Tanker	
Texas City	Jacksonville	6.0	2,100,000 min	Barge	Barge is
Texas City	Miami	4.5	2,100,000 min	Barge	239 ft long
Baton Rouge	Jacksonville	5.9	2,100,000 min	Barge	x 64 ft wide
Baton Rouge	Miami	4.4	2,100,000 min	Barge	
Delaware City	Jacksonville	3.9	2,100,000 min	Barge	
Delaware City	Miami	4.7	2,100,000 min	Barge	
Delaware City	Syracuse	11.2	672,000 min	Barge	Barge is 230 ft long x 43 ft wide
Baton Rouge	Omaha	9.2	400,000 min	Barge	8 mo oper.
Texas City	Omaha	11.2	400,000 min	Barge	8 mo oper.
Baton Rouge	Cincinnati	4.3	420,000 min	Barge	Barge is
Baton Rouge	Chicago	5.0	420,000 min	Barge	195 ft long
Baton Rouge	Louisville	3.7	420,000 min	Barge	x 35 ft wide
Baton Rouge	Knoxville	6.3	420,000 min	Barge	
Texas City	Cincinnati	5.9	420,000 min	Barge	
Texas City	Chicago	6.5	420,000 min	Barge	
Texas City	Louisville	5.0	420,000 min	Barge	
Texas City	Knoxville	7.8	420,000 min	Barge	
Texas City	Pittsburgh	6.4-7.1	420,000 min	Barge	
Texas City	Minneapolis	6.4-7.1	420,000 min	Barge	
Texas City	Birmingham, AL	6.1	420,000 min	Barge	

TABLE 3-1, cont.

VIA RAIL

<u>FROM</u>	<u>TO</u>	<u>RATE, ¢/GAL</u>	<u>RATE BASIS, GAL</u>
Seattle	Spokane	11.6	28,800 min
San Francisco	Fresno	23.2	28,800 min
Texas City	Albuquerque	33.1	28,000 min
Texas City	Denver	15.2	28,800 min
Texas City	Los Angeles	16.8	28,800 min
Texas City	Salt Lake City	26.3	28,000 min
Davenport, Iowa	Omaha	19.7	28,000 min
Pensacola	Atlanta	9.4	28,800 min
St. Louis	Denver	33.6	28,800 min
Delaware City	Binghamton, NY	9.0	to 135,000
Delaware City	Binghamton, NY	4.8	135,000 min
Delaware City	Syracuse, NY	11.4	to 135,000
Delaware City	Syracuse, NY	5.9	135,000 min
Delaware City	Roanoke, VA	12.5	to 135,000
Delaware City	Roanoke, VA	7.3	135,000 min
Delaware City	Harrisburg	6.2	to 135,000
Delaware City	Harrisburg	3.4	135,000 min
Cleveland	Cincinnati	5.1	135,000 min
Texas City	Fresno	19.6	135,000 min
Texas City	Pittsburgh	41.0	135,000 min
Texas City	Minneapolis	38.8	135,000 min
Texas City	Atlanta	20.7	135,000 min





LEGEND:

- VARIOUS RATES
- ▨ 0-5¢/GALLON
- ▧ 5-10¢/GALLON
- ▩ 10-15¢/GALLON
- ▬ 15-20¢/GALLON
- SELECTED COMMERCIALY NAVIGABLE WATERWAYS
- ▲ EXISTING METHANOL PLANTS
- EXISTING OXYGEN PLANTS

FIGURE 3-1
METHANOL SHIPPING COST CONTOUR MAP
 STONE & WEBSTER ▲

SECTION 4

Process Design and Selection

4.1 DESIGN PHILOSOPHY

The design of the biomass-based methanol plant (BMP) is predicated upon the concepts of standardizing construction and maximizing shop fabrication to reduce plant cost. A manufacturing facility is envisioned to produce the several skids of equipment that would be connected in the field. Only minor changes in skid-mounted components are required for different sites. Several plants could be shop fabricated each year.

The scale of the plant is based on the largest vessel in the plant that can be shipped by truck or rail without special permits. Based on estimated gasifier yields and throughput, the constraining vessel is the gasifier. The gasifier yield and scale-up are described below. The base case plant produces 7.68×10^6 gal/yr of fuel grade methanol.

Outside services such as oxygen, water, sewerage, and electricity are available at the site as needed. Otherwise the plant is constructed as a grass roots plant, unassociated with any other chemical plant (except the oxygen plant). Environmental constraints are discussed in Section 4.5. Fuel grade methanol is being produced, consequently purification requirements are lower than for industrial grade methanol (see Section 5.2).

The overall process design is based on three primary requirements:

- Methanol production is maximized.
- Energy integration is used to avoid importing supplementary fuel and electricity.



- Except for the gasifier, all processing equipment selected is commercially available, i.e., no "black box" unit operations are used to resolve design problems.

A brief process description is given in Section 4.4. The detailed process description and design rationale are given in Section 5.

4.2 GASIFIER YIELDS

Originally, a computer simulation model of the SERI gasifier was to have been developed only to evaluate yield changes as a function of gasifier operating pressure. Because of the lack of completed data under pressurized, oxygen-blown conditions, an inaccurate energy and material balance for process design and economic evaluation was likely if experimental data were used exclusively. The use of a simulation model provides better simultaneous closure on mass, energy and elemental balances than can be obtained by forcing closure on the experimental data. SWEC has assumed that if a computer simulation model can accurately simulate the better, more plentiful, SERI experimental data obtained at atmospheric pressure (air or oxygen blown), it can be extrapolated to higher pressures since (1) the fundamental characteristics of the reactor were observed to be the same at both atmospheric and higher pressures, and (2) the thermodynamic basis of the model is sound. The modeling effort described below has been used to provide gasifier yield predictions for use in the process design energy and material balances.

4.2.1 Reaction Zone Models

The SERI downdraft fixed-bed gasifier has been modeled in two parts as shown in Figure 4-1:

1. An overall chemical equilibrium model encompassing the top of the gasifier through the flaming pyrolysis zone (FPZ) from which some "char" has been withheld, and



2. an energy and material balance model for the reacting char zone in which sensible heat from gas cooling is available for carbon-steam and carbon-carbon dioxide reactions. The resulting gas is in water gas shift equilibrium.

4.2.1.1 Flaming Pyrolysis Chemical Equilibrium Model

A computer program made available by the National Aeronautics and Space Administration (NASA) has been used to simulate the flaming pyrolysis zone of the SERI downdraft gasifier. The model, "Computer Program for Calculation of Complex Chemical Equilibrium Compositions, Rocket Performance, Incident, and Reflected Shocks, and Chapman-Jouget Detonations," has been previously used by SWEC to model coal gasifiers. The condition for chemical equilibrium is established in the program by the minimization of the Gibbs free energy. The model is adiabatic. The accuracy of the assumption of overall chemical equilibrium depends on the temperature and residence time of the reacting species. Equilibrium conditions are favored by high temperature (above 1,600°F) and long residence time (several seconds).

Not all of the wood is gasified in the FPZ, since char surface temperatures are well below gas phase temperature. The method used to model this experimentally observed phenomenon was removal of carbon from the NASA program input and reintroduction of the "missing" carbon to the char reaction model.

Table 4-1 illustrates the input information required for the NASA program, species considered, and the program outputs. Since the fuel is normalized to one mole of carbon, the actual dry wood composition of $C_{1.0} H_{1.473} O_{0.635} N_{0.0019}$, after char is removed, becomes $C_{0.8} H_{1.473} O_{0.635} N_{0.0019}$ which is equivalent to $C_{1.0} H_{1.843} O_{0.795} N_{0.0023}$.

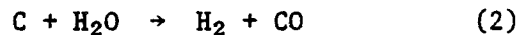
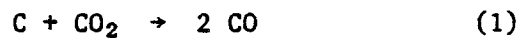
The model is adiabatic while an actual reactor has heat loss (heat losses are higher in the SERI pilot reactor than would be expected in a commercial unit). Nitrogen in excess of oxidant nitrogen has been added to the model to simulate heat loss. This nitrogen is inert and acts as a sensible heat sink



in the program. The excess nitrogen is subtracted out of NASA program results before the char zone model is simulated. By varying the oxygen-to-fuel ratio and the heat loss, equilibrium temperature can be set.

4.2.1.2 Char Zone Model

The char zone model assumes that as the gas from flaming pyrolysis cools, the sensible heat released is available for the following endothermic reactions of the remaining char:



By assuming that the gas is in shift equilibrium at all times, the reaction pathway chosen is immaterial, since any reaction products relax to shift equilibrium. Heat released during the shift is also put into the char reactions. To speed closure, endothermic shifting is avoided and the carbon-CO₂ reaction is preferentially picked to keep the calculated shift equilibrium constant less than or equal to the actual constant at all temperatures. This is shown in Figure 4-2.

For any temperature change from inlet zone conditions, the extent of reaction of char is calculated. Then the amount of shifting required and heat of reaction for the shift are calculated. The shift heat of reaction is then applied to the CO₂-char reaction and the amount of additional shifting required is calculated. This iteration is repeated until the heat of reaction for the shift increment is less than one percent of the initial sensible heat change. Two iterations are shown in Figure 4-2. The temperature change in the reacting zone model is used to determine the moles of char remaining and the corresponding equilibrium gas composition.

The inputs and outputs for the char reaction model are shown in Table 4-2.



4.2.2 Calculation Procedure

For the given dry wood composition, a moisture content and oxygen-to-dry fuel ratio are selected. Carbon representing experimental char remaining after flaming pyrolysis (about 20 weight percent) is removed from the wood. An estimated heat loss is incorporated by adding nitrogen to the oxidant in excess of the oxygen impurity nitrogen.

Fractional Heat Loss =

$$\frac{(\text{lb N}_2 \text{ added/lb dry wood}) (\text{Btu/lb N}_2, \Delta H \text{ from inlet to equil. temp.})}{(\text{HHV Btu} / \text{lb of dry wood})}$$

The actual heat loss depends on the equilibrium temperature reached.

The NASA program is then run. Parameters can be changed until a suitable (but not unique) equilibrium temperature (corresponding to experimentally observed conditions) is obtained. The excess nitrogen is removed. If necessary a non-equilibrium level of methane (from the cracking of tars) is introduced (see below) to account for experimentally observed methane levels. To introduce methane, carbon removed from the NASA simulation is reacted with available equilibrium hydrogen. Since this reaction is exothermic a new "equilibrium" temperature is calculated based on the heat release and the enthalpy properties of the gas phase species. This composition and temperature are used as inputs to the reacting char zone model. The mass flow of gas (usually based on one mole of dry wood plus corresponding oxidant) and the remaining char are also inputs.

The reacting char zone model is run from the initial temperature to lower temperatures, in selected temperature increments (usually 25-50°F), until a suitably low char quantity remains (usually 2-3 percent of the wood carbon). This defines the outlet temperature and gas composition from the gasifier models. If the resulting temperature for proper char reduction is too low (compared to experimental results, i.e., below 1100-1300°F) or all char is



consumed before the gas temperature reaches experimental grate temperatures, the initial temperature for the reacting char zone model is inaccurate. In the former case, the initial temperature is too low and more oxygen is needed to boost the NASA equilibrium model temperature. In the latter case, the initial (NASA program) temperature is too high and less oxygen should be used.

4.2.3 Simulation Results

The composition (ignoring CH_4) of SERI atmospheric, air-blown run 7-26 (personal communication T. Reed to H. M. Kosstrin) was approximated by assuming a 15 percent heat loss with 20 percent of the wood carbon remaining after flaming pyrolysis. Heat loss is defined as sensible heat difference of excess nitrogen from equilibrium temperature to inlet temperature divided by dry wood HHV. The experimental air/fuel ratio (plus heat loss nitrogen) was used. Flaming pyrolysis zone exit temperature was 1530°F (832°C). Grate temperature was 1180°F (638°C). The gas compositions are compared in Table 4-3. By adding methane, char zone temperature would increase to 1608°F (inlet) and 1380°F (outlet) with the same degree of char conversion and heat loss. Differences between experiments and the model are partly due to poor closure on the experimental carbon and oxygen balances. With about 3 percent methane in the product, to reduce char zone outlet temperature to 1202°F (650°C) which was in the experimentally observed range, heat loss would increase from about 15 percent to 15.7 percent. Therefore, changes to the model operating conditions within the range of likely experimental operating conditions can produce balanced yields within the (accuracy) range of experimental yields.

The model is not sensitive to the amount of char withheld from the FPZ since the final result is still in shift equilibrium. Withholding more char raises FPZ outlet temperature which then provides more sensible heat to react to the additional withheld char.

The operating conditions that the model was used to simulate under oxygen-blown conditions are shown in Table 4-4. The resulting gas composition and



operating conditions that were used for the process design material balance are shown in Table 4-5. A gas composition through the reacting char zone as a function of temperature is shown in Figure 4-3. Tar, ash, and sulfur were added later since these are trace compounds and do not affect the results of the model.

Equilibrium methane level was less than that experimentally observed for all cases. Therefore, methane was independently (of the models) added between zone models for all runs. For sensitivity of gas composition to operating pressure, a series of simulations at different pressures were made, maintaining FPZ exit temperature (by changing oxidant level). While equilibrium methane concentration increased with pressure, the resulting level was still less than observed methane levels. By adding additional methane to reach experimentally observed levels, any sensitivity of the model to pressure change is lost. Therefore, model gas composition and yield as a function of pressure will need to be experimentally confirmed to verify a (high pressure) material balance.

The hydrogen-to-carbon monoxide ratio from the gasifier predicted by the model is higher than that observed experimentally. The oxygen-to-fuel ratio required is lower. These predictions can be explained by the higher heat loss in the experiments. To maintain a specified gasifier operating temperature, heat loss must be made up by burning more feedstock-derived raw synthesis gas (H_2 to H_2O and CO to CO_2), consuming more oxygen, and reducing H_2 plus CO yield. Since experimental and model CO_2 levels coincide, fuel hydrogen seems to be preferentially consumed (at constant temperature) when heat loss increases. This is expected since hydrogen autoignites at lower temperatures than carbon monoxide and has wider flammability limits.

4.3 GASIFIER SCALE-UP AND DESCRIPTION

A sketch of the full scale SERI gasifier intended for use in the BMP is shown as Figure 4-4. The gasifier has been scaled up to approximately a ten foot outside diameter size. Allowing for nozzles, the vessel still will be rail shippable or truck shippable with permits. A 9 in. refractory lining



has been assumed in most areas to maintain a maximum skin temperature of 150°F. This leaves the inside diameter at 8 ft-6 in. A 6 in. refractory lining is used in the hopper area below the grate (line CC). A castable type refractory is used. The refractory is assumed to be installed in the fabrication shop then patched and cured in the field.

Wood feed is introduced into the top head of the gasifier at two points. It is intended to keep the top of the gasifier full of fresh feed to minimize freeboard and, to avoid combustion at the top of gasifier (in the freeboard) that would upset the intended temperature profile in the reactor. Oxygen is introduced below the top of the wood bed at the top of the flaming pyrolysis zone (FPZ). To avoid hot spots and cold spots in the FPZ, the oxygen should be evenly distributed across the bed. A sparger system of the type shown in Figure 4-5 should be used. Sparger hole design should be patterned so that tars and particulates cannot clog the holes.

A central agitator has been incorporated into the design to assure an even flow of solids and gases down the gasifier. A water cooled shaft supported and sealed at the top and below the gasifier vessel is used. To accommodate a bottom shaft housing and motor, the char hopper (bottom) of the gasifier has been split into two cones. The agitator arms are installed in the field through the manways (three manways are provided for access to all areas of the gasifier). Four levels of agitation are provided each with three arms spaced 120 degrees apart. The agitator arms are assumed to be made of refractory material that can be screwed or bolted onto the water-cooled shaft. It does not seem feasible to water-cool the arms because of heat losses and potential tar condensation and associated problems. Design of the arms and shaft assembly is a difficult mechanical problem that must be addressed prior to building the gasifier.

At the bottom of the reacting char bed, an inconel grate is attached to the shaft. In the field, layers of ceramic balls are placed on top of the grate to help control the flow of solids out of the char bed. A refractory projection immediately above the grate around the circumference of the gasifier is used to keep the balls on top of the grate; the clearance



between the bottom of the refractory projection and the grate is less than the diameter of the balls. Retaining posts may have to be placed around the gasifier below the grate to prevent deflection of the grate during operation. These posts can be joined to the gasifier shell via refractory clips to avoid hot spots on the shell.

The gas product exits the gasifier immediately below the grate on one side. Approximately 2 ft of gasifier height below the grate is kept at the 8 ft-6 in. inside diameter to reduce particulate entrainment by the gas. The particulate not entrained drops into the two cone hoppers and ultimately to the char removal lock hopper (G-202) below the gasifier. Downstream of the gasifier, a cyclone (M-201) is used to remove entrained solids. These solids are sent to ash and char lock hopper (G-202) via conveyor C-203. Connections are provided for steam blowing the sides of the gasifier hoppers to prevent bridging.

The throughput of the gasifier has been determined by estimating a "heat release rate" per unit of reactor cross section. The heat release rate corresponds to a material flow rate through the fixed bed in terms of higher heating value. This measurement is based on actual phenomena, since the rate of partial oxidation gasification corresponds to the actual combustion heat release rate in the gasifier.

The determination of a reasonable throughput rate for a scaled-up SERI gasifier was based on the actual rates achieved in larger scale downdraft gasifiers and the rates achieved on the SERI 6 in. reactor. Information was obtained on two larger scale downdraft reactors. The first gasifier is a unit built by Biomass Associates and tested by Forintek of Canada. This unit is similar in design to the Swedish units operated (air-blown, atmospheric) during the Second World War and has a 9 in. I.D. grate/choke plate with a 13 in. I.D. reactor. The second gasifier is a unit produced by the Buck Rodgers Company. This gasifier is similar to the SERI design and is 24 in. I.D. but operates at atmospheric pressure and is air blown. The Forintek unit has operated on wood at an average grate area loading of 2.0×10^6 Btu/hr-ft² giving approximately 3 percent char and less than



1 percent tar. In the 24 in. I.D. vessel gasifying wood, the Buck Rodgers Company guarantees a throughput equivalent to 1×10^6 Btu/hr-ft² but has operated at approximately 2.55×10^6 Btu/hr-ft².

It is recognized by gasifier experts that as the scale of the units is increased it is difficult to maintain throughput rates demonstrated in smaller units. For a 8.5 ft diameter, 56.7 ft² cross section, SWEC has assumed a heat release rate of 1.78×10^6 Btu/hr-ft². This is equivalent to approximately 12,000 lb/hr of dry ash free wood with a 8,417 Btu/lb HHV. The effects of different heat release rates on the base case design economics are discussed in Section 7.7.

4.4 DESIGN SYNOPSIS

The outline of the process design is discussed in this section. The detailed process description can be found in Section 5.2. Specific process options are discussed in Section 5.3. A general discussion of processing requirements and options is given below.

Unless other products are more valuable, methanol production should be maximized. Energy integration is important to reduce fuel and electricity imports. All equipment other than the gasifier should be commercial. Other assumptions and constraints depend upon the basic assumptions and environmental considerations (see Section 5.1 and below).

The sequence of process steps for methanol production from wood via gasification is straightforward. The wood to be gasified must be sized and dried to specification. This requires screening for oversized and undersized material. The prepared wood must be pressurized in a lock hopper in order to be fed to the gasifier. The raw gas must be converted to the proper stoichiometry for methanol synthesis (see Section 7.5). Because the shift conversion of carbon monoxide to hydrogen (to produce the proper hydrogen-to-carbon oxides ratio) produces additional carbon dioxide, carbon dioxide removal occurs after the shift operation.



The carbon monoxide shift is insensitive to operating pressure and can occur at gasifier pressure. The removal of carbon dioxide is dependent on carbon dioxide partial pressure. Since complete removal of carbon dioxide is not required (nor desired), the pressure at which carbon dioxide is removed is an economic consideration; removal can occur at any pressure between gasification (150 psig) and methanol synthesis (1000 psig).

The method of carbon dioxide removal depends on capital cost and energy consumption. At intermediate pressures between gasification and methanol synthesis, the most economic process is likely to be Benfield scrubbing (see Oil and Gas Journal, January 10, 1977, p 78). Additionally, the Benfield potassium carbonate will pick up any sulfur dioxide generated in the flaming pyrolysis zone of the gasifier and does not need refrigeration. The acid gas scrubbing is located after the first stage of raw gas compression and before the second and last stage of compression. The raw gas is scrubbed to remove particulate and tars before compression.

The next step is to remove any residual sulfur that may remain in the gas to avoid poisoning the methanol synthesis catalyst. Adsorption of reduced sulfur species onto zinc oxide is the only practical method to reduce sulfur to less than 0.1 ppm. Conversion to raw methanol can then proceed. The raw methanol product must then be purified. In this case, the methanol must only be degassed and dehydrated. Inert gases (nitrogen and methane) must be removed from the methanol reactor system to avoid accumulation.

Energy integration is an important design factor. Because most of the major design steps are straightforward, the plant design options deal largely with matching energy sources and sinks to minimize equipment costs and energy import. Design options that have been used include indirect drying of wood to recover the wood moisture latent heat as steam, fluidized-bed combustion of wood fines and other combustible wastes to balance the steam load, and utilization of the methanol synthesis purge gas (after hydrogen recovery and recycle) as diesel fuel to generate plant electricity. Details of these designs are provided in Section 5.3.1.



Proper treatment of waste streams is also an important design factor for the plant. Low cost capture of the hydrogen sulfide in the acid gas with iron oxide, and phenol extraction from gasifier condensate are discussed in detail in Section 5.3.

4.5 ENVIRONMENTAL STANDARDS AND PROCEDURES FOR SPOKANE

4.5.1 General

In 1974, the Washington State Legislature established the Council on Environmental Policy (CEP), and charged it with the responsibility for implementing the State Environmental Policy Act of 1971 (SEPA). As directed by the Legislature, the duties of the CEP were transferred to the Department of Ecology (DOE) on July 1, 1976. The rules governing the review of projects by DOE are currently under review by the Legislature and changes may be made in the present rules.

In the case of the Spokane methanol-from-wood process plant, the two most important environmental concerns relate to air and liquid waste emissions from the plant. An agency of the City of Spokane will be designated as the "lead agency" and will be responsible to determine if the impact of the project exceeds a "threshold limit" and, therefore, requires the preparation of an Environmental Impact Statement (EIS). The lead agency bases its determination in part on information presented by project's sponsor on the "environmental checklist" form.

The project sponsor and the designated lead agency can mutually agree that an EIS is required in which case the environmental checklist is omitted. When an environmental checklist is prepared, and the lead agency does not find that threshold limits are exceeded, approximately two months are required from the date of submittal for the review and approval of all interested state and local agencies. If the lead agency does make a "Determination of Significance," approximately 90 days are required from the date of the checklist submittal, in addition to the time required by the sponsor to prepare the EIS, in order to hold hearings, receive comments on the EIS and prepare the final EIS.

The "environmental checklist" procedure can be submitted in parallel with efforts to obtain a state wastewater discharge permit, a state "Notice of Construction" permit and building permits. The issuance of these permits may not be made, however, until the environmental review process is completed. Approximately 90 days should be allowed to obtain the "Notice of Construction" and state wastewater discharge permits.

4.5.2 State Permit Procedures

The state "Notice of Construction" permit can be issued within 90 days for a small source. There is no specific time limit to begin construction, but the period allowed is usually between 18 months and five years.

4.5.3 Air Emissions

Air emission points from the methanol plant are the wood solids and tar stream incinerator (maximum firing rate of 12×10^6 Btu/hr), and either a purge gas steam boiler or diesel engine (maximum firing rate of 15×10^6 Btu/hr).

All air emissions are likely to be less than 100 tpy so a prevention of significant deterioration (PSD) permit would not be required, though a state "Notice of Construction" permit would be. Primary responsibility for implementing and enforcing air pollution regulations lies with Washington's local agencies. For Spokane, the agency is the Spokane County Air Pollution Control Authority. The state agency retains jurisdiction over statewide sources such as pulp mills, aluminum plants, and auto emissions. Washington's ambient air quality standards for SO_2 and ozone are more stringent than the federal standards. In addition, there are ambient standards for fluorides and settled particulates (fallout).

Best Available Control Technology (BACT) generally is required. Since Spokane is a nonattainment zone for particulates, any new source has to show that it would not delay attainment of standards, and apply Lowest Achievable Emission Rate (LAER) if particulate emissions are over 100 tpy. If



emissions are less than 50 tpy, LAER is not required, and if between 50 and 100 tpy the issue is uncertain.

Problem emission limits are:

Particulates:	0.05 gr/dscf (standard conditions are 60°F and 29.92 in. of Hg; corrected to 7 percent oxygen).
SO ₂ :	Less than 1000 ppm corrected to 7 percent O ₂ .
NO _x :	BACT, probably 0.4 lb/10 ⁶ Btu for the incinerator.
Opacity	20 percent
Volatile Organic Compounds (VOC)	Only regulated in ozone nonattainment areas, which Spokane is not.
Odors:	"Recognized good practice and procedure to reduce odors to a reasonable minimum."

4.5.4 Wastewater Effluents

The site for the methanol from wood plant has been assumed to be accessible to the Spokane sewer system. Since the plant will not discharge wastewater directly to a receiving water, the plant's wastewaters must only be pretreated for discharge to the City of Spokane's municipal treatment plant. The pretreatment required includes adjusting the pH between 5.5 and 9.0, and removal of any substances which are present in concentrations high enough to be noxious, malodorous or toxic to the biota present in the biological treatment process.

A review of the expected waste characteristics indicates that pretreatment of certain waste streams will be required. High phenol concentrations in the raw gas condensate process waste stream may produce toxic effects at the municipal treatment plant. On a preliminary basis it has been decided, therefore, to limit the phenol concentration from the raw gas condensate to



no more than 100 mg/l. The pH of demineralizer regeneration wastes may fluctuate outside the allowable limits. A neutralization system will be required to assure compliance of this waste with the allowable limits.



TABLE 4-1
NASA Equilibrium Program Inputs and Outputs

Inputs

Fuel First Component Molecular Composition	C _{1.0} H _{1.843} O _{0.795} N _{0.0023} (dry, ash free wood)
Fuel First Component State	Solid
Fuel First Component Temperature	298°K
Fuel First Component Weight Fraction	0.8235
Fuel Second Component Molecular Composition	H ₂ O (wood moisture)
Fuel Second Component State	Liquid
Fuel Second Component Temperature	298°K
Fuel Second Component Weight Fraction	0.1765
Oxidant First Component Molecular Composition	O ₂
Oxidant First Component State	Gas
Oxidant First Component Temperature	389°K
Oxidant First Component Weight Fraction	-
Oxidant Second Component Molecular Composition	N ₂
Oxidant Second Component State	Gas
Oxidant Second Component Temperature	389°K
Oxidant Second Component Weight Fraction	-
Oxidant/Fuel Ratio	wt/wt
Pressure	atm. (absolute)

Outputs

Temperature (°K)	
Gas Molecular Weight	
Gas Mole Fractions (example shown)	
CO	0.23659
CO ₂	0.13849



TABLE 4-1 (Cont)

H ₂	0.25584
H ₂ O	0.20791
NH ₃	0.00010
N ₂	0.16038
CH ₄	0.00070

Products which were considered but whose mole fractions were less than 0.01 for all assigned conditions: (same example)

C(s)	C	CH	CH ₂	CH ₂ O	CH ₃	CN
CNN	CN ₂	C ₂	C ₂ H	C ₂ H ₂	C ₂ H ₄	C ₂ H ₆
C ₂ N	C ₂ N ₂	C ₂ O	C ₃	C ₃ O ₂	C ₄	C ₅
H	HCN	HCO	HNCO	HNO	HNO ₂	HNO ₃
HO ₂	H ₂ O(s)	H ₂ O(l)	H ₂ O ₂	N	NCO	NH
NH ₂	NH ₃	NO	NO ₂	NO ₃	N ₂ N ₄	N ₂ O
N ₂ O ₄	N ₂ O ₅	N ₃	O	OH	O ₂	O ₃
CH ₄	CH ₄ O	C ₃ H ₈	C ₆ H ₆			



TABLE 4-2

Inputs and Outputs for the Reacting Char Zone Model

Inputs

Gas composition (in mole fractions) from NASA program results
Gas mass flow (lb/hr)
Char mass flow (lb/hr)
Initial and final temperatures
Number of temperature increments

Outputs

Temperature
Gas composition by moles and mole fractions
Moles of char remaining

Automatic Defaults

Insufficient char for reaction
Insufficient water for reaction
Insufficient carbon dioxide for reaction



TABLE 4-3

Predicted and Experimental Yields for the SERI Gasifier

Component	SERI Run 7-26 (air blown)		SWEC Computer Model
	Mole Fraction	Mole Fraction	Mole Fraction Adding CH ₄
H ₂	0.1571	0.1578	0.1101
H ₂ O	0.0760	0.0600	0.0799
CO	0.1756	0.1549	0.1409
CO ₂	0.1294	0.1218	0.1273
N ₂	0.4435	0.5054	0.5238
CH ₄	<u>0.0185</u>	<u>0.0000</u>	<u>0.0178</u>
	1.0001	0.9999	0.9998



TABLE 4-4

Operating Conditions for Simulation of an Oxygen-Blown SERI Gasifier

Moisture Content	=	15-20 weight percent (dry basis)
Heat Loss	=	2 percent (of inlet HHV)
Pyrolysis Zone Temperature	=	Approximately 1470-1650°F (800-900°C)
Carbon Conversion	=	97-98 percent
Gasifier Exit Temperature	=	Approximately 1202°F (650°C)
Methane in Exit Gas	=	3-4 volume percent
Pressure	=	150 psig



TABLE 4-5

SERI Gasifier Model

Operating Conditions and Yield for the Process Design Material Balance

Dry, Ash Free Wood Composition	$C_{1.0}H_{1.473}O_{0.635}N_{0.0019}$ (MW = 23.6714)
Moisture Content	16.0 percent (dry basis)
Oxidant-to-Fuel Ratio	0.3447 lb/lb wood (ash free) as fed
Oxygen Purity	98 percent by volume
Oxygen Feed Temperature	241°F
Wood Feed Temperature	77°F (approximate)
Heat Loss	2.0 percent of fuel HHV
Wood Carbon Conversion	97.95 percent
Pyrolysis Zone Maximum Temperature	1638°F
Gas Exit Temperature	1213°F

Exit Gas Composition	Mole Percent
H ₂	35.22
H ₂ O	11.54
CO	30.51
CO ₂	18.87
N ₂	0.68
CH ₄	<u>3.18</u>
	100.00



FIGURE 4-1
SCHEMATIC OF SERI GASIFIER MODEL

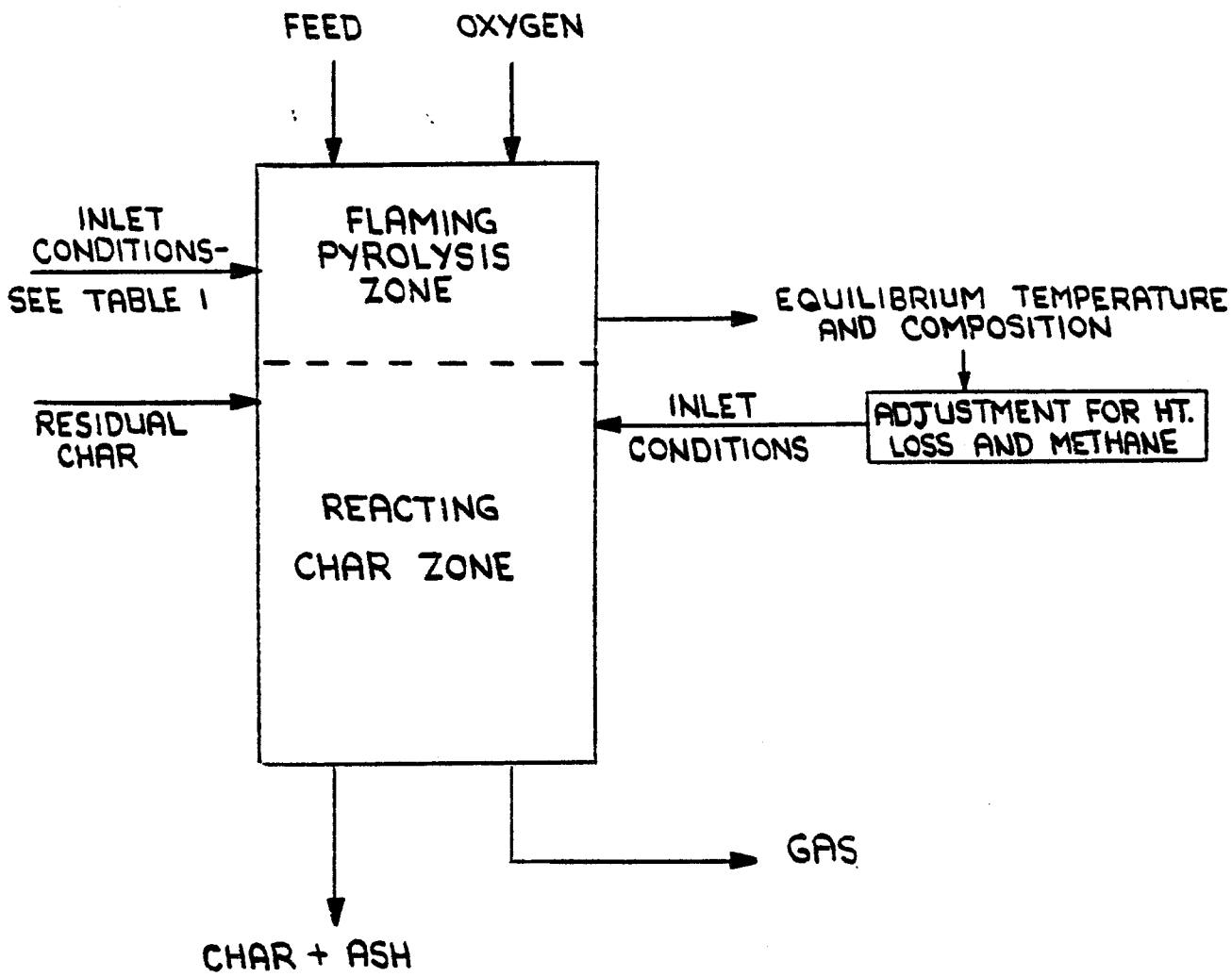
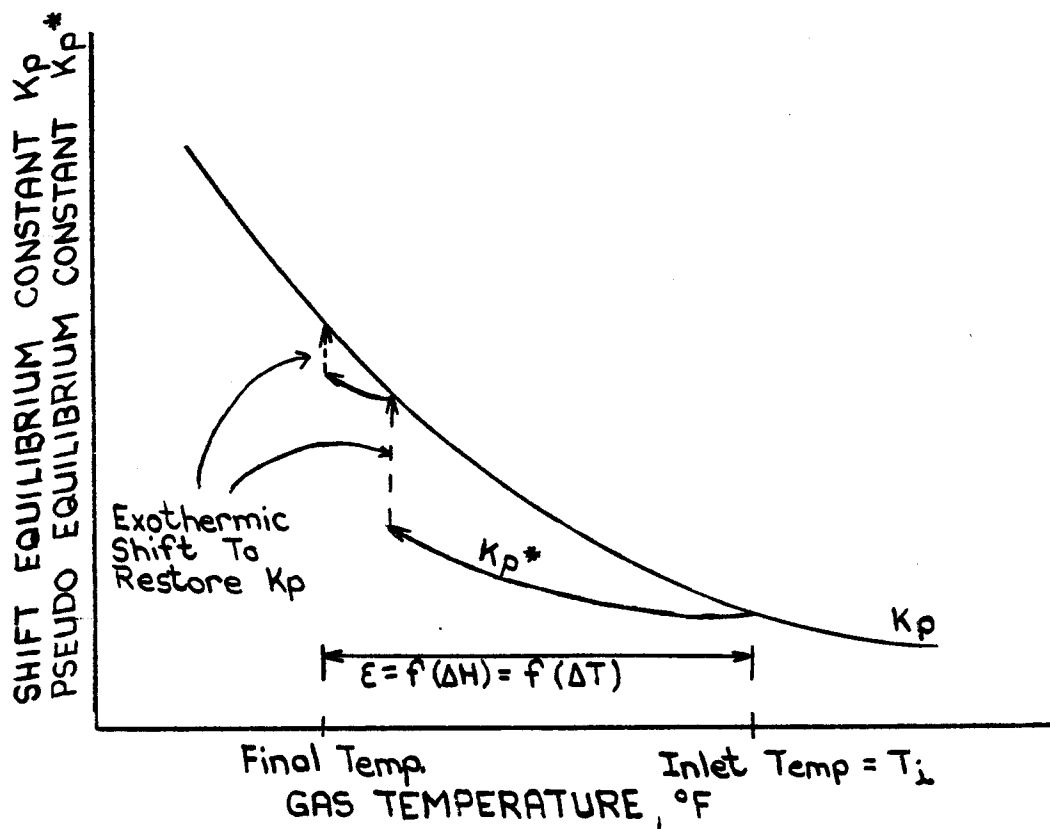
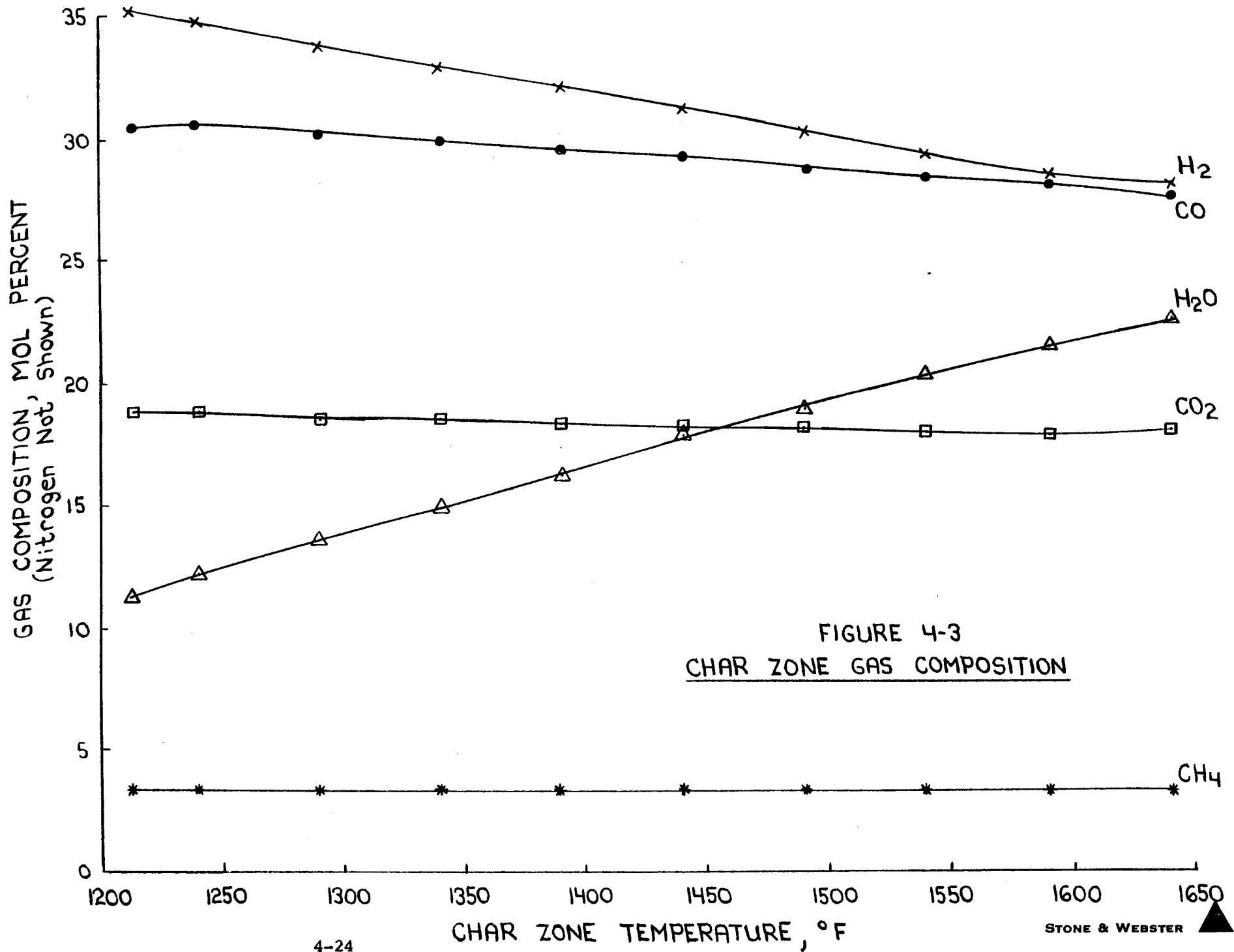


FIGURE 4-2
CHAR ZONE
APPROACH TO EQUILIBRIUM



$$K_p^* = \frac{(\text{moles CO}_2 @ T_i - \epsilon) (\text{moles H}_2 @ T_i)}{(\text{moles CO} @ T_i + 2\epsilon) (\text{moles H}_2\text{O} @ T_i)}$$

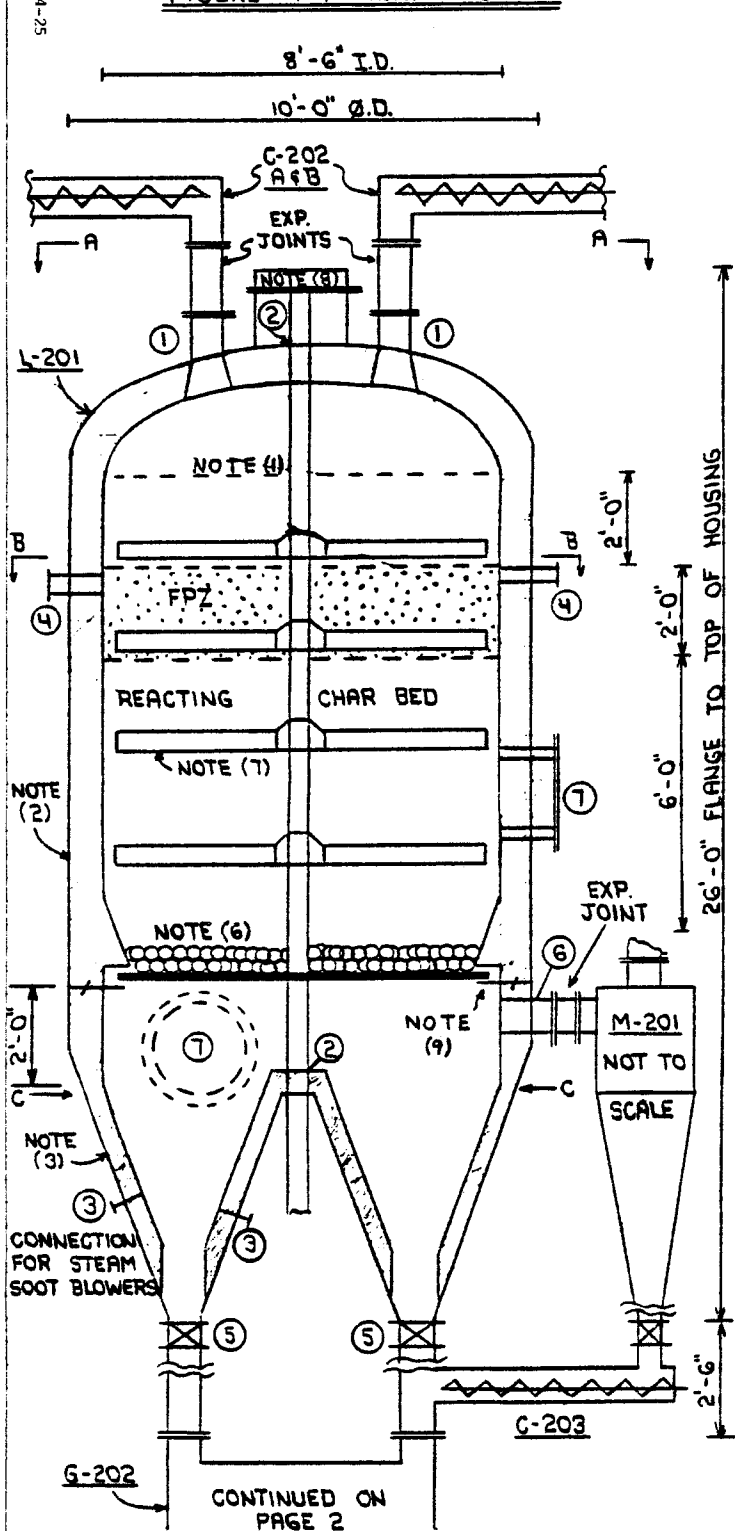
where $\epsilon = \text{moles carbon reacted} = f(\Delta T)$



STONE & WEBSTER ENGINEERING CORPORATION

Client	SER1	Project	METHANOL PLANT	Sheet	1 of 2
Location		Service	GASIFIER/ASH & CHAR HOPPER	Job No.	14665

FIGURE 4-4 SER1 GASIFIER



Shell, Inside Dia.	10'-0"	TOP	INTER	BOTTOM	
Tray Nos.					
PRESSURE					
Psig. Kg/cm ² gs.	Oper. 150				
	Design 200				
TEMPERATURE					
°F. °C.	Oper. 150 Shell				
	Design 600 Shell				
MATERIAL	Shell SA212				
	Trays				
	Valves				
CORROSION ALLOWANCE	Shell				
	Heads				
	Trays				
Tray Spacing					
Type of Liquid Flow					
Type of Trays					
Construction:		Inspection:			
Code:		Emerg. Vac. Design:	Yes	No	
Stress Relieve:	Yes	No	Radiograph:	Yes	No
Insulation:	Yes	No	Type:		Thickness in.
Support:	Skirt Length:	ft.			Lugs. No.
Weight:	Empty:				Full:
NOZZLES:	Mark No.	Size in.	Qty.	Rating	Facing
Feed					
O H Vapor					
Reflux In					
Bottoms					
Drains					
Reboiler Vapor					
Reboiler Liquid					
Vent					
Steam Out					
Safety Valve					
Manholes	7	24	3	300	
Thermowells	-	1	24	300	
Locate in Bottom of DC'S From Tray Nos.					
Gage Glass					
Press. Conn.					
Level Control					
WOOD INLET	1	12	2	300	
SHAFT HOUSING	2	3	2	300	
STM. SOOT BLW	3	1/2	24	300	
O ₂ NOZZLES	4	1	24	300	
ASH OUTLETS	5	6	2	300	
GAS OUTLET	6	18	1	300	

Tray Layout Reference:

Remarks:

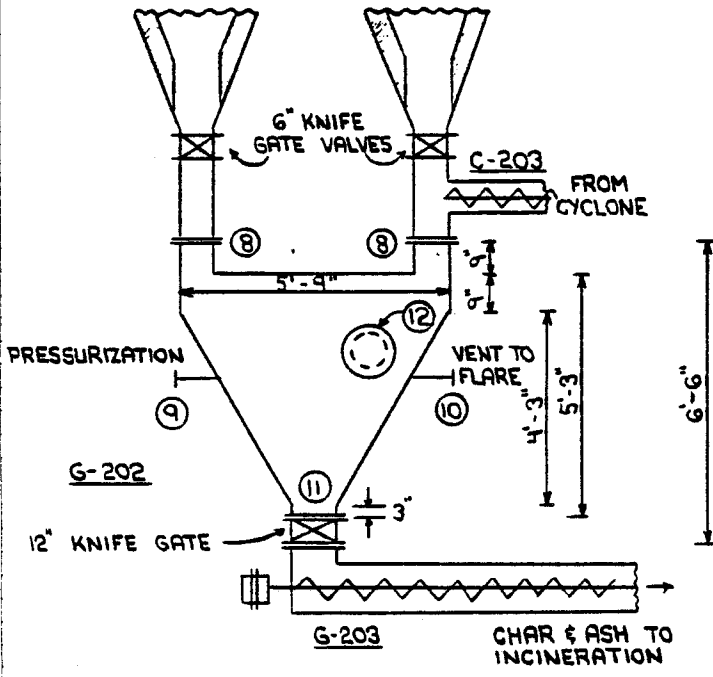
- (1) 2' Wood Height Req'd. Above O₂ Nozzles
- (2) 9" Refractory Lining Above CC
- (3) 6" Refractory Lining Below CC
- (4) Assume Refractory Is 2500°F Rated Castable
- (5) Shell Thickness = 3/4"
- (6) Ceramic Balls; Size and Number of Layers Yet To Be Determined.
- (7) Three Agitator Arms At Each of Four Levels.
- (8) Motor Drive Connection
- (9) Refractory Clips For Deflection Guards

Issue	Date	Description of Issue	By	App'd	Proj.

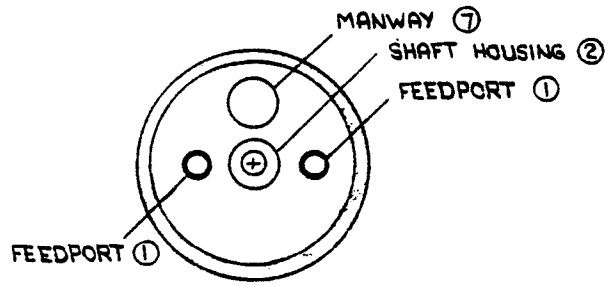
ITEM NO. L-201/G-202

FIGURE 4-5 GASIFIER VESSEL DETAILS

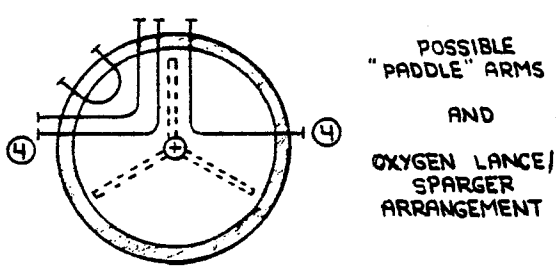
4-26



SECTION A-A (FROM PG. 1 of 2) - 1/2 PREVIOUS SCALE



SECTION B-B (FROM PG. 1 of 2) - 1/2 PREVIOUS SCALE



POSSIBLE
"PADDLE" ARMS
AND
OXYGEN LANCE/
SPARGER
ARRANGEMENT

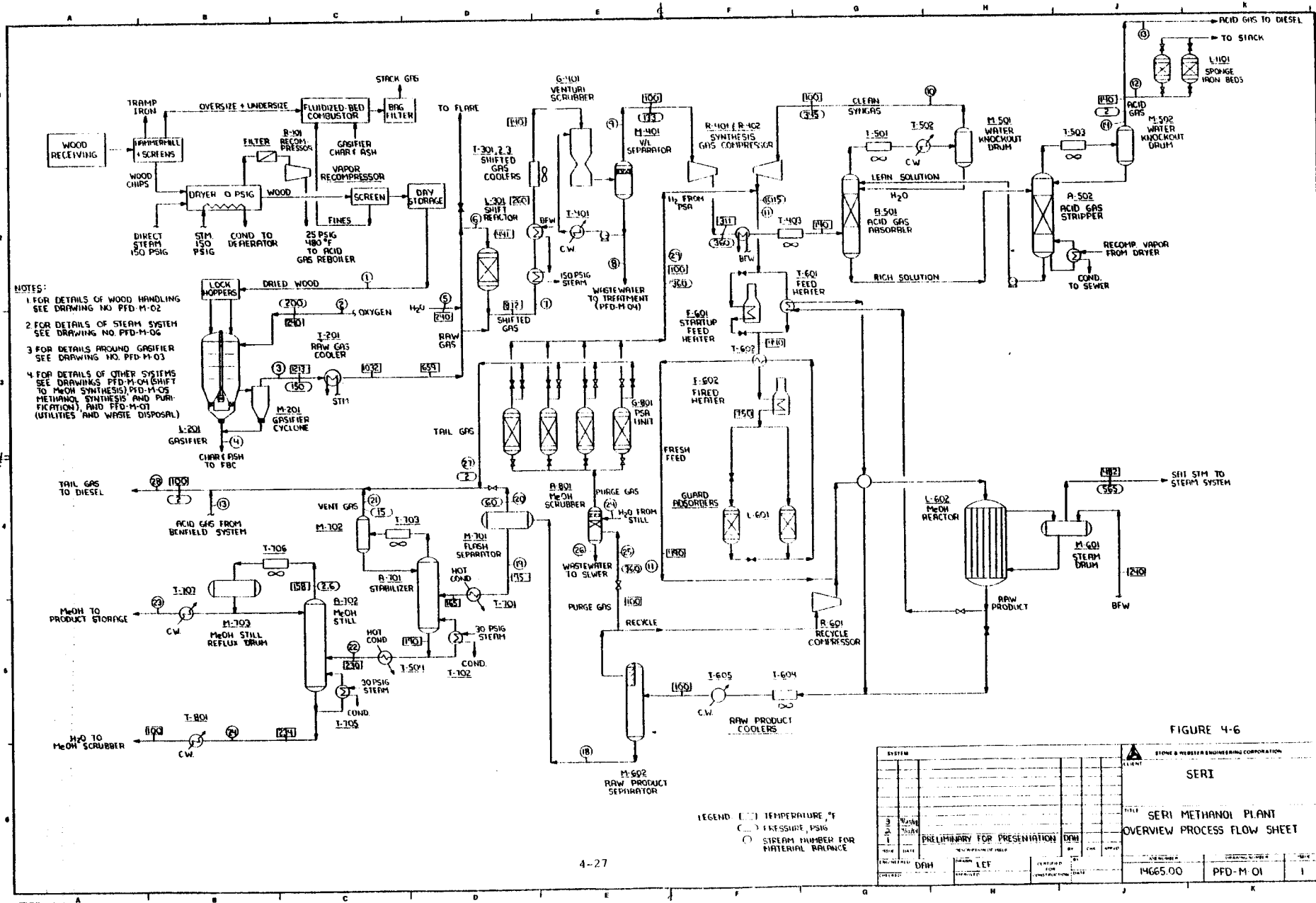
G-202		TOP	INTER	BOTTOM
Shell, Inside Dia.	5'-9"	Tapering		12"
Tray Nos.				
PRESSURE	Oper.			
Psig. Kg/cm ² ga.	Design			
TEMPERATURE	Oper.			
°F. °C.	Design			
MATERIAL	Shell			
	Trays			
	Valves			
CORROSION ALLOWANCE	Shell			
	Heads			
	Trays			
Tray Spacing				
Type of Liquid Flow				
Type of Trays				
Construction:		Inspection:		
Code:		Emerg. Vac. Design:	Yes	No
Stress Relieve:	Yes No	Radiograph:	Yes	No
Insulation:	Yes No	Type:	Thickness in.	
Support:	Skirt Length: ft.		Lugs: No.	
Weight:	Empty:	Full:		
NOZZLES: FOR				
Feed	G-202	Mark No.	Size in.	Qty. Rating Facing
O H Vapor				
Reflux in				
Bottoms				
Drains				
Reboiler Vapor				
Reboiler Liquid				
Vent				
Steam Out				
Safety Valve				
Manholes				
Thermowells	-	1	4	
Locate in Bottom of DCS From Tray Nos.				
Gage Glass				
Press. Conn.				
Level Control				
ASH HOP. INLET	8	6	2	
PRESS. GAS	9	2	1	
VENT TO FLARE	10	2	1	
ASH OUTLET	11	12	1	
MANHOLE	12	24	1	
Tray Layout Reference:				

Remarks:

G-202 SHELL THICKNESS = 3/4"
 MATERIAL SA212

Issue	Date	Description of Issue	By	App'd	Proj.

ITEM NO. L-201/G-202



SECTION 5

CONCEPTUAL DESIGN

5.1 DESIGN AND SITE BASIS

As part of Task 1, a site definition comprising information needed for the process design and economic evaluation is required. This section details this information and the methods used to provide the information. The descriptions of the methanol transportation cost analysis and oxygen and wood availability and cost analysis, which are necessary for locating a site are presented in Section 3.

5.1.1 Plant Size (Capacity)

Plant capacity has been set by the maximum gasifier size and throughput selected (see Section 4.3). Maximum gasifier size has been set at 10 ft OD to be rail shippable and still allow good internal distribution of solids. Gasifier inside diameter (ID) is then about 8.5 ft, allowing for about 9 in. of refractory. Gasifier flow area is about 56.7 ft². Assuming a feed rate of 1.78 MM Btu feed (HHV) per hour per square foot, and a dry wood heating value of 8417 Btu/lb, the dry ash-free wood flow into the gasifier is about 12,000 lb/hr. This translates into a methanol production rate of about 77 tons per day. The gasifier is the largest (diameter) vessel used in the process and is the constraining piece of equipment with respect to plant scale.

5.1.2 Steam Availability, Pressures, and Cost

The energy balance around the process (excluding the oxygen plant) indicates that all plant steam requirements can be met without importing fossil fuels or steam. If part of the electricity for plant use is produced with process-generated steam, no net steam is available for export. The following discussion examines the process options. Saturated steam at 565 psig is generated by the methanol synthesis steam generator. Steam generation from high temperature sources such as the shift and methanol purge gas (if

burned) can also be at this pressure. Steam for power generation will be at the pressure of the methanol synthesis steam generation minus an allowance for superheating and line loss. The maximum steam turbine inlet conditions are then about 550 psig and 750^oF (270^o superheat). Steam will be available at 150 and 30 psig for process steam requirements. The major process steam consumers are wood drying (150 psig), acid gas stripping (30 psig) and methanol distillation reboilers (30 psig). Some of the 30 psig steam required will come from compression of the evaporated moisture from wood drying. Some of the 150 and 30 psig steam needed will be generated (saturated) from a waste wood combustor and diesel engine (operating on methanol purge gas) exhaust.

For process option evaluations, steam at 150 psig or less will be valued at \$3.10/MM Btu which is approximately the cost of producing it from wood at \$30 per dry ton and 80 percent efficiency. Process steam at 550 psig/750^oF will be valued at its equivalent in electrical energy and available steam (\$4.37/MM Btu) as shown in Table 5-1. Direct steam drives instead of electric motors are not appropriate since steam distribution will be more expensive than electrical distribution, and steam power efficiencies drop appreciably as turbine size (power) decreases. With small turbines and inexpensive electricity available, electricity generation using steam without cogeneration is not economical.

5.1.3 Electricity Availability and Cost

All plant drivers will be electric motors. Motors larger than 10 hp will be 480 V, 60 cycles, 3 phase except for the synthesis gas compressor and wood moisture compressor drives which will operate on 4160/2400 V. Motors from 1-10 hp will be 240 V, 60 cycle, 3 phase. Fractional horsepower motors will be 120 V, 60 cycles, single phase. Power will be available from Inland Power and Light at 3.22¢/kWh plus a demand charge of \$5.60/kW/month. Steam can be used to generate plant electric power at 4160/2400 V after process steam requirements have been met. Diesel power generation, if used, will be at 4,160/2,400 V. Any costs associated with bringing power to the plant boundary will not be included in the cost estimate. Line voltage to the plant is assumed to be 13,200/7,200 V.

5.1.4 Cooling Water and Process Water

The source for cooling and process water make-up will be the City of Spokane, Water Department. Year-round average water temperature is 53°F. Water quality is shown in Table 5-2. Cooling water will be recirculated and cooled in a cooling tower to 70°F. The Spokane summer 1 percent wet bulb temperature is 64°F (One percent refers to the amount of time during the four warmest months that the wet bulb temperature exceeds 64°F). Cooling water temperature rise will be 30°F.

5.1.5 Site Elevation and Rainfall

Spokane's elevation is 2357 ft. The plant site will be at the same elevation and is assumed to be flat and unimproved. Average annual precipitation in Spokane is 20 in. The 10 yr maximum 24 hr rainfall is 2 in.

5.1.6 Air Temperatures

Record temperatures in Spokane are approximately -26°F and 108°F. One percent summer dry bulb temperature is 91°F. The five percent summer dry bulb temperature is 85°F. The 99 percent dry bulb temperature is 0°F. (During December through February, the temperature is 0°F or higher 99 percent of time.) A 90°F inlet air temperature was used to size the air coolers. Because insulation was not directly estimated, no winter dry bulb temperature was specified.

5.1.7 Oxygen Delivery and Price

Oxygen at 98 percent purity will be delivered to the plant at 200 psig from a turnkey facility owned and operated by a major oxygen vendor. Oxygen cost will be \$60,000 per month for 60 tons per day (take-or-pay) plus the cost of 400 kWh/ton at prevailing electricity costs (3.2¢/kWh plus \$5.60/kW/month or \$16.29/ton at 90 percent availability). Total oxygen cost at 90 percent (gasifier plant) availability is \$52.82/ton, as received.



5.1.8 Methanol Purity and Product Distribution

The methanol purity specification is shown in Table 5-3. This specification is that used by the California Energy Commission (CEC) for the CEC's neat methanol fleet programs, except for water and ethanol content. Because the methanol produced will likely be blended with gasoline, the water specification has been set at 250 wt ppm instead of 500 wt ppm to allow for some water absorption during blending and storage. Also because of blending, ethanol produced during methanol synthesis is allowed in the product (about 0.5 wt percent). This allows the removal of a methanol/ethanol separation column. Although blenders using methanol for octane enhancement normally use industrial grades of methanol (see Table 5-4), no higher alcohol, ketone, ether, or aldehyde concentration restrictions have been used for this plant. (The only alcohol purity requirement for federal tax reduction eligibility of blends is 190 proof.)

The product methanol will be sold to a fuel blender. It will be the blender's responsibility to arrange for tax benefits and transport methanol away from the plant by rail or truck. The blender's location and distribution network are assumed to be in the vicinity of the plant so that methanol and blended product transportation costs are not a factor in methanol product pricing. Storage capacity for seven days' production and off-loading facilities for methanol product will be included in the plant equipment.

5.1.9 By-products and Wastes and Their Disposition

The by-product and wastes from the plant and their disposition are listed in Table 5-5. No by-products will be sold. No potential market for the carbon dioxide in the Spokane area has been identified.

5.1.10 Wood Handling

Wood will be delivered to the plant by truck as whole tree, green chips. Space for seven days' open storage will be provided. Green chip moisture content is assumed to be 50 percent. Wood will be reduced in size to pin chips with a thickness of 1/4-3/8 in. and a 1 1/2 in. maximum length. The

wood will be dried to 16 percent moisture content. Undersized wood will be combusted, supplemented with acceptable sized wood as needed.

5.2 DETAILED PROCESS DESCRIPTION

5.2.1 Section 100 Wood Receiving and Handling (see PFD-M-02)

Whole tree green chips (nominally 3/8 in. x 3/8 in. x 1 1/2 in. and 50 percent moisture) are delivered by truck to the facility. A supplementary rail branch is shown on the plot plan. The weight of wood delivered is determined by weighing the truck before and after unloading on a truck scale (G-101). Grab samples to determine wood moisture content are taken by hand. The truck dumps its load onto the storage pile. A wheeled tractor shovel (G-102) is used to move wood around the pile and into the reclaim hopper (G-103) that initiates wood processing. To avoid problems with spontaneous combustion and wood decomposition, the pile is used on a first-in, first-out basis. The wood storage pile is approximately one acre in area and contains approximately 2100 tons of green chips.

The tractor shovel transfers wood from the pile into the reclaim hopper (G-103). A drag chain conveyor (part of G-103) is used to move wood out of the hopper onto the scalper feed conveyor (C-101) which is used to transport wood to the top of the material sizing equipment. At the top of the belt conveyor (C-101), a self-cleaning magnetic separator (G-104) is used to remove tramp iron. A scalping screen (G-105) with 35 mm openings (IFO or interface openings, i.e., spacing between adjacent discs) is used to remove oversized pieces and ice lumps from the wood chips. Depending on the operating condition, these oversized pieces are landfilled via a disposal dumpster (G-118) or transferred with the tractor shovel to "fines" storage for eventual combustion and steam generation.

Wood chips less than 35 mm (largest dimension) fall through the scalping screen to the primary (disc) screen (G-106) which passes (accepts) material less than 10 mm (largest dimension). Oversized material falls into a hammermill (hog mill G-107) where it is reduced in size. The milled material is pneumatically conveyed back to the top of the disc screen (G-106).



Small rocks, etc, are periodically removed from the hammermill and land-filled via the dumpster.

The accepts from G-106 fall into the fines (disc) screen (G-108). Material less than 5 mm (largest dimension), i.e., "fines," falls through the screen and is conveyed via C-103 to a storage silo (G-112). Acceptable wood chips greater than 5 mm fall onto belt (dryer feed) conveyor C-102 that feeds the wood dryer. A bypass chute on G-108 that is normally closed allows acceptable chips (greater than 5 mm) to be diverted to the fines conveyor (C-103) if the supply of fines is not adequate to match steam demand. Approximately 7 weight percent of the wood flow to G-108 is expected to be fines.

The "uniform" 5-10 mm (largest dimension) chips acceptable for gasification are dried from a nominal 50 percent moisture content to 17 percent moisture content in an indirectly heated torus disc dryer (G-110). The green chips are conveyed upward from the fines screen (G-108) discharge on a belt conveyor (C-102) that elevates the chips again for gravity flow and empties into a screw feeder (G-109). The screw feeder discharges into the dryer; 150 psig steam is used as the indirect heating media in the dryer. A small amount of direct superheated steam is added to the dryer to promote drying and moisture removal from the dryer. The water evaporated from the wood is compressed from atmospheric pressure to 25 psig in a spiraxial (positive displacement) compressor (R-101) for use as reboiler steam in acid gas removal (see Section 500).

The dried wood chips are removed from the dryer via a rotary feeder (G-119) into a fines (deck) screen (G-111). This screen is used to remove new fines (approximately 1 percent of the dried wood with smallest dimension less than 5 mm) produced in the dryer. These fines are transported via conveyor C-105 into the fines storage silo. The accepts from drying (the gasifier feedstock) are conveyed via a bucket elevator (G-113) to a 24-hr capacity storage silo (G-114). The storage silo has a nominal capacity of 150 tons and a circular traveling screw for reclaiming chips through a central bottom discharge. The chips are removed via the reclaimer onto a gravimetric feeder (G-115) that measures wood flow to the gasifier. The gravimetric feeder discharges onto a covered belt conveyor (C-201) that lifts the chips to an

elevation sufficient to gravity feed two bucket elevators (G-205A&B) that carry the chips into the gasifier lock hoppers.

The fines and accepts required for combustion to generate steam are removed from the fines storage silo (G-112) with a screw reclaimer that spans the circular bottom of the silo and moves through an arc that is pivoted tangent to the circle. The reclaimed material is conveyed up into bucket elevator G-117 via covered belt conveyer C-104. Bucket elevator G-117 feeds the metering bin (G-1111) of the fluidized-bed combustor (G-1101) that combusts the solids for steam generation in boiler T-1102.

5.2.2 Section 200 Gasification (see PFD-M-03)

Dried wood chips are lifted from the storage silo (G-114) and gravimetric feeder (G-115) by a belt conveyer (C-201) to a sufficient height to transfer the chips by gravity through a diverter/flop gate into two bucket elevators (G-205A&B). Each bucket elevator then lifts chips to a sufficient height to transfer the chips through a flop gate into one of the two lock hoppers on each feeder system (G-201A&B). The lock hoppers feed a metering bin. Each set of two lock hoppers is cycled (top open and accepting feed, top closed and pressurizing, bottom open and discharging feed, top and bottom closed and depressurizing). Therefore, the metering bin is always pressurized. Nitrogen is used as the pressurizing gas. A proprietary valving system (supplied by the G-201 vendor) reduces the pressurizing gas requirement.

The flow of wood into the gasifier is controlled by the metering screw in each metering bin. Wood chips are removed from the metering bins via the injection screws. Pressurized bucket elevators (G-204A&B) are used to convey the chips up to the top (inlet) of the gasifier. The wood chips are then moved horizontally from the bucket elevator discharge to the gasifier feed inlets by pressurized screw feeders (C-202A&B). Since the pressure between the gasifier and the metering bin is equalized, there is little blowback of gasifier product into the lock hoppers or leakage of nitrogen into the gasifier.



Details of the gasifier (L-201) are presented in Section 4.3. Briefly, woodchips are introduced into the top of the gasifier where they are partially devolatilized. Oxygen introduced below the top layer of chips is used to partially combust the gas phase species and provide heat for gasification. The devolatilized chips pass through the flaming pyrolysis zone where combustion occurs into a reacting char bed zone where the hot, partially combusted gases react with the remaining carbon in the chips to form carbon monoxide, carbon dioxide, water, and hydrogen. The gases leaving the gasifier are in shift equilibrium.

The solids remaining in the gasifier (i.e., char and ash) fall through a grate into a lock hopper (G-202). Solids recovered from the gasifier cyclone (M-201) are conveyed via a screw (C-203) into G-202. With the upper valves on G-202 closed, the hopper is depressurized, and the accumulated char and ash fall into the removal and quench screw (G-203). Water is sprayed into G-203 to cool the char below its ignition point so that the char will not smolder while it is stored. The quench screw deposits the char and ash onto a conveyor (C-1101) that carries the char and ash to storage before it is burned for disposal in the fluidized bed combustor (G-1101).

The product gas from L-201 exits the gasifier below the grate and enters cyclone M-201 where entrained particles greater than (approximately) 10 microns are removed. The gas is then cooled in T-201 (generating steam) in preparation for shifting the raw gas (Section 300). During startup and shutdown, the flow of gas is directed from T-201 to the flare.

5.2.3 Section 300 Gas Shift (see PFD-M-04)

The raw gas needs to be shifted ($\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2$) to transform the hydrogen-to-carbon monoxide molar ratio from about 1.15 to approximately 2.1 for methanol synthesis. To avoid reduction of the shift catalyst, a minimum water-to-hydrogen (WTH) ratio must be maintained in the shift reactor (L-301). This minimum ratio increases with temperature. Because the shift reaction is exothermic, the limiting WTH constraint must be met at the reactor exit.

Maximum system heat recovery occurs when the amount of water and/or steam used is minimized. If all the raw gas is introduced into the shift reactor, the inlet temperature into the reactor must be reduced to approximately 660°F (from 1032°F) to reach the desired conversion of carbon monoxide to hydrogen. (The shift equilibrium favors higher concentrations of hydrogen at lower temperatures.) The only H₂O added in this case is the water injected to cool the gas. However, the minimum WTH ratio is violated at the resulting reactor outlet temperature of 860°F. If additional water is injected, heat recovery from the raw gas and shifted product is reduced and more water must be condensed from the shifted product.

The amount of water injection and heat recovery can be kept approximately the same by bypassing some of the raw gas around water injection and the shift reactor. The same amount of water added to less raw gas will reduce the inlet temperature to the reactor, favoring more conversion of carbon monoxide to hydrogen and reducing the WTH ratio required; i.e., a lower shift reactor outlet temperature. With the same overall conversion of carbon monoxide to hydrogen, the combined shifted stream and bypass will have approximately the same temperature as when all the raw gas was passed through the reactor.

Approximately 40 percent of the raw gas is bypassed around the shift reactor (L-301). Water is injected into the remainder of the gas via spray nozzles. The shift inlet temperature is approximately 470°F. The shifted product and the bypassed gas are combined and then cooled in a waste heat boiler (T-301) and boiler feedwater economizer (T-302). The waste heat boilers T-301 and T-201 share a common steam drum (M-301). Final cooling (in Section 300) to 140°F is done with an air-cooled heat exchanger (T-303).

5.2.4 Section 400 Gas Cooling and Compression (see PFD-M-04)

The cooled, shifted gas is further cooled to 100°F and washed with water in a venturi scrubber (G-401) to remove particulates and tars and to condense water prior to gas compression. The scrubbed gas is separated from the scrubbing water in a vapor-liquid separator (M-401). The liquid from M-401



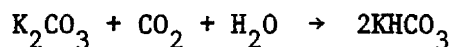
is recirculated to the scrubber via a pump (P-401). The water is filtered to remove particles and then cooled against cooling water in exchanger T-401. A tubular screen filter system (G-402) with two tubes is used. One filter is backwashed with process water while the other is in service. The backwash flow is sewered. Net condensate from M-401 is sent to an API-type gravity separator (M-402) to remove insoluble oils and tars from the condensate. The condensate is then sent to wastewater treatment (Section 900) to remove dissolved phenols before sewerage. The insoluble oils are sent to incineration in Section 1100.

The gas from M-401 is compressed in the first stage of the synthesis gas compressor (R-401). The gas is compressed to an intermediate pressure (365 psig) between the gasifier and methanol synthesis pressures to enhance the operation of the acid gas removal system by increasing carbon dioxide partial pressure. The compressed gas is cooled against boiler feedwater in T-402 and then air cooled in T-403 to control the temperature of shifted gas into the acid gas absorber (A-501). The clean synthesis gas from Section 500 is compressed to methanol synthesis pressure in the second stage of the synthesis gas compressor (R-402). Recycled hydrogen from Section 800 is added to the gas before compression. Since the synthesis gas must be heated before introduction into the (sulfur) guard bed (L-601), no cooling is done after the second compression.

5.2.5 Section 500 Acid Gas Removal (see PFD-M-04)

Carbon dioxide and sulfur compounds must be removed from the shifted gas prior to methanol synthesis. Sulfur compounds will poison the synthesis catalyst. Carbon dioxide conversion to methanol consumes an extra mole of hydrogen per mole of methanol produced compared to carbon monoxide conversion. When initial hydrogen production is less than twice that of carbon monoxide, such that carbon monoxide shifting is required, the carbon dioxide should be removed to minimize carbon dioxide conversion to methanol because carbon dioxide conversion will reduce methanol yield (see Table 5-6). However, some carbon dioxide (approximately 2-4 mole percent) should be kept in the gas because it promotes the catalytic synthesis.

A Benfield acid gas removal system which uses a potassium carbonate solution containing proprietary activators has been specified. The solution absorbs carbon dioxide at high pressure by converting



The reaction is exothermic. The reverse reaction occurs readily at low pressure near the boiling point of the solution when the heat of reaction is provided. The sulfur dioxide generated in the gasifier will be converted irreversibly to potassium sulfate. Most of the hydrogen sulfide and carbonyl sulfide generated in trace quantities in the gasifier will be absorbed by the Benfield solution and desorbed with the carbon dioxide. A small amount of the potassium carbonate will react irreversibly with carbon monoxide to form potassium formate. The potassium sulfate and formate must be purged from the Benfield solution.

The carbon dioxide and sulfur compounds (acid gas) are removed from the synthesis gas in the acid gas absorber (A-501) by countercurrent contact with lean (stripped) Benfield solution. Net evaporation from the lean solution at the top of the column is avoided by cooling the gas leaving A-501 to 100°F, condensing most of the water in the column overhead gas. To cool the column overhead, an air-cooled exchanger (T-501) and a water-cooled trim cooler (T-502) are used. The condensate is separated from the cleaned gas in a water knockout drum (M-501). The condensate is recirculated to the top of A-501 via a pump (P-504).

The (carbon dioxide) rich solution is sent (under its own pressure) to the top of the stripper column (A-502). The column operates at low pressure. Indirect steam is used to "strip" the acid gas from the rich solution by heating and boiling the solution at the bottom of the stripper in reboiler T-504. Since the stripping reaction is endothermic, the vapor rising in the column provides the enthalpy to remove the acid gas from solution. To reduce water losses from the solution, the acid gas is air cooled in T-503 to 140°F. The condensate is removed from the acid gas in knockout drum M-502. The stripped lean solution at the bottom of A-502 is pumped up to the absorber pressure with P-501 and recirculated to A-501.



The bulk of the steam used to run the stripper is compressed evaporated water from the wood dryer (G-110). The remainder of steam needed is generated in Section 1000. Low pressure steam must be used to avoid depositing salts on the solution side of the reboiler (T-504) and fouling the heat exchange surface. A spare reboiler is provided to allow periodic cleaning of both the hot side and solution side surfaces while operating. The hot side of T-504 will be prone to fouling, since the vapor from G-110 will contain traces of wood resins and oils.

Because impurities, mainly potassium sulfate and potassium formate, will tend to build up in the Benfield solution, a purge stream is taken from the rich solution. Makeup solution is mixed in tank (Q-501) and pumped via P-503 to the solution makeup tank (Q-502) that is also used to hold the solution inventory when the columns are drained. The makeup solution is pumped into the system with P-502.

Part of the acid gas is sent to the diesel engines in Section 1100 to reduce the hydrogen concentration in their fuel gas to satisfy the manufacturer's specifications for preventing predetonation. The remainder is sent to the sponge iron bed (L-1101) to remove sulfur species before it is vented. The gas must be compressed to 15-20 psig (in R-1103) to be properly treated in L-1101. Since much of the remaining water is condensed in T-505 using cooling water prior to compression in R-1103, the condensate is removed from the vapor in knockout drum M-504. The condensate is recirculated back to the stripper (A-502).

The steam condensate leaving T-504 is dirty (compared to the boiler water system) and is sent to the sewer system via an oil/water separator (M-1103). For sewer discharge, the temperature of the condensate should not be higher than about 100°F. To reduce the temperature of the condensate and to prevent vapor being sent to M-1103, the condensate is flashed to near atmospheric pressure in M-503. The steam produced is vented. The liquid from M-503 is cooled from its dewpoint temperature to 100°F against cooling water in T-506 which is spared so that one unit is operating while the other is being cleaned of wood resins and tars.

5.2.6 Section 600 Methanol Synthesis (see PFD-M-05)

Before the synthesis gas can be processed into methanol, the remaining traces of reduced sulfur compounds must be reduced to less than 0.1 ppm to avoid poisoning the methanol synthesis catalyst. A few ppm hydrogen sulfide and carbonyl sulfide will not be removed in the Benfield acid gas removal system. These compounds can be removed from the synthesis gas by adsorption onto and reaction with zinc oxide; but to effectively and economically utilize the zinc oxide, the zinc oxide bed must be operated at approximately 750°F.

The synthesis gas from acid gas removal and second-stage compression (R-402) is heated against methanol reactor effluent in T-601 from approximately 340 to 440°F. During startup, a fired heater (F-601) is used. The gas is then heated in T-602 against synthesis gas leaving the guard bed (L-601). The final temperature increase from 700° to 750°F is provided by a fired heater (F-602). The gas then passes into a fixed bed adsorber (L-601). Two adsorbers are provided so that one is onstream while the other is being recharged with zinc oxide. The zinc oxide is removed after its adsorption capacity has been reached. Zinc oxide lifetime, i.e., bed breakthrough, is determined by the outlet concentration of the reduced sulfur species from L-601.

After cooling the sulfur-free gas against the feed gas to L-601, the synthesis gas is combined with recycled gas from the methanol raw product separator (M-602) and recycle compressor (R-601). The fresh feed and recycle streams are heated against synthesis reactor effluent in T-603 to obtain the proper reactor feed temperature.

The methanol reactor (L-602) uses the Lurgi design. The reactor is similar to a vertical shell and tube heat exchanger with a fixed tube sheet. The shell side contains boiler feedwater which picks up the heat generated by the reaction, producing steam and keeping the reaction temperature favorable for a good methanol yield. The synthesis and recycle gas pass through the tubes which are packed with catalyst. The steam produced by the exothermic reactions is sent to a steam drum (M-601) for vapor/liquid separation. The



water circulates by natural convection. The pressure at the steam drum sets the steam operating temperature and the methanol reaction temperature.

The product gas leaving the reactor is cooled first in T-601 against feed to the guard bed, then in T-603 against feed to the methanol reactor. To condense the raw methanol, the product gas is air cooled in T-604, then cooled against cooling water in T-605 to 100°F. The liquid product is separated from unreacted gas in M-602. The liquid product consisting of methanol, water, ethanol, and other trace by-products is sent to purification (Section 700). Most of the unreacted gas is recycled back to the methanol reactor via recycle compressor R-602. To remove nonreacting gases (nitrogen and methane) from the system, a purge stream is taken from the gas leaving M-602. The purge stream is treated in Section 800 to recover hydrogen lost with the purge.

5.2.7 Section 700 Methanol Purification and Storage (see PFD-M-05)

Because the product methanol is intended for use as fuel and not as a chemical feedstock, only the noncondensable gases (light impurities, carbon monoxide, hydrogen, nitrogen, methane, and carbon dioxide) and water must be removed from the raw product. Therefore, the equipment required for distillation is much smaller per unit of methanol produced and less complicated than that required for chemical grade methanol production.

The raw product from M-602 is reduced in pressure to 60 psig. The gases flashed are separated from the liquid in flash separator M-701. The vent gases are sent to the diesel engines (Section 1100) to be burned. During startup or upsets, liquid from M-701 is diverted to intermediate tank Q-701. The contents of Q-701 are reintroduced to M-701 via pump P-701.

The remaining noncondensable gases are removed from the liquid in the stabilizer (A-701) which operates at about 15 psig. The stabilizer is a column designed to provide sufficient methanol vapor flow to strip all the remaining noncondensable gases from the liquid. The liquid feed to A-701 is preheated against steam condensate in T-701. Steam for heating the column

bottoms is provided through a reboiler (T-702) that uses 30 psig steam. The methanol vapor leaving the column is condensed and returned. An air-cooled exchanger (T-703) is used. The liquid methanol is separated from the noncondensable gas in M-702 and returned to the column via P-704. A small caustic addition system (G-701) is used to neutralize trace by-product acids from synthesis in A-701.

A separate column is used instead of a flash plus a final vent on the distillation column to reduce the loss of methanol with the noncondensable gases vented. Since the vapor overhead rate in the distillation column is set by distillation requirements, the vapor flow is much higher than in the stabilizer. Consequently, if noncondensibles are vented from the still instead of the stabilizer, the methanol losses are higher. The small scale and minimal cost of the stabilizer equipment and low operating energy economically justifies the recovery of the approximately 250 gallons per day of methanol that would be lost in a flash and distillation column noncondensibles vented.

The liquid from the stabilizer bottom is preheated against steam condensate in T-704 and introduced into the distillation column (A-702). The column is used to reduce the water content of the overhead methanol product to less than 250 ppm (by weight). Most of the trace by-product ethanol goes overhead with the methanol. An air-cooled heat exchanger (T-706) is used as a total reflux condenser. Net methanol is taken out of the reflux drum (M-703) and pumped to storage tanks (Q-702A-D) via pump P-703. Reflux methanol is pumped back into the top of the column with pump P-705. Off-specification product can be sent to the intermediate tank (Q-701). The reboiler (T-705) for the column uses 30 psig steam. Net water and trace, C₃+ alcohol by-products are removed from the bottom of A-702 and sent to Section 800 for use as a water wash.

In the event of an overpressure or emergency pressure release for the flash separator (M-701), stabilizer (A-701), and/or methanol still (A-702), the vented gas is routed to the blowdown drum (M-704) which acts as a surge drum and vapor/liquid separator upstream of the flare. From M-704 any condensate



is transferred back to the flash separator via pump P-702A&B. Vapor is sent to the flare.

5.2.8 Section 800 Methanol Purge Gas Treatment (see PFD-M-05)

The purge gas from the methanol synthesis loop contains mostly methane and hydrogen. For the reasons discussed in Section 5.3.1, most of the hydrogen is removed from the purge gas with a pressure swing adsorption (PSA) unit (G-801). Methanol present in the purge gas will not readily desorb from the molecular sieve material used in the PSA unit. Therefore, it is washed from the purge gas upstream of G-801. The water removed from the raw methanol during purification is used as wash water. The wash water is cooled against cooling water in T-801, then pumped (using P-801) to the pressure used in the scrubber (A-801).

The PSA unit operates at about 365 psig. This pressure is less than the pressure of the purge gas as it leaves M-602 (raw product separator), but it provides a good hydrogen recovery (less gas is lost during depressurization) and allows the hydrogen that is recovered to be recycled back to the synthesis gas after the first stage of compression. Therefore, the pressure of the purge gas is reduced before the gas enters the PSA unit.

The purge gas from M-602 at reduced pressure is sent into the methanol scrubber (A-801). The wash water from A-801 is sewered. The washed gas enters the PSA unit in which all species but hydrogen (and some nitrogen) are adsorbed onto the bed material. The hydrogen passes through at the operating pressure.

Four vessels are provided for G-801 and they are cycled to provide continuous operation. All the adsorbed species are desorbed when the pressure in each of the vessels is dropped to near atmospheric. This tail gas is sent to the diesel engines (Section 1100).

5.2.9 Section 900 Wastewater Treatment (see PFD-M-04)

Allowing for temperature, pH and volume control, and removal of miscellaneous oils, prior to discharge, the only wastewater stream that cannot be discharged to a municipal treatment plant is the net condensate stream from Section 400. Because this stream contains a high concentration of phenolic tars, it must be treated to avoid upsetting the biological agents used in a municipal sewage treatment plant. A phenol extraction system is used instead of a two-stage aerobic treatment system (see Section 5.3.2 for a comparison).

The water discharge from oil/water separator M-402 is pumped via P-904 into the top of the extraction tower (A-901). Water flows downward by gravity countercurrently to isopropyl ether (IPE) that extracts the phenols out of the water phase. The water leaving the bottom of the tower is sent to the sewer via an oil/water separator (M-1103). The IPE leaving the top of A-901 is heated against IPE-phenol distillation column (A-902) overhead vapors in T-901, then introduced into the middle of A-902.

After T-901, the overhead vapor from the still is cooled and condensed in T-902 against cooling water. A condensate drum (M-901) is used to separate noncondensed gases from the liquid IPE. The condensed IPE is recirculated back to A-902 by gravity and to A-901 with pump P-902. Phenols are moved from the bottom of A-902 and pumped via P-903 to storage (Q-1102). Low pressure steam is the heating media for the IPE - phenol distillation column reboiler (T-903).

Since IPE is slightly soluble in water, IPE must be added occasionally to the system. IPE is stored in Q-901. If oxygen is present, IPE can combine with oxygen to form potentially explosive peroxides. To avoid peroxide formation, a small amount of sodium sulfite is added to the water being treated and to the IPE fed to the storage tank (Q-901).



5.2.10 Section 1000 Process and Boiler Water Treatment and Steam Plant
(see PFD-M-06)

City of Spokane water is heated from approximately 53°F to 90°F against diesel cooling water in T-1001 before entering the activated carbon filter (G-1001) that is used to remove chlorine in order to protect the demineralizer resins. The water next enters the mixed bed demineralizer (G-1002) that removes trace quantities of dissolved minerals from the water. G-1001 is regenerated by backwashing. G-1002 is regenerated by ion exchange using caustic and hydrochloric acid stored in Q-1002 and Q-1003, respectively. The accumulated salts are removed by backwashing. These regeneration wastes are treated in Q-1105.

Demineralized water is stored in Q-1001. Sodium phosphate is added via G-1004 to control the pH and hardness of the process water. The process water is pumped to the required pressure with P-1003. Water for the shift is heated to 240°F in the first-stage compressor cooler (T-402). Demineralized water is also sent to the deaerator (M-1001) for boiler feedwater (BFW) makeup via P-1004 and T-402.

Three different pressure levels of steam are generated. The highest pressure steam (565 psig) is produced only in the Lurgi methanol synthesis reactor (L-602). Thirty psig steam is produced from the diesel exhaust; 150 psig steam is generated in the shift and gasifier cooling area and by the fluidized-bed combustor (FBC) to meet the plant steam demand. The 150 psig steam produced in the FBC is let down to 30 psig as needed.

Heat for the deaerator is provided by the steam condensate returns. Supplemental 30 psig steam is available for deaeration. Boiler feedwater (BFW) pumps P-1005, P-1006, and P-1007 are used to deliver the BFW from the deaerator to the proper steam drums.

Steam (565 psig) from the methanol reactor steam drum (M-602) is sent to steam turbine R-102 that partially runs vapor recompressor R-101. Exhaust



from R-102 is flashed at 30 psig in M-1002. Vapor from M-1002 is added to the 30 psig steam header. Condensate from M-1002 is sent to the condensate steam separator (M-1004).

Steam (150 psig) from the shift steam drum (M-301) supplemented with 150 psig steam from the FBC steam drum (M-1101) is released into the wood dryer (G-110) and is sent to the dryer's heat transfer discs. Condensate return from the dryer is flashed in M-1003 to 30 psig. Steam recovered from M-1003 is added to the 30 psig steam header. Condensate from M-1003 is sent to M-1004.

Steam (30 psig) from the diesel waste heat boiler steam drum (M-1102), supplemented by 150 psig steam from M-1101 (let down to 30 psig) and flashed steam from M-1002 and M-1003, is sent into the 30 psig steam header. This steam header supplies T-702 (stabilizer reboiler), T-705 (methanol still reboiler), and T-903 (phenol still reboiler) and provides the remainder of steam required for T-504 (acid gas stripper reboiler). No condensate is returned from T-504, but condensate returned from T-702, T-705, and T-903 is sent to M-1004.

Any vapor from the condensate steam separator (M-1004) is vented. Condensate from M-1004 is returned to the deaerator via exchangers T-701 (stabilizer feed heater) and T-704 (methanol still feed heater).

5.2.11 Section 1100 General Utilities, Waste Disposal, and Offsites (see PFD-M-07)

5.2.11.1 Waste Incineration and Heat Recovery

Char and ash from the (gasifier area) quench screw (G-203) are conveyed via C-1101 to the char and ash storage bin (G-1112). A bucket elevator (G-1114) is used to convey char and ash from G-1112 into the fluidized bed combustor (FBC) metering bin (G-1111). Wood fines and oversized wood from the fines storage silo and bucket elevator (G-117) are also sent to G-1111.

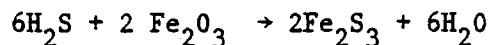


The FBC (G-1101) is used to burn char (with ash), wood fines, phenols from Section 900, and miscellaneous oils. Additional wood is burned to provide sufficient steam for the plant. The oils and phenols are sent to storage tank Q-1102 and pumped into the bed via P-1101.

The FBC system contains a fluidized bed with a bed outlet for large or heavy tramp material. The tramp material is screened to separate tramp material that is discarded into dumpster (G-1109) from bed sand that is reinjected pneumatically with a blower. A combustion air blower is provided to supply combustion and overfire air. The combustion exhaust is cooled in the FBC waste heat boiler (T-1102) and FBC economizer (T-1101). Steam is recovered in M-1101. The cooled gas is pushed into a cyclone (M-1105) for large particulate removal and a baghouse (G-1116) for final particulate cleanup via an I.D. fan (R-1101). Particulate from M-1105 and G-1116 is sent to a dumpster (G-1110). Baghouse exhaust is sent to the FBC stack (G-1105).

5.2.11.2 Acid Gas Sulfur Removal

The acid gas from Section 500 is mainly carbon dioxide with perhaps 50-150 ppm hydrogen sulfide and carbonyl sulfide that must be removed to prevent an odor problem with disposing the acid gas. A sponge iron bed (L-1101) is used to adsorb the sulfur:



The acid gas from M-502 is compressed to 15-20 psig with a positive displacement blower (R-1103). The compressed gas is cooled with cooling water in T-1106 to an appropriate inlet temperature to the sponge iron bed (L-1101). The sponge iron is not regenerated. When hydrogen sulfide begins to break through the bed, a small standby bed (L-1102) is used while the sponge iron in L-1101 is changed. After L-1101 the acid gas (mainly carbon dioxide) is vented through a stack (G-1115).



5.2.11.3 Wastewater Disposal

A schematic of wastewater disposal is shown as Figure 5-7. Condensate from the acid gas stripper reboiler (T-504), treated gasifier condensate from A-901, and the miscellaneous plant drains are routed to oil/water separator M-1103. Light and heavy (compared to water) oils are pumped via P-1105 to the liquid wastes storage tank (Q-1102). Water from M-1103 is sent to the sewer.

Backwash from the scrubber filter (G-402), backwash from the activated carbon filter (G-1001), and runoff from the woodpile are sent to equalization ponds Q-1103A&B that act as holding ponds to control the rate of sewer discharge. The ponds are pumped out at a steady rate with P-1103.

The backwash from the mixed bed demineralizer (G-1102) is sent to the demineralizer wastes sump (Q-1104) where the acid and caustic flows from regeneration are held and mixed. Additional caustic or hydrochloric acid is added to keep the discharge (to sewer) pH between 5.5 and 9.0. The backwash is sent to the sewer via P-1104. Cooling tower blowdown is also sent to Q-1104 for pH control.

The water from methanol synthesis and purge gas washing (A-801), and sanitary wastes are sent directly to the sewer.

5.2.11.4 Flare System

Emergency and/or depressurization vents lead to the flare (G-1102) from all process areas that need to be isolated by valves. The pilot fuel for the flare is propane. Vents from high pressure areas are designed for 15 psig operation to reduce their size. See PFD-M-07 for vent piping details. The flare location is shown on Figure 5-11.

5.2.11.5 Cooling Water

The cooling water steady state operating flows and circuits are shown on PFD-M-07. A mechanical draft cooling tower (T-1105) is used to cool the



circulating water. Cooling water is distributed to the plant cooling water users via P-1102. A 5:1 (maximum) concentration buildup of dissolved solids in the cooling water caused by drift and evaporation is controlled by a continuous blowdown of 2 gpm. City water is used as makeup. Sulfuric acid provided from G-1107 is used to control scaling. Sodium hypochlorite is added from skid G-1108 to control biofouling.

5.2.11.6 Diesel Generators

Two gas-fueled diesel/generator sets (G-1103 and G-1104) are used to provide part of the electric power requirement of the plant and serve as emergency power sources. Normal operating fuel is the tail gas from the PSA unit (G-801) combined with Section 700 continuous vent gases (from M-701 and M-702) and enough acid gas (carbon dioxide) from Section 500 to keep the hydrogen concentration in the fuel below 12 mole percent to control pre-detonation. Propane is available as an emergency, supplemental, and startup fuel. Each diesel is equipped with an air-start system.

The diesels share a waste heat recovery system consisting of a waste heat boiler (T-1104), steam drum (M-1102), economizer (T-1103), and stack (G-1113).

5.2.11.7 Plant Fuel System

Wood fines and wastes fired in the FBC (G-1101) and purge and vent gases from methanol synthesis and purification used in the diesel engines (G-1103 and G-1104) are the major sources of fuel used in the plant. Propane is used to fire the guard adsorber final heater (F-602), pilot the flare, and fire the three startup heaters required - guard adsorber startup feed heater (F-601), (methanol) reactor startup heater (F-603), and shift startup superheater (F-301). Propane is also used to run the diesel generators when tail gas from the PSA unit is not available. A propane storage tank (Q-1101) is provided.

5.2.11.8 Plant Nitrogen System

Nitrogen is used in the plant to pressurize process equipment (especially lock hoppers G-201A&B), purge process equipment, and for instrument "air" for actuating valves. The latter application uses nitrogen instead of a dried, compressed air system because dry nitrogen is readily available from the oxygen plant at low cost.

While nitrogen might be available from the oxygen plant at up to 70 psig, it has been assumed that low pressure nitrogen from the oxygen plant's low pressure column is used. Nitrogen enters the methanol plant at 4 psig. Approximately 50 scfm are sent to the lock hopper compression system. Approximately 27 scfm are sent to compressor R-1102 that produces 1000 psig nitrogen for plant pressurization requirements and as an emergency instrument "air" supply. High pressure nitrogen is stored in M-1104. Another 335 scfm of nitrogen is sent to compressor R-1104 which boosts the pressure to 100 psig for distribution as instrument "air." Variations in instrument nitrogen demand are dampened by instrument nitrogen receiver M-1105. Piping for the instrument system consists of a 2 inch header with smaller piping delivering nitrogen to individual valves.

5.2.11.9 Plant Air System

The plant air system provides air for pneumatic motors and tools, and functions as a back-up supply of instrument "air" in the event of failure of the instrument nitrogen compressor (R-1104). An air compressor (R-1105) rated to deliver 400 cfm at 100 psig and an air receiver (M-1106) are used. Piping for the system consists of a main 2 inch header into required areas with smaller piping to individual outlets.

5.2.11.10 Fire Protection System

The fire protection system is designed to provide 2000 gpm of fire water to the plant fire water loop, to provide Halon as required to the plant control room and to provide foam to the methanol storage tanks.



The water supply for fire protection is provided by one of two fire water pumps (P-1106A&B) each rated at 2000 gpm with a 150 psi differential head. One pump is electric motor-driven and the other pump is diesel engine-driven to provide fire protection if all plant electric power is lost.

The fire protection system also includes two jockey pumps (P-1107A&B) to maintain pressure in the fire water protection main at a minimum of 140 psig. If the fire water system pressure drops below 130 psig, the electric motor-driven fire pump (P-1106A) is automatically started. If P1106A is unavailable, P1106B will start automatically at a system pressure of 125 psig. Both pumps operate until manually stopped.

The fire water distribution system consists of a 10 inch diameter underground piping loop surrounding the entire plant. Branch lines with hydrants and hose houses are provided throughout the plant. Fixed monitors are used to protect the columns. The distribution piping also supplies water to the various required spray systems that are actuated by fire detection devices. The areas covered by fixed spray systems are the wood chip handling and feeding systems, combustible gas compressors, diesel generators, storage tanks, process area pipeways, and the fluidized bed boiler area. Plant buildings are protected by sprinkler systems with the exception of the plant electrical and control rooms which are protected by a Halon system.

The methanol storage tanks are protected by a fixed foam system (G-1117) consisting of a 3000 gallon foam tank, foam generators, associated piping, and a building housing the system.

The water supply for firefighting consists of the contents of the equalization ponds (Q-1103A&B) with City of Spokane municipal water (from mains) as backup.

5.2.11.11 Buildings

A maintenance/shop building with locker facilities has been provided. A single building houses the administrative offices, laboratory, electrical switchgear, and control room. A separate heated building is used to house



the water treatment and boiler feedwater systems to avoid freezing of the systems. The fire pumps and foam system are housed in a small building. Sheds are provided for the synthesis gas compressor, diesel engines/generators, nitrogen compressors, and cooling tower chemical feed skids.

5.2.12 Energy and Material Balances

The base case design material balance is shown in Figure 5-8. The base case energy balance is shown in Figure 5-9. Detailed components flows with stream numbers corresponding with Figures 5-1 to 5-6 are given in Table 5-7.

The overall conversion of gasified dry wood to methanol is 53 weight percent (50% based on total plant feed) for the base case. Carbon conversion to methanol is 40 weight percent. Most of the carbon lost is in the form of carbon dioxide (50 weight percent of wood carbon). For comparison, when methane is converted to methanol, carbon conversion is approximately 72 percent. It is estimated that when western subbituminous coal is converted to methanol, carbon conversion to methanol is approximately 47 percent (with purchased oxygen). Process changes that improve the base case yield are discussed in Section 7.

The energy efficiency of converting green wood to methanol is approximately 58 percent. Net drying losses are small because of heat recovery from the wood vapor. Fourteen percent of the wood energy (the methanol synthesis purge gas) is used to generate electricity. For comparison, when methane is used to produce methanol, methane enthalpy conversion to methanol is approximately 59 percent on a higher heating value basis. When western coal is converted to methanol, enthalpy conversion is approximately 54 percent, excluding air separation power.

5.3 PROCESS INTEGRATION AND DESIGN ALTERNATIVES CONSIDERED

The major design options and process integration for establishing a base case are concerned with waste and by-product streams because the basic processing steps (gasification, shift, quenching, compression, acid gas removal, methanol synthesis, and methanol purification) are sequential and



not subject to improvement by rearrangement. The operating conditions of the gasifier are designed to maximize carbon conversion to synthesis gas. The shift must produce sufficient hydrogen to convert the remaining carbon monoxide to methanol. The operating conditions selected for acid gas removal and methanol synthesis are near the optima for producing methanol from a hydrogen deficient feedstock. Therefore, improvements for the initial (base) design have come from integration of the plant and by-product streams for energy consumption reduction and selection of low-cost waste stream processing for safe disposal.

Most of the design options evolve from the consideration of four streams - wood dryer vapor, gasifier condensate, methanol synthesis purge gas, and the acid gas. These options are tempered by the small scale of the equipment, small equipment efficiencies and cost, and the desire to reduce plant capital and operating costs per unit of methanol produced.

5.3.1 Energy Integration

The major potential sources and uses of energy in the plant are listed in Table 5-8. To provide proper temperature differences, at least two pressure levels of steam production must be used - 565 psig for the steam produced by the Lurgi methanol reactor and approximately 30 psig required for the Benfield system. Thirty psig steam is also adequate for methanol distillation. If wood drying is by direct air heating, the production of steam is less than the demand unless about 7-7.5 MM Btu of wood are burned to produce steam. The purge gas is then adequate to dry the wood. However, no power is recovered from the energy sources and purge gas hydrogen is lost.

If the wood is dried indirectly, a third steam pressure is needed, 150 psig, to provide an adequate temperature difference in the dryer. The latent heat of the wood moisture can be recovered by compressing the dryer vapor to about 25 psig (the limit in a single stage spiraxial, positive displacement blower). The economics of vapor compression are shown in Table 5-9. Indirect drying is more efficient, since there are no air heating losses. Steam that is directly injected into the dryer to improve drying mass transfer is still recoverable. Blue haze from a direct dryer - a potential



air pollution problem - is also eliminated, although the condensate from the compressed moisture must be treated and kept separate from the steam system. The wood moisture and direct steam injected into the dryer is nearly sufficient to run the acid gas stripper. Therefore, the dryer vapor can be used to base load the acid gas stripper and 30 psig steam can be used to balance the load.

With an indirect dryer and vapor compression, the resulting steam demands are shown in Table 5-10. The shift steam can be dropped by injecting water into the raw gas. This also reduces heat exchanger surface. The remaining raw gas cooling and shift cooling can be used to produce 150 psig steam. Because there is no demand for 565 psig, this steam can be let down through a steam turbine to produce either 150 or 30 psig steam.

The flow of 565 psig steam is low relative to the design flows of efficient, multi-stage steam turbines. Because single stage turbine efficiencies with small wheels (due to low flow rates) are only about 50 percent efficient, letdown to 30 psig is preferred to get a useful amount of power. The steam turbine can be ganged to an electric motor to reduce some of the electric load. Approximately 200 horsepower can be produced. There are only two motors of this size in the plant (the synthesis gas compressor and dryer vapor compressor). Therefore, the steam turbine is placed on the dryer vapor compressor shaft.

One method to utilize the wood fines and the purge gas is to generate additional 565 psig steam; combined with the Lurgi steam, the flow of 565 psig steam would only be about 27,000 lb/hr. Expanding steam to 30 psig from 565 psig, 750°F (270°F superheat) has a theoretical steam rate of 15.96 lb/kWh. However, the turbine efficiency for this flow rate is only about 65 percent, and the net power produced would only be about 1500 hp. This configuration would require a gas-fired and a wood-fired high pressure boiler. The steam for the indirect dryer would not be available unless two steam turbines were used, extracting at different pressures. This would lower individual turbine efficiencies to about 60 percent or less, and less power would be extracted from the 150 psig back-pressured unit.



A better alternative is to burn the wood fines for balancing the steam load, and use the methanol purge gas (and purification vent gases) as a diesel engine fuel. The wood fines are burned to generate 150 psig steam. Some 30 psig steam can be generated off the diesel exhaust. The indirect steam condensed in the dryer is flashed to produce additional 30 psig steam. These two 30 psig steam sources with the let-down steam from the Lurgi methanol reactor are nearly sufficient to supply all the 30 psig steam needed. A small amount of 150 psig steam must be let down to 30 psig to meet the requirement. The base case steam plant energy flows are shown in Figure 5-10.

The diesel engine(s) firing on the purge gas alone can produce the same amount of power as a steam cycle using the maximum production possible of 565 psig steam because the engines' diesel cycle is more efficient (about 35 percent) than a small, back pressured, noncondensing Rankine cycle (about 14 percent). While Rankine cycle efficiencies of 35 percent are possible, the cycle conditions require higher pressures, higher superheat temperatures, reheating, the more efficient turbomachinery available in larger sizes, and low temperature (vacuum pressure) steam condensers.

The purge gas and vent gases direct from methanol synthesis cannot be used as diesel fuel because of high hydrogen content. Hydrogen's fast flame speed will cause predetonations in a large, slow speed internal combustion engine. Because hydrogen is valuable as a methanol feedstock, removing the hydrogen from the purge gas prior to fuel use is advantageous. A pressure swing adsorption (PSA) unit is used to "remove" the hydrogen by temporarily adsorbing nonhydrogen species on a molecular sieve. Because an efficient PSA system can recover only 75-80 percent of the hydrogen, the remaining hydrogen concentration in the purge gas after PSA treatment is still too high for combustion. The purge and vent gases are diluted with some acid gas (mainly carbon dioxide) to reduce the hydrogen concentration. While the gas heating value declines, the combined purge gas and vent gases can still be used as a diesel engine fuel.

By recycling hydrogen from the purge gas, the amount of carbon monoxide shifted to hydrogen can be reduced and/or the amount of carbon dioxide removed as acid gas can be reduced. The net result is the same; more hydrogen is available to convert carbon to methanol, and therefore less carbon as carbon dioxide needs to be rejected. By reducing the carbon dioxide removal, less energy is needed to run the acid gas stripper, reducing steam consumption.

Two additional benefits accrue by using diesel engines. The acid gas used to dilute the diesel fuel hydrogen concentration also contains trace amounts of hydrogen sulfide which must be oxidized to avoid an odor problem upon discharge. Using acid gas as a fuel diluent reduces the amount of acid gas that must be specially treated. Also, to operate the plant safely, a backup source of electricity, i.e., a diesel generator, must be provided in any case to allow an orderly and safe shutdown of the plant.

5.3.2 Gasifier Condensate Treatment

The condensate from scrubbing the raw gas (to remove tars and particulate prior to compression) will contain tars. No analysis of the pilot gasifier tars has been made, but Stone & Webster has assumed that the tars will be phenolic in composition and therefore pose a danger to surface waters, groundwaters (a domestic water supply limit of 1 mg/l), and sewage systems. This is a worst-case assumption. Because the tar/phenol concentration in the condensate is expected to be about 2500 wt ppm, it has been decided to reduce the phenol concentration in this stream to 100 ppm by weight (100 mg/l) to avoid upsetting the City of Spokane municipal sewage system to which the condensate will be discharged. The extraction system described in Section 5.2.9 has been chosen.

The alternatives to extraction are listed in Table 5-11. Aerobic digestion via activated sludge would be the preferred method if a discharge to the sewer were not possible. Activated sludge can reduce phenol levels to less than 1 ppm in a two-stage process, but a large quantity of dilution water must be used because phenols are toxic to biological processes at "high"



concentrations. This additional volume of water to be treated raises the relative cost of aerobic digestion for low removal efficiencies. However, to reach a discharge level of less than 1 ppm by solvent extraction, the equipment required to obtain the separation would tend to grow exponentially because of mass transfer limitations.

For reduction to less than 100 ppm by aerobic digestion, the condensate is mixed with all other water wastes to provide the necessary dilution. Phosphate salts and caustic are added to the combined waste stream to provide nutrients and adjust pH for biological treatment in the powdered activated carbon/activated sludge (PACT) process unit. The unit is designed to provide 30 hr detention of influent wastes and have a mixed liquor suspended solids concentration of approximately 15 g/l. An estimated 190 lb/day of activated carbon is added which enables the system to tolerate a design concentration of 350 mg/l phenol. The effluent from the PACT process is clarified and a portion of the sludge is recycled to the PACT unit while the excess sludge is transported to an aerobic digester for sludge stabilization. The PACT process removes 95 percent of the BOD, 90 percent of the COD, and produces an effluent which contains less than 30 ppm of suspended solid and 10 ppm of phenol (90 ppm with respect to the gasifier condensate). The digester produces approximately 1500 gal per day of 4 percent solids sludge which is trucked offsite for disposal at an approved landfill. The overflow from the clarifier is within the neutral pH range required and is combined with other plant waste streams discharged to the sewer system. (For reduction to less than 1 ppm of phenol, the overflow from the clarifier would be treated in a second, separate PACT unit with a different strain of activated sludge).

The installed capital cost of single stage activated sludge treatment of the combined wastewater flow is approximately \$485,000. The installed capital cost of the extraction unit is approximately \$62,000. Nonlabor operating costs for the activated sludge system are about \$73,000 per year. Operating costs for the extraction unit are \$17,000 annually.



5.3.3 Acid Gas Treatment

The concentration of hydrogen sulfide in the acid gas will range from 35 to 300 volume ppm, depending on the level of oxidation of sulfur species in the gasifier. Disposal of the acid gas produces an odor problem in the immediate vicinity of the plant because hydrogen sulfide can be detected nasally at 0.5 ppm. The hydrogen sulfide content of the acid gas must be reduced before the acid gas can be discharged.

Removal of the hydrogen sulfide is complicated by the low pressure of the acid gas and the preponderance of carbon dioxide. Selective absorption, selective adsorption or oxidation must be used. Several candidate processes are listed in Table 5-12. The exact allowable discharge of hydrogen sulfide will depend on the actual stack location and wind conditions. For this analysis, removal to less than 5 ppm has been assumed.

Sponge iron oxide adsorption/oxidation has been selected because of low capital cost and is described in Section 5.2.11.2. If the plant size was larger, a liquid oxidation process - specifically Lo-CatTM - might be more economic because of the compression power costs required for the sponge iron system would not be required for a Lo-Cat system, although the Lo-Cat capital costs are higher.

If the methanol plant were located in an area where carbon dioxide could be sold, treatment of the acid gas could be left to a carbon dioxide vendor who would purchase the untreated acid gas stream. Acid gas treatment equipment would not be necessary in this case.

5.4 PLANT STARTUP, SHUTDOWN AND EMERGENCY PROCEDURES

5.4.1 General Plant Startup

Startup of the wood-to-methanol plant is dependent on initial supplies of nitrogen, electricity, steam, dry wood chips, and oxygen. The nitrogen is



needed for operation of instruments and controls, purging, and pressurization. The oxygen is needed for gasification. Nitrogen and oxygen are assumed to be available for startup from the oxygen plant. The other requirements are provided from within the plant boundaries. Cooling water, process water, and boiler feedwater systems are assumed to be operable once electricity has been provided.

Electricity for startup is produced from the diesel engines firing on propane (grid electricity can be used, but may result in higher demand charges). The diesels are equipped with compressed air starters. Electricity from the diesels is used to operate the wood handling system and the fluidized-bed combustor (FBC). The FBC is needed to generate steam for drying wood when the dried wood supply from silo G-114 is not adequate for startup. The FBC must be heated to approximately 700°F with propane fired in a startup burner before fines and/or green chips can be combusted.

If enough dried wood is available in G-114, the gasifier can be started before or simultaneously with the wood dryer and wood handling system. The gasifier is brought up to temperature from a cold start by burning wood with low pressure air provided by startup blower R-201. The air is introduced through the oxygen distribution system. Initial combustion is provided by pouring kerosine on top of the wood and applying a wick through the top manway. Once a good char bed and temperature distribution have been established in the gasifier, the air supply is shut off and oxygen is used to begin gasification and pressurization. The gasifier product is diverted to the flare downstream of waste heat boiler T-201. If the gasifier is still hot and full of wood, the gasifier can be started up on oxygen.

Except for drying, the wood handling system can be operated on an on/off basis; 150 psig steam is needed to preheat the dryer (G-110) and indirectly dry the wood for gasification. Until it is needed in the acid gas removal system, the moisture removed in the dryer is vented.

The shift system (Section 300) can be started up as outlined in Table 5-13. The steam for heating the reactor is provided by the FBC and the gasifier



waste heat boiler (T-201). As the shift reactor is brought on line, steam generation from shift boiler (T-301) is added. Once the venturi scrubber system is operational, the synthesis gas compressor can be started, using the recycle loops to maintain upstream pressure. The acid gas removal system (Section 500) is brought on line according to the procedures outlined in Table 5-14. The steam injected into the columns comes from the 30 psig header (the diesels). The steam for operating the reboiler (T-504) comes mainly from the recompressed wood moisture (R-101).

As the shift and acid gas are being started, the methanol synthesis system (Section 600) is brought on line as summarized in Table 5-15. Once the synthesis gas compressor (R-401 and 402) is operating, flow of synthesis gas can be introduced into the guard bed system. The startup feed heater (F-601) is used until the methanol synthesis loop is on-line. Once the guard beds are operating at temperature and the sulfur in the synthesis gas is within the tolerance limits of the methanol catalyst, the synthesis gas flow to the reactor (L-602) can be started.

When the methanol synthesis loop is operating, the methanol purification (Section 700) can be started as outlined in Table 5-16. Purge gas from Section 600 is scrubbed, then sent to the PSA unit (G-801). The tail gas from the PSA unit is sent to the diesel engines that are switched off of propane fuel.

Overall startup time for the plant depends on the temperatures in the plant, since much of the startup requirements involve heating vessels at a controlled rate. From a cold start, total plant startup time should be about 12-14 hours depending on the dry wood supply. From a temporary shutdown (hot start), startup time need only be an hour or so to return the separating columns back in steady state operation.

5.4.2 Normal Shutdown

In a planned shutdown, equipment should be taken out of service in such a manner that rapid changes of temperature and pressure are avoided. The normal shutdown procedures for the individual sections of the plant are given



in Table 5-17. Purging and depressurization is not necessarily required for short shutdowns. The normal sequence for shutdown is the following:

1. Stop synthesis gas flow to the guard absorbers and shut down Sections 600, 700, and 800. Flare synthesis gas. Switch diesels to propane.
2. Stop synthesis gas compressors, flaring raw synthesis gas before R-401. Shut down Sections 500, 400, and 300.
3. Stop feed of wood and oxygen into the gasifier. Shut down Section 200.
4. Stop the wood handling system, then the diesel engines and water systems (as the plant units cool down).

5.4.3 Emergency Procedures

The entire methanol plant must be shut down in an orderly fashion if the plant oxygen, nitrogen, electricity, feedstock to the gasifier, cooling water, process water, or boiler feedwater systems are lost.

If plant oxygen is lost, the feed of wood to the gasifier is halted. The gas product is flared before the shift to avoid tar deposition. The diesels are switched to propane. The downstream sections are held under pressure with the syngas compressor shut off if the interruption is temporary. Otherwise the plant sections are stopped using the normal procedures, except that R-401 and R-402 are already shut down.

If plant nitrogen (alone or with oxygen loss) is lost, the nitrogen surge drum, M-1104, is used to maintain instrument air until shutdown and venting is complete. Purge nitrogen is not available, but vented steam could be used.



If purchased electricity is lost, the synthesis gas compressor should be stopped and the diesel engines switched to propane. Otherwise ordinary shutdown procedures can be followed. If both diesel engines are inoperable, the plant should not be operating on purchased electricity alone.

If feedstock to the gasifier is temporarily lost, the oxygen to the gasifier can remain on temperature control until it is shut off completely. The gas product should be flared when the oxygen flow drops below a minimum value (that must be determined experimentally) to avoid tar deposition in the shift area. The plant does not have to be shut down until the gas is flared. When the gasifier product is flared, the synthesis gas compressor should be stopped and ordinary shutdown procedures followed. If feedstock loss is not temporary, oxygen flow to the gasifier should be cut back as slowly as possible while shutting the plant down under ordinary procedures. If high temperature or low temperature excursions occur in the gasifier, the gas product should be immediately flared and the oxygen feed stopped.

If any water system is lost, the plant should be shut down immediately starting with the gasifier and synthesis gas compressor. This is unlikely since all pumps are spared and a supply of demineralized water is stored. No water system should be operated continuously if spare pumps are not available for service.

5.5 INSTRUMENTATION AND CONTROL SCHEME (SEE PFD-M-02 TO M-07 AND TABLE 5-18)

5.5.1 Section 100 Wood Receiving and Handling

No major control schemes are provided in this section of the plant. Flows are controlled by adjusting variable speed drives or turning motors on or off manually. Since the vapor compressor (R-101) is fixed speed, positive displacement, a pump-around loop controlled on inlet pressure has been added to avoid pulling a vacuum in the dryer. Dryer temperature controls are discussed with the steam plant (Section 1000).



5.5.2 Section 200 Gasification

The flow of wood into the gasifier is controlled by the level of wood in the top of the gasifier. The steady state set point for wood flow is determined from the control room. Therefore, the cycle rate for the lock hoppers G-201A&B is set by the operator. The metering bins' level detectors inform the operator that the lock hopper cycle rate must be adjusted.

If wood flow to the lock hoppers exceeds the capacity of any individual hopper, the condition is detected by a level transmitter that diverts wood flow to the companion hopper. If the adjacent hopper cannot accept wood, the flow of wood to both lock hoppers is shut off by stopping either bucket elevator G-205A or B. The diverter/flop gate feeding G-205A and B is also repositioned, if necessary. If both elevators are stopped, conveyor C-201 and therefore the flow of solids out of the storage silo (G-114) are also stopped.

The temperatures in the gasifier are controlled by the flow of oxygen to the gasifier. Thermocouples at all levels of the gasifier are used to monitor gasifier temperatures. Sufficient combustion must occur in the flaming pyrolysis zone to provide heat for reacting the char. Temperature set points will be determined by gas analysis (downstream of the gasifier) and monitoring char production. The flow of oxygen is also used to maintain gasifier operating pressure. The oxygen is available at a fixed pressure of 200 psig. Downstream (of the gasifier) pressure is set at 150 psig. To control both oxygen flow and pressure over a wide range of flows, two control valves in series are used.

A low pressure signal from the downstream pressure transmitter will decrease oxygen pressure drop through the two valves and vice versa. A pressure control valve and relief to flare are provided before T-201 to facilitate switching from low flow and/or pressure startup or shutdown conditions to normal operation.



Low pressure air instead of oxygen is used for startup and shutdown. A positive displacement blower (R-201) is used to provide air.

No temperature control is provided for T-201; this boiler operates by natural convection.

5.5.3 Section 300 Gas Shift

The flows of raw gas to the shift reactor (L-301) and shift bypass are controlled through a flow control valve on the bypass line. The flow rates are measured and the ratio of the flows is kept at approximately 1.5:1 (shift-to-bypass) with the control valve. Injection of water is used to control the temperature of the raw gas into the shift reactor. The flow of water is controlled by the temperature difference between the raw gas out of T-201 and the gas temperature after water injection, using the temperature into the shift as a set point and flow of raw gas to the shift reactor as a ramp proportioner. The temperature and composition out of the shift reactor are also monitored to alert the operator to changes in inlet gas composition (the temperature rise in the shift reactor is limited by inlet gas composition and the approach to shift equilibrium allowed with respect to catalyst activity) that would require changing the set point inlet temperature to L-301 to maintain the proper outlet H_2/CO ratio. Periodic raw gas analysis is also used to adjust the inlet temperature to the shift. A minimum water injection rate based on gas flow is used to avoid reducing the shift catalyst. If this causes inlet shift temperature to drop below the set point, the operator should adjust the shift bypass flow.

The gas cooling and production of steam in T-301 and T-302 are not controlled directly, but are limited by available surface area and temperature difference. The air cooler, T-303, is used to control the temperature of gas to the scrubber (G-401). The dampers' position on the air cooler fan is determined by the temperature difference between the inlet and outlet temperatures using the outlet temperature and air dry bulb temperature as set points.



5.5.4 Section 400 Gas Cooling and Compression

Shifted gas enters the section at a constant temperature of 140°F and is scrubbed and cooled in G-401 to a set temperature of 100°F by controlling the scrubber water temperature out of T-401. The flow of cooling water into T-401 is determined by the temperature difference between the scrubber water temperature into and out of T-401, with gas exit temperature variation from 100°F as a ramp variable. Scrubber water flow is constant.

Net liquid from M-401 is removed when the liquid level in the drum reaches the maximum allowable level. Light oils and heavy oils are removed from M-402 by the same method. Solids are backwashed from G-402 when the pressure drop across the filter reaches a maximum allowable level.

Depending on the vendor, capacity control and pressure control of the synthesis gas compressor stages may be integral with the compressor. A simple external control scheme is shown. Since the compressor uses positive displacement, too little inlet flow will create suction (reduced pressure) at the inlet. Inlet flow less than the maximum continuous rate for each stage is boosted by a recycle loop from discharge to suction. Flow through this loop is set by comparing inlet and discharge pressures with the design pressures. With the reciprocating compressor running at constant speed at maximum continuous output, excess inlet flow will increase upstream pressure above the Gasifier Section (200) set point and should be corrected at the gasifier. The temperature out of T-403 is set by controlling the damper position.

5.5.5 Section 500 Acid Gas Removal

Control of the Benfield System is the proprietary information of the Union Carbide Corporation, Engineering Products and Processes Division. Control of steam to the acid gas stripper reboiler (T-504A&B) will be by temperature control. With a regulated exit pressure from T-504, the exit temperature will be used to control flow. Temperatures below the dew point call for more steam. The flow of recompressed moisture from the wood dryer (G-110)



is constant. Steam flow is adjusted by changing the amount of 30 psig steam added from a set point of zero flow. Temperature could alternatively be controlled off of the return temperature of potassium carbonate solution to stripper A-502.

If all the recompressed moisture is not condensed, it is flashed/vented from steam drum M-503. The pressure in M-503 is regulated by let-down valves. The condensate from M-503 is cooled to a suitable discharge temperature with cooling water in T-506. The outlet temperature of condensate from T-506 is used to control cooling water flow from a set point of zero flow.

The acid gas from the Benfield System is cooled from 140°F to 100°F in T-505 to reduce the load on compressor R-1103. The flow of cooling water to T-505 is controlled by the outlet gas temperature from T-505.

5.5.6 Section 600 Methanol Synthesis

The exact control scheme for the methanol synthesis loop is the proprietary information of the Lurgi Kohle and Mineralöltechnik GmbH, Frankfurt, West Germany. Temperature in the methanol reactor is controlled by circulating water which is boiled to generate saturated steam. Reactor temperature is controlled by setting steam pressure. Liquid level in the steam drum (M-601) is maintained by controlling boiler feedwater flow via a drum level transmitter.

The temperature of synthesis gas into the guard adsorber (L-601A&B) is controlled by the heat release in fired heater F-602.

5.5.7 Section 700 Methanol Purification

The pressure in flash separator M-701 is controlled by a pressure regulator. Excess pressure is vented via a relief valve into blowdown drum M-704. Off-specification methanol produced during startup can be diverted by the operator to intermediate tank Q-701. Normally, the level in M-701 is used to control raw liquid methanol flow into the stabilizer (A-701). Flow



through the control valve is kept constant as long as high and low liquid levels are not breached.

The temperature of the feed into A-701 is controlled by the flow of steam condensate through T-701, since inlet temperature to T-501 is essentially fixed by the Section 600 operating conditions and the M-701 pressure.

The pressure in A-701 is controlled by the vent pressure on reflux drum M-702 which is set by a controllable regulator. Total reflux from M-702 is used; a level controller in M-702 prevents exceeding the maximum liquid level in the drum if the column flow rates are increased. The temperature out of air cooler T-703 is controlled by the damper setting on T-703, but feedback from the pressure controller on M-702 is used to assure enough air flow across T-703 to totally condense the condensible species. High levels of noncondensable gases (determined by gas analysis) will require lowering the stabilizer operating pressure or increasing the stabilizer reflux ratio.

The flow of steam to the stabilizer reboiler is controlled by the column bottom temperature; decreasing temperature (due to increased column flow) calls for more steam flow and vice versa. The quantity of steam is also monitored by the steam condensate temperature. More steam is required if the condensate temperature is below the saturation temperature at the regulated pressure. Ideally, the steam flow set point should correspond to zero subcooled condensate to avoid uncondensed steam. Excess steam is vented, indicating to the operator that supplemental steam production should be decreased. (See Section 1000.)

Flow of crude methanol from A-701 is controlled by level control; flow through the control valve is constant as long as high and low liquid levels are not violated. The temperature of the feed into the methanol still (A-702) is controlled by the flow of steam condensate through T-704, since the temperature out of A-701 is held constant at steady state operation.

The pressure at the top of A-702 is controlled by maintaining total condensation at a set point temperature. This in turn is monitored by observ-

ing reflux drum pressure and adjusting the cooling air flow accordingly. A safety relief vent for upset conditions and removing trace noncondensibles is provided. The flow of product to storage from the reflux drum (M-703) is determined by level control, since total reflux of methanol is used.

The flow of steam to the A-702 column reboiler is controlled by the column bottom temperature (and monitored with the steam condensate temperature) in the same manner that steam flow for A-701 is controlled. Water is removed from the bottom of A-702 by monitoring liquid level in the sump.

The temperature of the methanol product to storage is controlled by the flow of cooling water into T-707.

5.5.8 Section 800 Methanol Purge Gas Treatment

Since the temperature of water out of A-702 is controlled, the temperature of water into A-801 can be controlled using the outlet temperature from T-801 to set cooling water flow into T-801. The pressures of water and purge gas into A-801 are regulated. Control of the PSA System (G-801) is provided by the vendor, Union Carbide.

5.5.9 Section 900 Wastewater Treatment

The flow of isopropyl ether (IPE) for extraction in tower A-901 is controlled by monitoring the flow of wastewater to the column. Flow of water out of A-901 is controlled by the level of water in the sump of the tower.

The pressure at the top of the phenol still is set by the vent pressure on separator M-901. The temperature set point for still condenser T-902 is the corresponding IPE dew point. This temperature is maintained by controlling the flow of cooling water to T-902. A pair of manually set proportioning valves set reflux flow to A-902. Sufficient holdup is provided in M-901 to prevent overflowing the drum. Liquid level in M-901 is monitored to assure an adequate supply of IPE.



The flow of steam to A-902 is controlled in the same manner as flow of steam to A-701 and A-702 - off column temperature with steam condensate temperature as a monitor of the set point. The flow of phenol from A-902 is controlled by the level of phenol in A-902's sump.

5.5.10 Section 1000 Process and Boiler Water Treatment and Steam Plant

The controls of the activated carbon filter (G-1001), mixed bed demineralizer (G-1002), and chemical injection skids (G-1003, G-1004, and G-1005) are provided by their respective vendors. The level of demineralized water in Q-1001 is used to control the input of raw water to the treatment system.

Synthesis gas cooling (in Section 400) is used to heat demineralized water to the deaerator. To avoid overheating the water, a bypass around T-402 is controlled by the mixed (bypass plus T-402 effluent) stream outlet temperature. A similar control scheme is used to preheat process water for the shift to a constant 240°F. The deaerator temperature is set by the pressure maintained on the deaerator vent. The condensate return is supplemented by additional steam to maintain this temperature and produce the necessary steam flow to remove noncondensibles. Exact deaerator controls are provided by the vendor, supplemental steam and cooling water are provided.

The flow of makeup boiler feedwater (BFW) to each of the three steam drums is controlled by the level of BFW in the respective drum. The rate of steam production from each drum depends on the heat flux into their companion exchangers; the drum pressure is regulated. No flow controls are placed on the flow of high pressure steam from M-601 into steam turbine R-102. The pressure downstream of R-102 is controlled at the steam separator M-1002.

The flow of direct 150 psig steam to the dryer is controlled by the outlet temperature on the dryer exhaust. The flow of indirect 150 psig steam to the dryer is controlled by monitoring the vapor flow from condensate flash drum M-1003 and the temperature of steam and condensate from G-110. If the flow of 150 psig steam to G-110 is not adequate to dry the wood, the temperature will drop below the 150 psig dew point. If the flow of steam

out of M-1003 exceeds the level expected from flashing only 150 psig condensate, an excessive amount of 150 psig steam is being introduced to the dryer. These two indicators control the flow of balancing 150 psig from the FBC steam drum (M-1101) to the dryer.

The control of the flow of 30 psig steam to exchangers T-702, T-705, T-903, and T-504 has already been described. To assure sufficient 30 psig steam, 150 psig steam from the FBC steam drum (M-1101) is let down. The rate at which the 150 psig steam is supplied at (let down to) 30 psig is equal to the rates at which 30 psig steam flow is changed in the above four exchangers. If 30 psig steam demand is not being met, the pressure in the steam line from M-1101 will drop, indicating that the firing rate of FBC should be increased. If excess 30 psig steam is being introduced to the header, it is detected by the rate of vapor flow through the vapor/condensate drum M-1004 downstream of the exchangers. If high (excess) vapor flow occurs, the amount of 150 psig steam let down to 30 psig is reduced. However, vapor flow may still occur when zero 30 psig steam is being provided by the FBC, since there are other sources of 30 psig steam which cannot be reduced, i.e., steam flow to the dryer and from M-601 via M-1002. This excess 30 psig steam is vented through M-1004.

Since closing the let-down valve may result in excess steam being produced, the pressure on the 150 psig supply line from the FBC boiler is tied into the G-1111 metering bin (fuel feed rate) for the FBC.

The control of the temperature of raw methanol out of T-701 and T-704 has already been described. To avoid overheating the methanol with steam condensate, a bypass control is integrated with the temperature control of T-701 and T-704.

Liquid levels in steam separator/flash drums M-1002, M-1003, and M-1004 are maintained by level transmitters operating flow control valves. Liquid is needed in these drums to avoid gas bypassing into the condensate returns.



5.5.11 Section 1100 General Utilities and Waste Disposal

The temperature of acid gas to the sponge iron beds (L-1101 and 1102) is controlled by cooling water. Indication of bed breakthrough of H₂S is monitored with an automated detector.

Controls for skids G-1106, G-1107, and G-1108 are provided by the vendor. Controls for the FBC and the diesel generators are also vendor supplied.

Removal of oily wastes from M-1103 is controlled by monitoring sump levels.

Pressurized vents are relieved to 15 psig and maintained by pressure regulation at 15 psig up to a point adjacent to the flare.

Cooling tower water temperature is controlled by the damper setting on the tower fan. Makeup water for cooling is controlled by the water level in the cooling tower sump.

5.6 SPARING PHILOSOPHY

The plant has been designed to operate at a 90 percent stream factor. The equipment sparing and duplication to meet this goal are described below. Because a single gasifier train was part of the plant design basis, the gasifier has not been spared. The lock hopper system has two trains each capable of operating at 85 percent of design flow. Two gasifier feed points are required for proper feed distribution. The synthesis gas compressor and dryer vapor recompressor are not spared because of cost. Critical (lead time) spare parts should be maintained on site or be guaranteed available from the vendor. The two exchangers (T-504 and T-506) that handle dryer vapor are spared to allow on-line cleaning of the exchangers. Fouling from wood "blue haze" is expected. The guard adsorber (L-601) is spared to avoid plant shutdown while replacing the bed material; because gasifier sulfur products are not well documented, two identical beds are provided as a safety margin. The main sponge iron bed (L-1101) is spared with a smaller unit to be able to change the main bed while the plant is on stream. Two



diesel engines are provided to allow utilization of the PSA tail gas and vents over a wider range of operating rates and to allow safe operation of the plant if one engine is not operating. The engines are of different size strictly because the power specification fell within the two-size ranges, and two of the larger capacity were not needed. All pumps have been spared to avoid loss of service; all pumping requirements are provided by a single operating pump.

5.7 PLANT LAYOUT AND SKID LAYOUTS

The biomass-based methanol plant (BMP) is divided into eleven functional sections in which utilization of shop fabrication and skid mounting of equipment has been stressed to reduce field installation costs. The normal maximum allowable size of any skid was set by standard rail shipment. This allows an envelope of 12 ft wide by 12 ft high by 80 ft long. Using specialized rail cars can increase the height and length allowances.

The overall layout of the BMP is shown in Figure 5-11. A more detailed layout of the process area that identifies individual skid locations is shown as Figure 5-12. The layout of the BMP provides the required access to equipment for maintenance and operation. The process areas have been placed in close proximity in order to minimize piping run lengths. The location of major piping in the process area is shown in Figure 5-13. The skid layouts identifying individual pieces of process equipment are shown in Figures 5-14 to 5-23. Figure 5-24 is the layout of the waste treatment building.

Section 100 - "Wood Handling and Preparation" has not been designed for skid mounting. This is due to the layout and elevations required for the conveyors, screens, and other equipment that make field installation more practical. Figures 5-25 and 5-26 give the elevation views of the wood handling and preparation section facing south and east, respectively. In the remaining plant sections, most of the equipment is skid mounted except for the towers which are free standing. The skids include all piping, insulation, electrical, painting, and instrumentation that can be completed in the shop. This reduces the amount of more expensive field installation. Connections between skids are field installed.



There are a total of 46 individual skids of which 13 are vendor-supplied packages. Seven are air-cooled heat exchangers which are placed on the tops of other skids. The total number of skids is large, but many skids are of moderate size. Fewer skids could be used if the allowable individual skid size was increased. However, field installation work would not be significantly reduced. The overall skid list and skid dimensions are given in Table 5-18.

5.8 MATERIALS OF CONSTRUCTION

General

The use of carbon steel has been limited to 750°F maximum; 1 Cr-½ Mo steel is used up to 1050°F. Refractory linings on carbon steel have been used above 1050°F. Lines exposed to oily or acidic condensates are 18 Cr-8 Ni type 304 stainless steel or lined carbon steel.

Section 100 - Wood Handling

Vendor supplied equipment for this area is carbon steel or iron. Ultra-high-molecular-weight polyethylene is used as a lining material on chutes, hoppers, etc. to assure good flow of solids.

Section 200 - Gasification

Wood handling equipment is carbon steel. The gasifier and downstream equipment are refractory-lined carbon steel.

Section 300 - Gas Shift

All equipment downstream of water injection is carbon steel. The shift bypass only is 1 Cr-½ Mo.

Section 400 - Gas Cooling and Compression

Equipment in contact with the scrubbing liquor and condensate is 304 stainless steel.

Section 500 - Acid Gas Removal

Benfield equipment is carbon steel with the exception of T -504A&B which are 304 stainless steel. Condensate coolers T-506A&B and condensate flash drum M-503 are also made of 304 stainless. Equipment in contact with wet carbon dioxide may be lined carbon steel or stainless.

Section 600 - Methanol Synthesis

Gas service into and out of T-602 and L-601 is 1 Cr- $\frac{1}{2}$ Mo. Lurgi material specifications are not available. Fired heater tubes are 1 Cr- $\frac{1}{2}$ Mo.

Section 700 - Methanol Purification and Storage

All equipment is carbon steel.

Section 800 - Purge Gas Treatment

All equipment is carbon steel.

Section 900 - Wastewater Treatment

Equipment in contact with water condensate, phenol or iso-propyl ether is 304 stainless steel.

Section 1000 - Steam Plant and Process Water Treatment

All equipment is carbon steel except for hydrochloric acid and caustic transfer equipment.



Section 1100 - General Utilities, Waste Disposal and Offsites

Except for reagent handling equipment and (equipment) vendor-specified exceptions, all equipment is carbon steel.

5.9 COMMERCIAL PROJECT SCHEDULE

Figures 5-27 and 5-28 represent overall project summary schedules for construction of the first and "nth" biomass-based methanol plants. The schedule for the first plant is presented for comparison purposes. Project duration from application for environmental review and construction permits through startup is approximately 24 months for the first plant and 18 months for the "nth" plant.

The functions that have the most significant impact on the project duration are the critical elements that determine the project schedule. These are the permitting procedure, the detailed engineering and procurement program, equipment delivery, and plant startup.

The overall reduction in construction time from the first plant to the "nth" plant is six months. Two months of the six month reduction may be obtained via an abbreviated permitting procedure. A three month reduction is accomplished during the project execution phase and is attributable primarily to improvements in detailed engineering and equipment procurement and delivery. The final one month improvement in project schedule is due to increased operating experience which leads to smoother plant startups.

The permitting procedure, which is described in greater detail in Section 4.5, is assumed to include the requirement of an Environmental Impact Statement (EIS) for the first plant. Although detailed engineering and procurement of materials for long delivery lead time items (with provision for cancellation charges) can begin during the permitting phase, no equipment is released for fabrication until permits are issued. The time frame for the permitting procedures for the "nth" plant is two months shorter than that for the first plant due to the assumption that sufficient



operating information will be available from the plants which are already in operation to preclude the need for an EIS. These assumptions can only be verified in practice, as site-specific considerations, extent of involvement of county and municipal governments, and perception of local need will all play a role in determining the extent of the required permitting procedures.

The next critical element is the detailed engineering and procurement program. As mentioned earlier, the procurement of components and materials for long delivery items should begin during the permitting phase. The longest lead time piece of equipment for this project appears to be the Synthesis Gas Compressor (R-401 and R-402), which has a current (May 1984) delivery time of 30 weeks after approval of vendor drawings. The detailed engineering and procurement program for the "nth" plant is shorter than that for the first plant because a substantial portion of the engineering is repeat work, and qualified bidders who are familiar with the equipment specifications will have been established on previous projects.

Equipment delivery time is shorter for the "nth" plant than it is for the first plant, because the procurement program has benefited from the learning curve of the previous projects. Vendors are more familiar with the procurement requirements and can reduce interruptions in the equipment fabrication schedules. The learning curve efficiency improvement also occurs in skid construction and field assembly of the skids.

Plant startup of the "nth" plant is shorter than for the first plant, as the experience gained in previous startups and in operating plants can be used to anticipate startup problems and minimize delays.



5.10 EQUIPMENT LISTS AND DESCRIPTIONS

5.10.1 Buildings and Sheds List

<u>Name</u>	<u>Overall Plot Dimension</u>
Administrative Office/Lab/ Control Building	36 ft x 80 ft
Maintenance Shops/Locker Room	48 ft x 60 ft
Water Treatment Building	42 ft x 70 ft
Fire Pump Building	12 ft x 16 ft
Synthesis Gas Compressor Shed	20 ft x 20 ft
Diesel Engine/Generator Sheds	30 ft x 12 ft and 24 ft x 12 ft
Nitrogen Compressor Shed	9 ft x 12 ft
Cooling Tower Chemical Feed Shed	12 ft x 12 ft

5.10.2 Major Lines List
(3 Inch Diameter or Greater)

Gas Service

<u>Size (In.)</u>	<u>From Account No.</u>	<u>To Account No.</u>	<u>Approx. Length (ft)</u>	<u>Pressure (psig)</u>
14	G-110	R-101	30	2
12	Section 200	Flare	460	15
12	Section 300	Flare		
12	Section 400	Flare		
12	Section 500	Flare		
12	T-602	Flare	400	15
12	M-602	Flare		
12	M-704	Flare	360	2
12	Q-702A-D	Flare		
6	Section 800	Flare		
12	A-502	M-502	100	2
8	L-201	T-201	20	150
8	T-201	L-301	15	150
8	L-301	G-401	60	150
8	M-502	R-1103	70	2
8	R-101	T-504	290	25
6	R-1103	L-1101	30	15
6	M-401	R-401	15	150
6	M-502	G1103&1104	260	2
4	R-401	A-501	50	360
4	A-501	M-501	60	360
4	M-501	R-402	40	360
4	R-402	F-601/T-601	150	1005
4	F-601	T-602	35	1005

4	T-602	F-602	25	1005
4	F-602	L-601	45	1005
4	L-601	T-602	10	1005
4	T-602	T-603	15	1005
3	G-801	G-1103&1104	200	2

Liquid Service

<u>Size (In.)</u>	<u>From Account No.</u>	<u>To Account No.</u>	<u>Approx. Length (ft)</u>	<u>Pressure (psig)</u>
4	Q-702A-D	Tankers	200	100
3	G-401 Scrubber Loop		40	133
3	M-503	M-1103	320	25
3	T-504	M-503	20	25

Cooling Water, Steam, and Condensate Return

<u>Size (In.)</u>	<u>From Account No.</u>	<u>To Account No.</u>	<u>Approx. Length (ft)</u>	<u>Pressure (psig)</u>
6	Cooling water main and return		1100	100
3	T-506 & T-504 C.W. loops		40	100
6	R-102	M-1002	20	30
6	M-1002	Section 700	240	30
5	M-301	G-110	105	150
4	M-1102	Section 700	200	30
4	M-1101	G-110	80	150
3	M-1004	Section 700	50	30
3	Section 700	M-1001	70	30



5.10.3 Equipment Descriptions

Section 100 - Wood Handling

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
C-101	Scalper Feed Conveyor	24 in. wide trough belt conveyor, 83 ft long. 16 tph capacity, inclined 25° with walkway, drive, belt, cover, and supports.
C-102	Dryer Feed Conveyor	24 in. wide trough belt conveyor, 72 ft long. 16 tph capacity, inclined 25° with walkway, drive, belt, cover, and supports.
C-103	Fines Conveyor	24 in. belt conveyor, 103 ft long. 16 tph capacity, inclined 25° with walkway, drive, belt, cover, and supports.
C-104	Stored Fines Conveyor	18 in. trough belt conveyor, 20 ft long, inclined with drive, belt, cover, and supports.
C-105	Dried Fines Conveyor	18 in. belt conveyor, 20 ft long, 1 tph with walkway, drive, belt, cover and supports.
G-101	Truck Scale	50 ton capacity, pitless with steel deck, 10 ft x 35 ft, printer and indicator.
G-102	Tractor Shovel	Clark 55C or equivalent wheeled tractor with 5 cu yd woodchip bucket.



Section 100 - Wood Handling (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
G-103	Reclaim Hopper and Chain Conveyor	4 ft x 12 ft woodchip reclaim hopper with two strand, mill chain conveyor approximately 20 ft long.
G-104	Magnetic Separator	Eriez Model SE 7323 or equivalent suspended, self-cleaning electro-magnet including rectifier.
G-105	Scalping Screen	Rader RDS-25 scalping disc screen or equivalent with 25 mm IFO (interface opening, i.e., spacing between adjacent discs) and discharge hopper.
G-106	Primary Screen	Rader RVSP primary "V" disc screen, size 60 x 10 or equivalent with 10 mm IFO with intake and discharge chutes and discharge hopper.
G-107	Hog Mill and Blower	Montgomery Model 18RC40 Blo - Hog or equivalent hammermill with flywheel integral fan, (rolled plate) screen, discharge piping, and shear pin protection.
G-108	Fines Screen	Rader RVSF Fines "V" disc screen or equivalent with 5 mm IFO, discharge and intake chutes and bypass chute.
G-109	Dryer Screw Feeder	Acrison Model 140 V or equivalent with inlet hopper and drive, 1600 cfh.



Section 100 - Wood Handling (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
G-110	Wood Dryer	Bepex Model 96-22 Torus Disc ^(R) dryer or equivalent.
G-111	Dryer Fines Deck Screen	Rotex Model 221 or equivalent, single deck size 40 in. x 56 in. vibrating screen.
G-112	Fines Storage Silo and Reclaim	25 ton (3000 ft ³) capacity with "V" reclaimer, variable speed.
G-113	Storage Bucket Elevator	Bucket elevator with spaced 12 in. x 7 in. buckets, 65 ft centerline to centerline height. 10 tph capacity belt drive.
G-114	Dried Chips Storage Silo and Reclaimer	150 ton (18,000 ft ³) capacity silo with rotating screw reclaimer.
G-115	Gravimetric Feeder	Thayer Scale Model M or equivalent, 24 in. belt. 1-10 tph capacity range, with drive.
G-116	Rail Car Weigh Scale	Optional
G-117	Fines Discharge Bucket Elevator	Bucket elevator with spaced 10 in. x 6 in. buckets, 36 ft centerline to centerline height, 7 tph capacity, belt drive.
G-118	Dumpster	Approximately 800 ft ³ , supplied by hauler.
G-119	Rotary Vane Feeder	Included with G-110.



Section 100 - Wood Handling (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
R-101	Dryer Vapor Recompressor	Roots 1436LGL Spiraxial Gas Com- pressor or equivalent, 5160 cfm, 25 psi pressure difference.
R-102	Steam Turbine	Skinner model 518 or equivalent. Single stage, 6700 lb/hr with 600 psig allowable inlet pressure, 30 psig allowable backpressure.



Section 200 - Gasification

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
C-201	Conveyor	8 tph capacity, approximate length 32 ft, 24 in. trough belt with cover.
C-202A&B	Top Screw Feeders	Circular screw conveyors, 16 in. dia, approximately 18 ft long. Pressurized design for 160 psig.
C-203	Cyclone Char Removal Screw	Circular screw conveyor, 10 in. dia, approximately 10 ft long, pressurized design for 160 psig.
G-201A&B	Lock Hopper Feeders	Each consists of two lock hoppers, metering bin, metering screw, injector screw, and gas pressurization system with pressure recovery. 6 tph capacity, 160 psig delivery pressure.
G-202	Char and Ash Lock Hopper	See specification sheets. (Figure 4-6)
G-203	Lock Hopper Char and Ash Removal and Quench Screw	Through screw conveyor with 9 in. dia auger and spray nozzles, approximately 10 ft long.
G-204A&B	Pressurized Bucket Elevators	Bucket elevator with spaced 10 in. x 6 in. buckets, 35 ft centerline to centerline, 6 tph capacity each. Pressurized design for 160 psig.
G-205A&B	Bucket Elevators	Bucket elevator with spaced 10 in. x 6 in. buckets, 35 ft centerline to centerline, 6 tph capacity each.

Section 200 - Gasification (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
L-201	Gasifier	See specification sheets. (Figure 4-5)
M-201	Cyclone	1900 acfm design with 6 in. castable refractory lining.
R-201	Startup/Shutdown Blower	Roots RASJ Model 1021 or equivalent 3800 cfm, 4 psi pressure difference.
T201	Gas Cooler/Waste Heat Boiler	Vertical, refractory lined shell, U tube exchanger. Estimated duty 1.523 MM Btu/hr. Natural convection.



Section 300 - Gas Shift

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
F-301	Shift Startup Superheater	Propane-fired heater. Estimated duty 0.357 MM Btu/hr.
L-301	Shift Reactor	6 ft dia, 8 ft TT, vertical vessel.
M-301	Shift Cooling Steam Drum	3 ft-6 in. dia, 3 ft-6 in. TT length, horizontal arrangement.
T-301	Waste Heat Boiler	Vertical, refractory lined shell, U tube exchanger. Estimated duty 5.77 MM Btu/hr. Natural convection.
T-302	Waste Heat Economizer	TEMA Type U, horizontal. Estimated duty 0.875 MM Btu/hr
T-303	Shifted Gas Air Cooler	Single bay, six rows of tubes with one 15 hp motor. Estimated duty 3.63 MM Btu/hr.

Section 400 - Gas Cooling and Compression

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
G-401	Venturi Scrubber	800 acfm design with adjustable throat. 1 ft dia inlet, 6 ft-9 in. top of inlet to centerline of outlet.
G-402	Scrubber Filter	Duriron Model TE-2, Dual 4, 3 1/4 x 36 or equivalent.
M-401	Vapor Liquid Separator	4 ft dia, 8 ft TT vertical with tangential side inlet from venturi.
M-402	Oil/Water Separator	2 ft dia, 5 ft TT length API type separator.
P-401A&B	Scrubber Recycle Pump and Spare	75 gpm at 110 psi head.
R-401 and R-402	First and Second Stage Synthesis Gas Compressor	Penn Pump, 4 throw Model HOE or equivalent reciprocating compressor with pulsation dampeners.
T-401	Scrubber Water Cooler	TEMA Types L, M, or N, horizontal. Estimated duty 0.557 MM Btu/hr.
T-402	First-Stage Compressor Cooler	TEMA Type U, horizontal. Estimated duty 0.623 MM Btu/hr.
T-403	First-Stage Compressor Air Cooler	Single bay, five rows of tubes with one 5 hp motor. Estimated duty 0.90 MM Btu/hr.



Section 500 - Acid Gas Removal

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
A-501	Acid Gas Absorber	Packed column, dimensions confidential ^a
A-502	Acid Gas Stripper	Packed column, dimensions confidential ^a
G-501	Carbonate Mixer	Not available ^a
M-501	Clean Syngas Water Knockout Drum	2 ft dia, 5.5 ft TT length ^a
M-502	Acid Gas Water Knockout Drum	3.5 ft dia, 8 ft-10 in. TT length ^a
M-503	Steam Blowdown Drum	2.5 ft dia, 7.5 ft TT length
M-504	Acid Gas Final Water Knockout Drum	2 ft-4 in. dia, 6.0 ft TT length
P-501A&B	Solution Circulation Pump and Spare	450 gpm ^a
P-502A&B	Solution Makeup Pump and Spare	10 gpm ^a
P-503A&B	Carbonate Mix Tank Pump and Spare	Not available ^a
P-504A&B	Knockout Drum Recirculation Pump and Spare	2 gpm with 30 psi head

^a provided as part of a turnkey package by the Union Carbide Corporation

Section 500 - Acid Gas Removal (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
P-505A&B	Knockout Drum Recirculation Pump and Spare	25 gpm with 40 psi head
Q-501	Carbonate Mix Tank	Not available ^a
Q-502	Solution Makeup Tank	9,000 gallons ^a
T-501	Clean Syngas Air Cooler	Estimated duty 1.37 MM Btu/hr ^a
T-502	Clean Syngas Trim Cooler	Estimated duty 0.20 MM Btu/hr ^a
T-503	Acid Gas Air Cooler	Estimated duty 8.80 MM Btu/hr ^a
T-504A&B	Stripper Reboiler and Spare	Estimated duty 11.45 MM Btu/hr ^a
T-505	Acid Gas Final Cooler	TEMA Type L, M, or N, vertical. Estimated duty 0.70 MM Btu/hr
T-506A&B	Condensate Cooler and Spare	TEMA Type U, horizontal. Estimated duty 1.148 MM Btu/hr

^aprovided as part of a turnkey package by the Union Carbide Corporation



Section 600 - Methanol Synthesis

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
F-601	Guard Adsorber Startup Feed Heater	Propane-fired, estimated duty 0.480 MM Btu/hr
F-602	Guard Adsorber Final Heater	Propane-fired, estimated duty 0.275 MM Btu/hr
F-603	Reactor Startup Heater	Specifications confidential ^a
G-601	Steam Ejector	Specifications confidential ^a
L-601A&B	Guard Adsorber and Spare	4 ft-6 in. dia, 7 ft-5 in. TT verti- cal with 5 ft-5 in. packed depth.
L-602	Methanol Reactor	Dimensions confidential ^a
M-601	Methanol Reactor Steam Drum	Dimensions confidential ^a
M-602	Raw Product Separator	Dimensions confidential ^a
R-601	Recycle Compressor	Specifications confidential ^a
T-601	Guard Adsorber Feed Heater	Estimated duty, 0.480 MM Btu/hr ^a
T-602	Guard Adsorber Feed/ Effluent Exchanger	TEMA Type U, horizontal. Estimated duty 1.411 MM Btu/hr

^aprovided as part of a turnkey package by the Lurgi Kohle und Mineralöltechnik GmbH

Section 600 - Methanol Synthesis (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
T-603	Methanol Feed/Effluent Exchanger	Specifications confidential ^a
T-604	Raw Product Air Cooler	Specifications confidential ^a
T-605	Raw Product Trim Cooler	Specifications confidential ^a

^aprovided as part of a turnkey package by the Lurgi Kohle und Mineralöltechnik GmbH



Section 700 - Methanol Purification

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
A-701	Stabilizer	2 ft dia, 18 ft-6 in. TT with 7 trays
A-702	Methanol Still	3 ft-6 in. dia, 59 ft-6 in. TT with 32 trays
G-701	Caustic Addition System	Confidential
M-701	Flash Separator	3 ft dia, 5 ft TT length, horizontal
M-702	Stabilizer Reflux Drum	1.5 ft dia, 3.5 ft TT length, vertical
M-703	Methanol Still Reflux Durm	4 ft dia, 5 ft-4 in. TT length, horizontal
M-704	Blowdown Drum	4.5 ft dia, 11 ft-3 in. TT length, horizontal
P-701A&B	Intermediate Tank Pump and Spare	4 gpm at 77 psi head
P-702A&B	Blowdown Drum Pump and Spare	4 gpm at 77 psi head
P-703A&B	Product Pump and Spare	20 gpm at 17 psi head
P-704A&B	Stabilizer Reflux Pump and Spare	30 gpm at 18 psi head
P-705A&B	Still Reflux Pump and Spare	50 gpm at 28 psi head



Section 700 - Methanol Purification (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
P-706A&B	Product Transfer Pumps	150 gpm at 20 psi head
Q-701	Intermediate Tank	12 ft dia, 10 ft high
Q-702A-D	Final Product Tanks	Each 12 ft dia, 45 ft TT length, horizontal
T-701	Stabilizer Feed Heater	Multi-tube double pipe, estimated duty 0.348 MM Btu/hr
T-702	Stabilizer Reboiler	Kettle reboiler with U tubes. Estimated duty 1.649 MM Btu/hr
T-703	Stabilizer Air Cooler	Single bay with 6 rows of tubes, 7.5 hp motor, estimated duty 1.52 MM Btu/hr
T-704	Methanol Still Feed Heater	TEMA Type U, estimated duty 0.210 MM Btu/hr
T-705	Methanol Still Reboiler	Kettle reboiler with U tubes. Estimated duty 7.54 MM Btu/hr
T-706	Methanol Product Air Cooler	Two bays each with 6 rows of tubes and a 15 hp motor. Estimated duty 7.54 MM Btu/hr
T-707	Methanol Product Final Cooler	TEMA Type U, estimated duty 0.261 MM Btu/hr



Section 800 - Purge Gas Treatment

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
A-801	Methanol Wash Tower	1 ft dia, 7 ft-6 in. TT, vertical with 3 ft of packing.
G-801	PSA Unit	Inlet flow 27,000 scfh; standard skid package provided by Union Carbide
P-801A&B	Washwater Pump and Spare	1 gpm at 385 psi head
T-801	Washwater Cooler	Multi-tube double pipe. Estimated duty 0.06 MM Btu/hr



Section 900 - Wastewater Treatment

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
A-901	Extraction Tower	1 ft dia, 8 ft TT, vertical with sieve trays
A-902	Phenol Still	1 ft dia, 9 ft TT vertical with additional 2 ft TT well at 6 in. dia, 6 ft of packing.
M-901	Condensate Separator	1.5 ft dia, 3.0 ft TT vertical
P-901A&B	Isopropyl Ether (IPE) Feed Pump and Spare	1 gpm at 15 psi head
P-902A&B	IPE Recycle Pump and Spare	1 gpm at 10 psi head
P-903A&B	Phenol Wastes Pump and Spare	1 gpm at 40 psi head
P-904A&B	Wastewater Pump and Spare	1 gpm at 55 psi head
Q-901	IPE Storage Tank	2.0 ft dia, 9.0 ft TT vertical
T-901	Phenol Still Preheater	TEMA Type L, M, or N, vertical. Estimated duty 0.010 MM Btu/hr
T-902	Phenol Still Condenser	Multi-tube double pipe, estimated duty 0.070 MM Btu/hr
T-903	Phenol Still Reboiler	Multi-tube double pipe, estimated duty 0.070 MM Btu/hr

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Section 1000 - Steam Plant and Process Water Treatment

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
G-1001	Activated Carbon Filter	Illinois Water Treatment Model No. ACS-366 or equivalent, 36 in. dia x 96 in. TT main vessel containing 34.5 ft ³ of activated carbon
G-1002	Mixed Bed Demineralizer	Illinois Water Treatment custom package deionizer Model No. AMB-3696 or equivalent. Service vessel is 36 in. dia x 96 in. TT with 12 ft ³ of cation resin and 18 ft ³ of anion resin. Includes acid and caustic ejectors and day tanks and associated piping and controls
G-1003	Ammonia Injection Skid	Includes feed tank, ammonia tank, mixer, feed pumps and controls
G-1004	Sodium Phosphate Injection Skid	Includes dissolving/mix tank, mixer, feed pumps and controls
G-1005	Sodium Sulfitite Injection Skid	Includes dissolving tank and mixer, feed pumps and controls
M-1001	Deaerator	Crane Co. (Cochrane Division) Jet Tray Deaerator, Model 6400 Uni-Pac T-30 or equivalent
M-1002	Steam Separator	3.5 ft dia, 6 ft-7 in. TT, vertical
M-1003	Condensate Flash Drum	3.5 ft dia, 7 ft-8 in. TT, vertical
M-1004	Condensate Flash Drum	3.5 ft dia, 7.0 ft TT, vertical

Section 1000 - Steam Plant and Process Water Treatment (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
Q-1001	Demineralized Water Tank	8000 gal, horizontal, fiberglass
Q-1002	Caustic Tank	10,000 gal, horizontal
Q-1003	Hydrochloric Acid Tank	10,000 gal, horizontal
P-1001A&B	Caustic Transfer Pump and Spare	5 gpm at 55 psi head
P-1002A&B	HCl Transfer Pump and Spare	5 gpm at 55 psi head
P-1003A&B	Process Water Pump and Spare	6 gpm at 200 psi head
P-1004A&B	Deaerator Pump and Spare	5 gpm at 10 psi head
P-1005A&B	150 psig BFW Pump and Spare	40 gpm at 190 psi head
P-1006A&B	30 psig BFW Pump and Spare	6 gpm at 40 psi head
P-1007A&B	565 psig BFW Pump and Spare	15 gpm at 615 psi head
T-1001	Water Heater	Attached to diesel engine water jacket, double pipe. Estimated duty 0.174 MM Btu/hr

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Section 1100 - General Utilities, Waste Disposal and Offsites

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
C-1101	Char and Ash Conveyor	18 in. belt conveyor x 30 ft long inclined with drive, belt, supports, and cover
G-1101	Fluidized Bed Combustor	Energy Products of Idaho 7.5 ft dia combustion cell or equivalent
G-1102	Flare	Maximum heat release 87.5 MM Btu/hr from 196 Btu/scf gas, 50 ft stack height with 50 ft exclusion radius
G-1103&4	Diesel Engines	Dresser Waukesha Enginator Models VHP9500G and VHP7100G or equivalent, rated at 950 kW, and 700 kW respectively, operating on natural gas at 1200 rpm, includes generators, air starters, and radiators
G-1105	FBC Stack	Provided by G-1101 vendor as per his design
G-1106	Demineralizer Backwash Regeneration Skid	Includes acid and caustic feed tanks, feed pumps and controls, details by vendor
G-1107	Sulfuric Acid Skid	Storage tank, feed pumps, and controls. Specifications by vendor.
G-1108	Sodium Hypochlorite Skid	Storage tank, feed pumps, and controls. Specifications by vendor.

Section 1100 - General Utilities, Waste Disposal and Offsites (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
G-1109	FBC Ash Dumpster	Approximately 1300 ft ³ , i.e., 30 ft rolloff
G-1110	Baghouse Ash Dumpster	Approximately 6 ft x 6 ft x 8 ft
G-1111	Metering Bin	Provided by G-1101 vendor, as per his design
G-1112	Char and Ash Storage Bin	5 ft dia hopper with 5 ft TT height plus 60° cone bottom and rotary feeder on outlet
G-1113	Diesel Stack	Carbon steel pipe and base
G-1114	Bucket Elevator	10 in. x 6 in. spaced buckets, with belt drive, 35 ft centerline to centerline of pulleys, 2 tph capacity
G-1115	Acid Gas Stack	Epoxy lined steel pipe and base
G-1116	FBC Baghouse	Provided by G-1101 vendor as per his design
G-1117	Fixed Foam System & Building	Includes 3000 gal foam tank
G-1118	Halon Fire Extinguishing System	Supplied by vendor
L-1101	Sponge Iron Bed	10 ft dia with 15 ft bed depth
L-1102	Sponge Iron Bed	3 ft dia with 10 ft bed depth

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Section 1100 - General Utilities, Waste Disposal and Offsites (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
M-1101	FBC Steam Drum	Part of Type A boiler, supplied by G-1101 vendor
M-1102	Diesel Waste Heat Boiler Steam Drum	3.5 ft dia by 3.5 ft TT, horizontal
M-1103	Oil/Water Separator	Horizontal API type separator, 3.5 ft dia, 7.5 ft TT
M-1104	Nitrogen Surge Drum	4.0 ft dia, 22.0 ft TT, horizontal
M-1105	Plant Instrument Nitrogen Receiver	1.5 ft dia, 4 ft TT, horizontal
M-1106	Plant Air Receiver	1.5 ft dia, 4 ft TT, horizontal
P-1101A&B	Liquid Wastes Pump and Spare	1/3 gpm at 60 psi head
P-1102A&B	Cooling Water Pump and Spare	375 gpm at 100 psi head
P-1103A&B	Equalization Pond Pump and Spare	40 gpm at 60 psi head
P-1104A&B	Demineralizer Sump Pump and Spare	80 gpm at 30 psi head
P-1105A&B	Oily Waste Pump and Spare	1/2 gpm at 100 psi head

Section 1100 - General Utilities, Waste Disposal and Offsites (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
P-1106A&B	Fire Water Pump and Spare	2000 gpm each, 150 psi head, one diesel drive, one electric drive, 300 hp motors
P-1107A&B	Jockey Pump and Spare	50 gpm at 25 psi head
P-1108A&B	Diesel Fuel Pump and Spare	80 gph at 40 psi head
Q-1101	Propane Storage Tank	5.0 ft dia, 12 ft-3 in. TT, horizontal
Q-1102	FBC Liquid Wastes Storage Drum	Connected 55 gal drums
Q-1103A&B	Runoff Equalization Ponds	Concrete lined basins, each 47,500 gal
Q-1104	Demineralizer Wastes Sump	5000 gal tank with eductors
Q-1105	Diesel Fuel Tank	Standard 270 gal tank
R-1101	FBC Baghouse I.D. Fan	Provided by G-1101 vendor as per his design
R-1102	Nitrogen Compressor	Ingersoll Rand Air Compressor Model 15T2 or equivalent, 41.2 cfm at 1000 psig
R-1103	Regeneration Blower	Roots RCS-J, Model 821 or equivalent, 1790 cfm with 15 psi differential

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Section 1100 - General Utilities, Waste Disposal and Offsites (cont'd)

<u>Account No.</u>	<u>Item Name</u>	<u>Description</u>
R-1104	Plant Instrument Nitrogen Compressor	400 scfm at 100 psi head
R-1105	Plant Air Compressor	400 scfm at 100 psi head
T-1101	FBC Waste Heat Economizer	Part of Type A boiler, supplied by G-1101 vendor
T-1102	FBC Boiler	Type A, supplied by G-1101 vendor
T-1103	Diesel Economizer	TEMA Type L or M, estimated duty 0.085 MM Btu/hr
T-1104	Diesel Waste Heater Boiler	TEMA Type U, estimated duty 2.328 MM Btu/hr
T-1105	Cooling Tower	Mechanical draft tower. Baltimore Aircoil Model VXT-400 or equivalent.
T-1106	Acid Gas Cooler	TEMA Type L, M, or N. Estimated duty 0.150 MM Btu/hr

TABLE 5-1

HIGH PRESSURE STEAM VALUE PER MILLION BTU'S
(550 psig, 750°F)

Electrical Generation and Condensation

Heat rate assumed: $15,400 \text{ Btu/kWh} = 65 \text{ kWh per MM Btu}$

Electricity cost: $4.2\text{¢/kWh} = \$2.73/\text{MM Btu fuel}$

Capital cost for steam turbine, generator
and condenser: $\$300/\text{kW}$

Annual capital recovery: 20 percent

Operating hours per year: 7000

Capital charges for electricity: $0.86\text{¢/kWh} = \$0.56/\text{MM Btu}$

Net steam value w/o cogeneration: $\$2.73 - \$0.56 = \$2.17/\text{MM Btu}$

Electrical Cogeneration With Process Steam

Heat rate extracting at 30 psig: $29,500 \text{ Btu/kWh} = 33.9 \text{ kW per MM Btu}$

Extracted steam value: $\$3.10/\text{MM Btu}$

Electricity value: $\$1.42/\text{MM Btu}$

Capital cost for steam turbine and generator: $\$150/\text{kW}$

Electricity capital charges: $0.43\text{¢/kWh} = \$0.15/\text{MM Btu}$

Net steam value if cogenerating: $\$3.10 + \$1.42 - \$0.15 = \$4.37/\text{MM Btu}$



TABLE 5-2

CHEMICAL ANALYSIS OF SPOKANE WATER

	<u>mg/l</u>
Calcium (Ca)	34.3
Magnesium (Mg)	17.4
Sodium (Na)	3.6
Carbonate (CO ₃)	0
Sulfate (SO ₄)	11.0
Chloride (Cl)	1.6
Silica (SiO ₂)	12.0
Alumina (Al ₂ O ₃)	0
Iron (Fe)	0
ABS	0.01
Alkalinity, phenolphthalein (as CaCO ₃)	0
Alkalinity, methyl orange (as CaCO ₃)	144
Hardness* (as CaCO ₃)	122
Total Solids	141
Mineral Matter	77
Volatile Material	64
Dissolved Oxygen (O ₂)	7.7
Fluorine (F)	trace
Nitrate Nitrogen (NO ₃ -N)**	0.8
Manganese (Mn)	0
pH	7.8

*Hardness as grains per U.S. gal--7.1

**As Nitrate--3.5 mg/l



TABLE 5-3

METHANOL PRODUCT SPECIFICATION

<u>Characteristic</u>	<u>Method</u>	<u>Specification</u> ^a
Methanol content		99.40 ^b
Water content	ASTM-D-1364	0.025 ^c
Residue of Evap (mg/100 mL)	ASTM-D-1353	1.0
Acid (as acetic acid)	ASTM-D-1613	0.002
Alkalinity (as ammonia)		0.00030
Chloride (max)	Spec. ion electrode	0.0001
Lead (g/gal)	ASTM-D-3116	0.01
Phosphorous (g/gal)	ASTM-D-3231	0.005
Sulfur	ASTM-D-2622	0.025
Iron, volatile (mg/L)	Photometry	0.02
Permanganate No. (min)	ASTM-D-1363	60 min
Acetone No. (max)	Chromatography	0.002
9-11 psi RVP Unleaded gasolines (seasonally) added for volatility (volume percent)	ASTM-D-439-79	9.0-11.0

^aweight percent except as noted

^bCalifornia Energy Commission specification is 99.85.

^bCalifornia Energy Commission specification is 0.05.



TABLE 5-4

SPECIFICATIONS FOR INDUSTRIAL GRADE METHANOL

<u>Property</u>	<u>Grade A</u>	<u>Grade AA</u>
Methanol content, wt percent, min	99.85	99.85
Acetone and aldehydes, ppm, max	30	30
Acetone, ppm, max		20
Ethanol, ppm, max		10
Acid (as acetic acid), ppm, max	30	30
Water content, ppm, max	1500	1000
Specific gravity, 20/20°C	0.7928	0.7928
Permanganate time, min	30	30
Odor	Characteristic	Characteristic
Distillation range at 101 kPa (760 mm Hg)	1°C, must include 64.6°C	1°C, must include 64.4°C
Color, platinum-cobalt scale, max	5	5
Appearance	Clear-colorless	Clear-colorless
Residual on evaporation, g/100 mL	0.001	0.001
Carbonizable impurities; color platinum-cobalt scale, max	30	30



TABLE 5-5

PLANT BY-PRODUCTS AND WASTES

<u>By-products</u>	<u>Disposition</u>
Undersized and oversized wood	Burned onsite for steam generation.
Carbon dioxide (acid gas)	Sent to sponge iron beds to oxidize sulfides, and subsequently exhausted.
<u>Waste Products</u>	
Char and ash from gasification	Burned onsite for steam generation.
Ash from incineration	Landfilled
Reagent purged from Benfield system	Drummed and landfilled (non-hazardous).
Blowdown and condensate from wood dryer	Sewered
Condensate from raw gas scrubbing	Sewered after treatment
Water from methanol synthesis and purification	Sewered
Tars from gasification	Extracted from condensate and incinerated

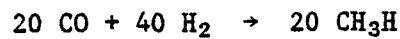


TABLE 5-6

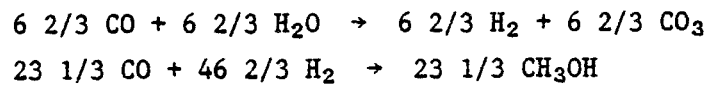
THE EFFECT OF CARBON DIOXIDE CONVERSION ON METHANOL YIELD

Initial Raw Gas: 40 moles H₂
 30 moles CO
 30 moles CO₂

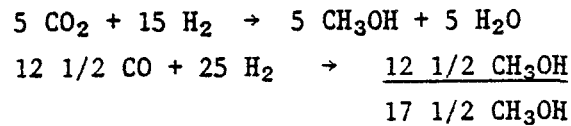
A) Maximum methanol without shifting:



B) Shifting to maximize methanol production:



C) If 5 moles CO₂ react with H₂:



NUMBER STREAM	COMPONENT PROPERTIES				1		2		3		4		5		6		7		8		9	
					Wood Feed to Gasifier		Oxygen Feed to Gasifier		Raw Gas Product		Gasifier Char + Ash		Water to Shift		Feed to Shift		Combined Shift Product		Net Scrubber Liquor		Scrubbed Raw Gas	
					LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR
COMPONENT	MOL WT	SG																				
CH ₃ OH	32.04																					
H ₂ O	18.016				128.3	2311			108.8	1961			135.0	2432	200.3	3609	145.9	2628	139.9	2520	6.0	108
CO	28.01							297.8	8061						172.7	4837	182.8	5120			182.8	5120
CO ₂	44.01							178.0	7834						106.8	4700	283.0	12,455			283.0	12,455
CH ₄	16.04							30.0	481						18.0	289	30.0	481			30.0	481
H ₂	2.016							332.3	670						199.4	402	437.2	881			437.2	881
N ₂	28.01					3.1	87	3.5	98						2.1	59	3.5	98			3.5	98
Lights	46.07																					
C ₂ H ₅ OH	46.07																					
C ₃₊ Alcohols	60.01																					
Dry, Ash-Free Wood	-				-	12,065																
Wood Ash	-				-	120					-	120										
Char	-										-	130										
Tars	-								-	6					-	4	-	6	-	6	-	-
O ₂	32.00					156.5	4848															
NH ₃	17.03							0.2	4						0.1	2	0.2	4	0.2	4	-	-
SO ₂ /H ₂ S	-							0.1	6						0.06	4	0.1	6			0.1	6
IPE	102.17																					
TOTALS					14,486	154.6	4935	940.7	19,121		250	135.0	2432	697.5	13,906	10827	21679	140.1	2530	942.6	19,149	
BPSD					WT/SD																	
GPH					M ³ /HR																	
°API					SP. GR.																	
LB/GAL					KG/M ³																	
MW																						
SCFH					m ³ /HR																	
REMARKS					5-83																	
SCF conditions - 60° F, 14.7 PSIA					m ³ conditions - 0° C 760 mm Hg abs																	

TABLE
5-7

COMPONENT MATERIAL BALANCE
 Client: SERT
 Plant: METHANOL PLANT
 STONE 6 WEBSTER ENGINEERING CORPORATION
 MW 291-3-70 Sheet 1 of 4

NUMBER STREAM	COMPONENT PROPERTIES			10		11		12		13		14		15		16		17		18		
				Clean Syngas to Synthesis		Total Flow to Synthesis		Acid Gas to Stack		Acid Gas to Diesels		Total Acid Gas		Isopropyl Ether Makeup		Treated Condensate		Phenols to Incineration		Raw Methanol Product		
				LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR
CH ₃ OH	32.04																			201.0	644.0	
H ₂ O	18.016			1.8	32	1.8	32	45.6	821	7.4	134	53.0	955			131.9	2520			24.7	445	
CO	28.01			182.7	5117	182.7	5117	0.1	~ 3		trace	0.1	3							0.2	6	
CO ₂	44.01			29.5	1298	29.5	1298	218.0	9515	35.5	1562	253.5	11,157							3.9	171	
CH ₄	16.04			29.6	475	29.6	475	0.3	5	0.06	1	0.4	6							3.7	59	
H ₂	2.016			436.6	880	464.5	936	0.5	~ 1	0.08	trace	0.6	~ 1							1.5	3	
N ₂	28.01			3.5	98	3.8	106	-	-	-	-	-	-							0.2	5	
Lights	46.07																			0.3	12	
C ₂ H ₅ OH	46.07																			0.04	2	
C ₃ + Alcohols	60.09																			0.03	2	
Dry, Ash-Free Wood	-																					
Wood Ash	-																					
Char	-																					
Tars	-															100ppm	trace			6		
O ₂	32.0															0.2	4					
NH ₃	17.03																					
SO ₂ /H ₂ S	-									trace		trace		trace								
IPE	162.17														5		5			trace		
TOTALS				683.7	7100	711.9	7164	264.5	10,425	~43.0	1697	307.6	12,122		5	141.1	2529			6	235.5	7145
DPSD				WT/SD																		
GPH				M ³ /HR																		
°API				SP. GR.																		
LB/CAL.				KG/M ³																		
MW																						
SCFH				CM ³ /HR																		

REMARKS	
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ISS.	DATE	DESCRIPTION OF ISSUE	CH'NO	APP'D
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COMPONENT MATERIAL BALANCE
 Client: SERI
 Plant: METHANOL PLANT
 STONE & WEBSTER ENGINEERING CORPORATION
 MB 201-3-70 Sheet 2 of 4

SCF conditions - 60° F, 14.7 PSIA CM³ conditions - 0° C 760 mm. Hg abs

TABLE 5-7 CONT.

NUMBER STREAM	COMPONENT PROPERTIES				19		20		21		22		23		24		25		26		27	
					Raw Methanol from Flash		Flash Overhead		Stabilizer Overhead		Stabilizer Bottoms		Methanol Product		Methanol Still Bottoms		Methanol Purge Gas		Methanol Dis- tillation Washw		PSA Tail Gas to Diesels	
					LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR
CH ₂ OH	32.04				200.8	6434	0.2	6	~0.5	15	200.3	6419	200.3	6419	-	-	1.0	31	1.0	31	-	-
H ₂ O	18.016				24.7	445	-	-	-	-	24.7	445	0.05	1	~24.6	444	-	-	24.5	441	0.2	3
CO	28.01				-	-	0.2	6	-	-	-	-	-	-	-	-	3.1	86			3.1	86
CO ₂	44.01				2.2	97	1.7	74	2.2	97	-	-	-	-	-	-	2.6	114			2.6	114
CH ₄	16.04				0.2	4	3.5	55	0.2	4	-	-	-	-	-	-	25.9	416			25.9	416
H ₂	2.016				-	-	1.5	3	-	-	-	-	-	-	-	-	35.2	71			7.9	15
N ₂	28.01				-	-	0.2	5	-	-	-	-	-	-	-	-	3.6	101			3.3	93
Lights	4.07				0.3	12	-	-	0.1	5	0.2	7	0.2	7	-	-	-	-				
C ₂ H ₅ OH	4.07				0.04	2	-	-	-	-	0.04	2	0.04	2	-	-	-	-				
C ₃₊ Alcohols	6.09				0.03	2	-	-	-	-	0.03	2	-	-	0.03	2	-	-	0.03	2		
TOTALS					~228.2	6996	7.3	149	3.0	121	225.2	6875	~200.6	6429	~24.7	446	71.4	819	~25.5	474	43.0	727
BPSD WT/50																						
OPH M ³ /HR																						
°API SP. GR.																						
LB/GAL. KG/M ³																						
MW																						
SCFH M ³ /HR																						

REMARKS	
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	SCF conditions - 60° F, 14.7 PSIA m ³ conditions - 0° C 760 mm. Hgabs

ISS	DATE	DESCRIPTION OF ISSUE	CHK'D	APP'D
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TABLE 5-7 CONT.		COMPONENT MATERIAL BALANCE Client: SERI Plant: METHANOL PLANT STONE & WEBSTER ENGINEERING CORPORATION	
		MB 201-3-70 Sheet 3 of 4	

NUMBER	COMPONENT PROPERTIES			28		29		30											
				Total Fuel Gas to Diesels		Hydrogen Recycle		Dried Acid Gas											
STREAM	MOL. WT.	SG		LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR	LB MOL/HR	LB/HR
COMPONENT				SCF-MOL/HR	SCF/HR	SCF-MOL/HR	SCF/HR	SCF-MOL/HR	SCF/HR	SCF-MOL/HR	SCF/HR	SCF-MOL/HR	SCF/HR	SCF-MOL/HR	SCF/HR	SCF-MOL/HR	SCF/HR	SCF-MOL/HR	SCF/HR
CH ₃ OH	32.04			0.7	21														
H ₂ O	18.016			8.8	137			13.2	238										
CO	28.01			3.3	92			0.1	~ 3										
CO ₂	44.01			42.0	1847			218.0	9595										
CH ₄	16.04			27.6	476			0.3	5										
H ₂	2.016			8.9	18	27.9	56	0.5	1										
N ₂	28.01			3.5	98	0.3	8												
Lights	46.07			0.1	5														
C ₂ H ₅ OH	46.07																		
C ₃ + Alcohols	60.09																		
H ₂ S	34.02								trace										
TOTALS				96.9	2694	28.2	64	232.1	9842										
BPSD				MT/SD															
GPH				M ³ /HR															
°API				SP. GR.															
LB/GAL.				KG/M ³															
				MW															
SCFH				m ³ /HR															

REMARKS

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SCF conditions - 60° F, 14.7 PSIA m³ conditions - 0° C 760 mm. Hgabs

TABLE
5-7
CONT.

COMPONENT MATERIAL BALANCE
Client: SERI
Plant: METHANOL PLANT
STONE & WEBSTER ENGINEERING CORPORATION

MB 201-3-70 Sheet 4 of 4

ISS. DATE DESCRIPTION OF ISSUE CH'KD APP'D

TABLE 5-8

Base Design Energy Sources and Uses
 (Approximate Values in Parenthesis in MM Btu/Hr)

Definite Sources

Methanol Purge Gas (13.0 - LHV)
 Shift Cooling (6.64)
 Methanol Reactor Steam (6.66)
 Gasifier (Raw Gas) Product Cooling (3.6)
 Wood Fines and Char (9.5-HHV)

Definite Users

Wood Drying (11.7-14.5 net)
 Shift Steam (2.1)
 Acid Gas Steam (11.45)
 Methanol Distillation (9.06)
 Plant Electric Motors
 Guard Bed Heater (0.275)

Possible Sources

Evaporated Wood Moisture (9.0)
 Green Woodchips
 Electric Power
 Fossil Fuels

Possible Users

Electricity/Steam Export to Oxygen
 Plant
 Wastewater Stripping Steam
 Steam Turbine Drives (backing out
 large motors)



TABLE 5-9

Estimated Economics for Indirect Drying and Vapor Compression
(Versus Direct Air Drying)

Reduction in dryer energy consumption	=	1.7 MM Btu/hr
Net gain from compressed vapor	=	9.5 MM Btu/hr
Compression power	=	445 kW
Added dryer costs	=	\$100,000
Added compressor costs	=	\$190,000
Assume vapor condensate handling costs	=	direct dryer haze control costs
Reduction in operating costs	=	\$130,900 ^a
Payback period	=	2.2 years

^a(11.2 MM Btu/hr x \$3.10/MM - 445 kW x \$.0407/kWhr) x 7884 hr

TABLE 5-10

Steam Demands with an Indirect Dryer (MM Btu/Hr)

	<u>30 Psig</u>	<u>150 Psig</u>
Methanol Distillation	9.06	-
Acid Gas Stripping	0.96	-
Shift Steam	-	2.1
Wood Drying	-	14.58 ^a (direct and indirect)

^aRelative to liquid H₂O at 77°F



TABLE 5-11

Phenol Treatment Methods

<u>Method</u>	<u>Feasibility for Net Gasifier Condensate</u> ^(a)
Solvent Extraction	Good - Low capital cost for simple removal. However, removal to low ppm level is expensive. Needs piloting to design for specific waste streams.
Activated Sludge/Aerobic Digestion	Good - High cost because of dilution required and small scale but effective removal - specific dosage and dilution will depend on specific contaminants. Should pilot.
Reverse Osmosis	Poor - Only about 70 percent removal per pass; several passes in series required.
Ozonation	Poor - Not all species will be effectively removed. Expensive.
Wet Oxidation	Poor - Not all species will be effectively removed. Expensive.
Incineration	Poor - Mostly water is evaporated, energy intensive - about 2.5 MM Btu/hr.
Activated Carbon	Poor - Not all species will be effectively removed.

^aBasis is 5 gpm net liquid production containing 2000-3000 wt ppm phenols.

TABLE 5-12

Methods for Eliminating Hydrogen Sulfide from Acid Gas

<u>Method</u>	<u>Comments</u>
Molecular Sieve Adsorption	Selective adsorption of water - gas must be dried. Low usage of adsorbent at low hydrogen sulfide partial pressures. Regeneration required.
Chemical Oxidation	Reagents such as potassium permanganate, ozone sodium hypochlorite or hydrogen peroxide are expensive. Acid or salt purge must be handled.
Caustic Scrubbing	Expensive because carbon dioxide will react with solution forming carbonates.
Sponge Iron Adsorption/ Oxidation	Simple technology with little capital equipment required. Compression of acid gas to 15-20 psig is required. Sulfur rate is so low that sponge can be thrown away. Selective for hydrogen sulfide.
Liquid Catalytic Oxidation	Commercial at low pressure with hydrogen sulfide concentrations of 400 ppm or higher. Suitable for removal to 5 ppm at low pressure. More expensive in small scale than sponge iron. Standard Stretford reagent is toxic (there is a non-toxic version). Lo-Cat reagent is nontoxic. Solid sulfur product.



TABLE 5-12 (cont'd)

Catalytic Incineration

Difficult to purchase vendor equipment in this small scale. Fuel costs will be about 1 MM Btu per 1000 scf. Capital intensive because of heat recovery required to reduce fuel costs.



Table 5-13

Startup Procedures for Gas Shift (Section 300)

1. Preheat steam lines to the shift reactor and drain any condensate formed. Maintain an open drain at the bottom of the reactor to prevent the buildup of condensate.
2. Introduce superheated steam (150 psia, 500^o F) from the shift startup superheater to the reactor and heat to 310^o F. Close the drain.
3. At 310^o F, substitute raw gas (cooled by water injection) at the normal operating temperature to raise the catalyst bed inlet temperature to 440^oF. Flare shift product and bypassed raw gas upstream of R-401 (synthesis gas compressor).



Table 5-14

Startup Procedures for Acid Gas Removal (Section 500)

- 1.^a Wash the acid gas removal system first with water, then with a weak potassium carbonate solution, and finally with water.
- 2.^a Drain
3. Purge equipment with nitrogen
4. Pressurize the acid gas absorber to operating pressure with nitrogen.
5. Heat system by steam injection to absorber and stripper.
6. Prepare solution in carbonate mix tank and circulate through system.
7. Open steam to stripper reboiler.
8. Introduce synthesis gas to system.
9. Start air coolers and open cooling water flow to system.
10. Route clean gas to flare (downstream of R-402) in order to purge system of nitrogen.
11. Once nitrogen has been removed from the system, open line to methanol synthesis.

^aRequired only for initial startup or after extended downtime.



Table 5-15

Startup Procedures for Methanol Synthesis (Section 600)

1. Reduce the synthesis catalyst (initial startup only).
2. Put the synthesis loop under low nitrogen pressure.
3. Fill the reactor jacket with boiler feed water.
4. Preheat the boiler feed water in the methanol reactor with saturated steam generated in the startup heater (F-603).
5. Start coolers.
6. Introduce feed to the synthesis loop by reducing flow to the flare from downstream of the guard adsorbers and start recycle compressor.
7. Open raw product line to purification.
8. Open purge line.



Table 5-16

Startup Procedures for Methanol Purification and Storage (Section 700)

1. Purge system with nitrogen to remove any oxygen.
2. Prepare caustic solution.
3. Open line from raw product separator (M-602) to flash separator (M-701).
4. Load stabilizer with liquid and start reboiler and air cooler. Run stabilizer on recycle until proper operating conditions are reached. Divert stabilizer product to intermediate tank (non-specification methanol in the intermediate tank is returned to the flash separator during normal operation).
5. Fill methanol still with liquid and start reboiler and air cooler. Bring still to operating conditions. Send net product to the intermediate tank and net water to Section 800.
6. When methanol product meets specifications, send net product to storage.



Table 5-17

Shutdown Procedures for Plant Sections 200 through 800

Section 800 - Purge Gas Treatment

1. Switch feed to unit to flare.
2. Stop new pressure swing cycles on PSA unit.
3. During last depressurization, begin switching diesels to propane fuel.
4. Shut down exchanger T-801 and pump P-801 after line from A-702 has been closed.

Section 700 - Methanol Purification and Storage

1. Close line from methanol still reflux drum (M-703) to product storage and send product to intermediate tank (Q-701). Stop pump P-703 and cooling water to T-707.
2. Close discharge water line from A-702 to Section 800.
3. Stop flow of methanol from A-701 to A-702 and divert to intermediate tank. Shut off condensate flow to T-704.
4. Shut off steam to methanol still reboiler T-705. When column cools, shut off product air cooler T-706 and reflux pump P-705. Drain column into intermediate tank.
5. Stop flow of raw methanol to stabilizer and divert flow from M-701 into intermediate tank. Shut off condensate flow to T-701.



TABLE 5-17 (cont'd)

6. Shut off steam to stabilizer reboiler T-702. When column cools, shut off air cooler T-703 and reflux pump P-704. Drain column into intermediate tank.
7. Purge methanol vapors out of system with nitrogen. Repurge M-701 when the last of the raw methanol product has been removed from M-602 into intermediate tank Q-701.

Section 600 - Methanol Synthesis

1. Stop feed of synthesis gas to the section and flare gas.

For Short Shutdowns

2. Isolate methanol synthesis loop and guard bed area and stop recycle compressor.
3. Maintain reactor jacket pressure by heating jacket.

For Longer Shutdowns

4. Depressurize methanol synthesis loop and guard bed area to low pressure, venting to flare.
5. Leave boiler feed water in reactor L-602 jacket.
6. Purge lines and L-602 tubes with nitrogen, venting to flare.

Section 500

1. Stop process gas flow to acid gas removal system.
2. Purge system with nitrogen at low pressure.

TABLE 5-17 (cont'd)

3. Shut off steam to acid gas reboiler.
4. Recover carbonate solution and pump to solution makeup tank.

Section 400 - Gas Cooling and Compression

1. Stop compressors when flow of gas to Section 500 is stopped. Divert gas to flare downstream of T-201.
2. Shut down T-403 and boiler feed water flow to T-402.
3. Shut down pump P-401 and stop cooling water flow to T-401.
4. When Section 300 is depressurized, purge Section 400 and Section 300.

Section 300 - Gas Shift

1. Stop T-303 when flow of gas to Section 300 is stopped.
2. Stop water injection.
3. Vent to flare. Purge reactor and lines with nitrogen.
4. Isolate reactor and maintain positive pressure with nitrogen.

Section 200 - Gasification

1. Stop oxygen feed to gasifier.
2. Stop wood feed to gasifier and lock hoppers.



TABLE 5-17 (cont'd)

3. Depressurize and vent system.
4. Start air blower (R-201) and burn out bed material (for prolonged shutdowns).



TABLE 5-18
VALVE, PIPING AND INSTRUMENT SYMBOLS

1/2







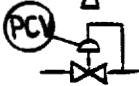


-  Manually Operated Valve
-  Valve, Normally Closed
-  Valve, Unspecified, Normally Open
-  Valve and Pneumatic Actuator
- Electronic Signal Connection
- ### Air Cooler Damper
-  Electric Motor
-  Pressure Relief Valve
-  Pressure Regulator (Control Valve)
- TE Temperature Element (Thermocouple)
- LT Level Transmitter (Type Unspecified)
- LIT Level Transmitter and Indicator
- PT Pressure Transmitter (Direct Connection)
-  Pressure Transmitter With Indirect (Diaphragm) Connection
- PIT Pressure Transmitter and Indicator
- FE Flow Element (Unspecified Measuring Device)
- FT Flow Transmitter (Type Unspecified)
- A Analyzer (Off Line) Used At This Location To Determine Composition And/Or Presence Of Specific Compounds
-  pH Analyzer

TABLE 5-18(Cont.)
VALVE, PIPING AND INSTRUMENT SYMBOLS, Cont.

TIC	Temperature Indicator and Controller
PIC	Pressure Indicator and Controller
FIC	Flow Indicator and Controller
FC	Flow Controller
FCV	Flow Control Valve
PCV	Pressure Control Valve
LCV	Level Control Valve
SCR	(Silicon Controlled) Rectifier, i.e. Electronic Motor Variable Speed Control
R	(Electric) Relay
CR	(Signal From) Control Room Operated

Table 5-19

Overall Skid List with Dimensions

<u>Section</u>	<u>Skid Number</u>	<u>Dimensions (ft)</u>
200	1	12x13x38
	2	11x12x35
	3	8x10x23
	4	8x6x6
300	1	10x12x20
	2	10x12x4
400	1	12x12x11
	2	Vendor Supplied
500	1	6x8x10
	2	12x12x16
	3	8x12x6
	4	8x6x4
	5	12x10x4
	6	12x10x4
	7	Vendor Supplied
	8	12x12x12
600	1	12x27x12
	2	12x15x4
	3	12x12x16
	4	6x4x6
700	1	8x13x7
	2	10x12x11
	3	6x10x4



Table 5-19 (Cont'd)

<u>Section</u>	<u>Skid Number</u>	<u>Dimensions (ft)</u>
	4	Vendor Supplied
	5	12x9x13
	6	12x15x8
	7	12x24x4
800	1	Vendor Supplied
	2	6x10x11
900	1	15x6x12
1000	1	5x5x12
	2	Vendor Supplied
	3	12x5x4
	4	12x5x4
	5	Vendor Supplied
	6	Vendor Supplied
	7	24x6x4
	8	Vendor Supplied
	9	Vendor Supplied
	10	12x5x12
1100	1	10x10x16
	2	6x6x12
	3	Vendor Supplied
	4	Vendor Supplied
	5	Vendor Supplied
	6	Vendor Supplied



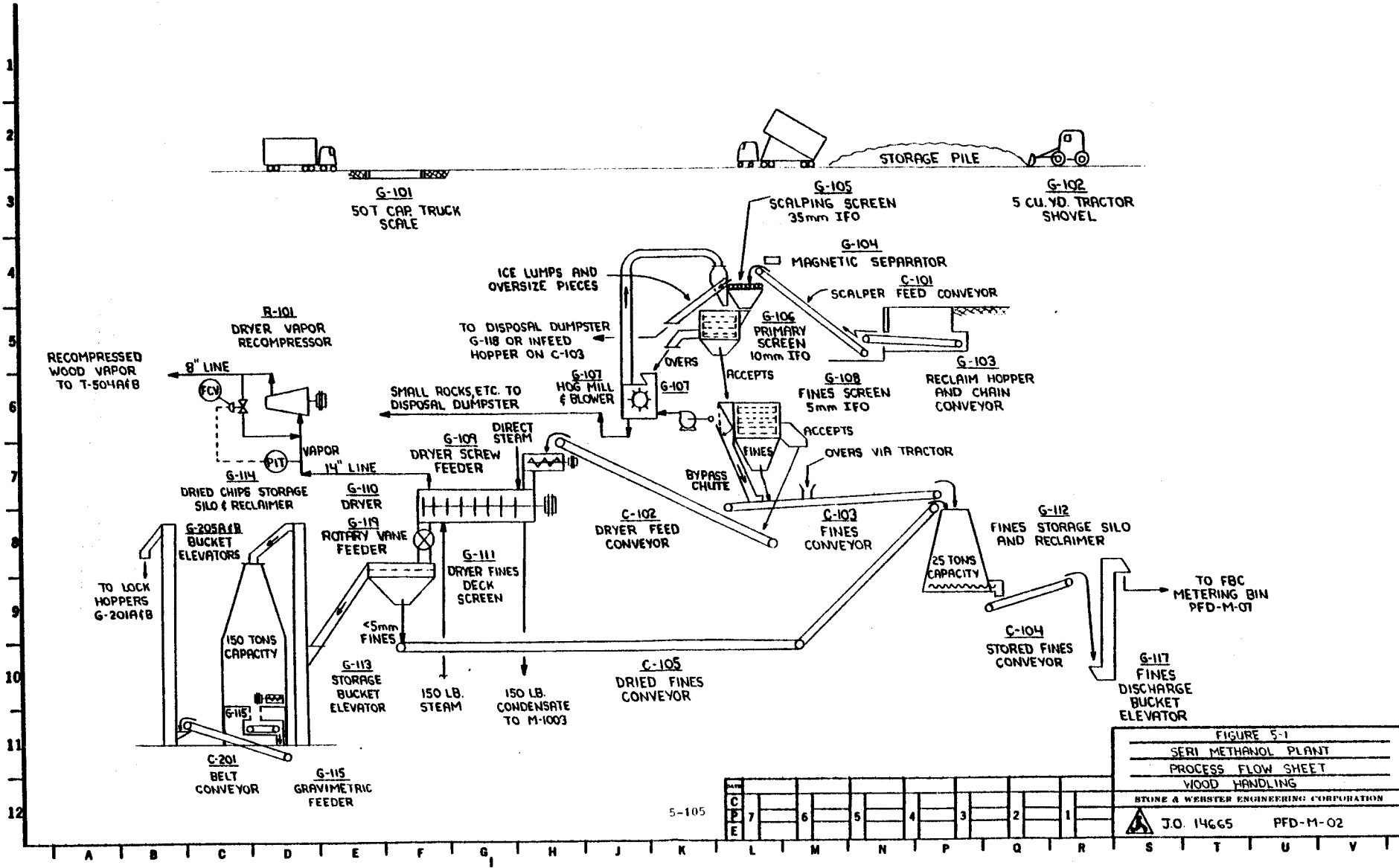
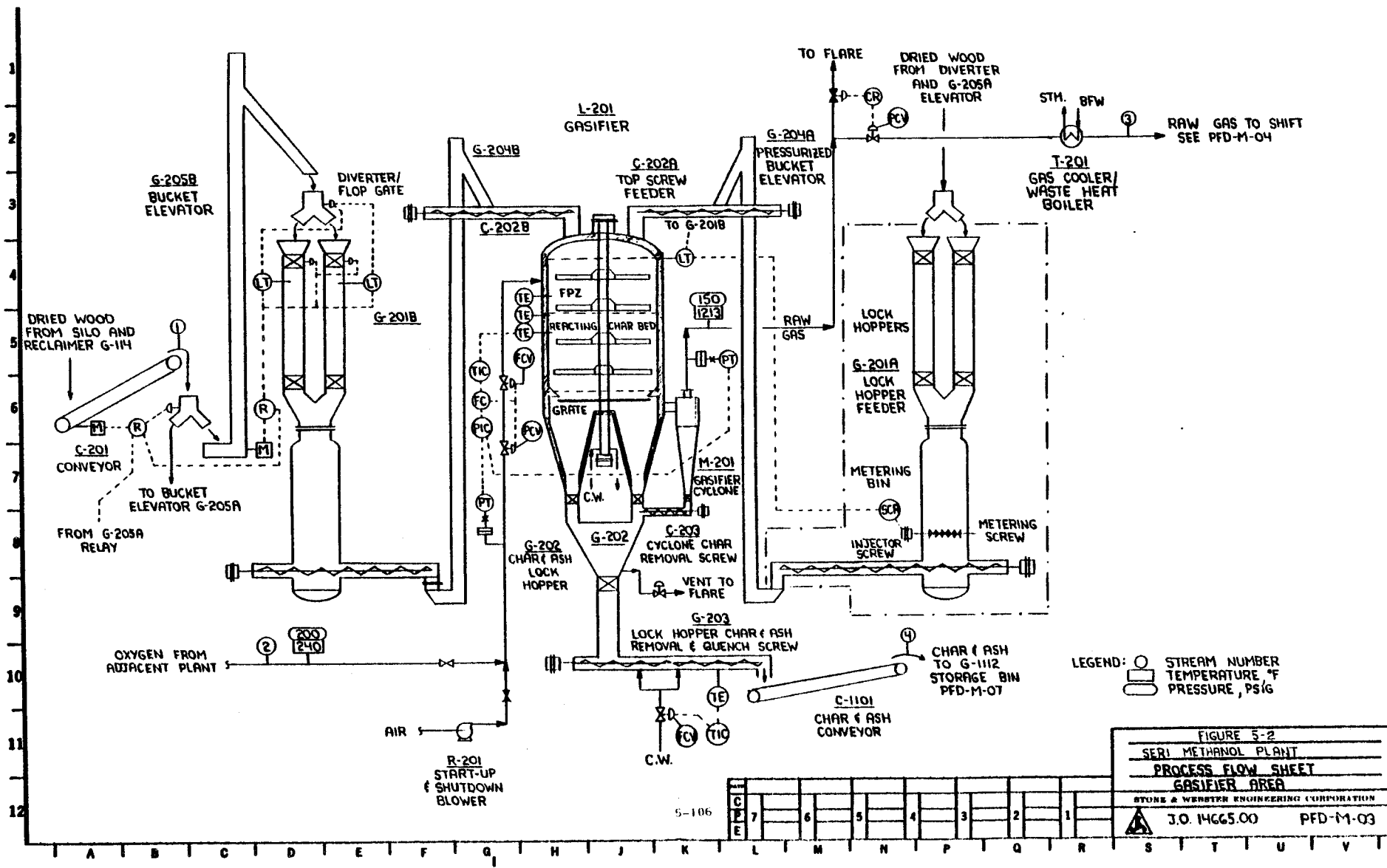
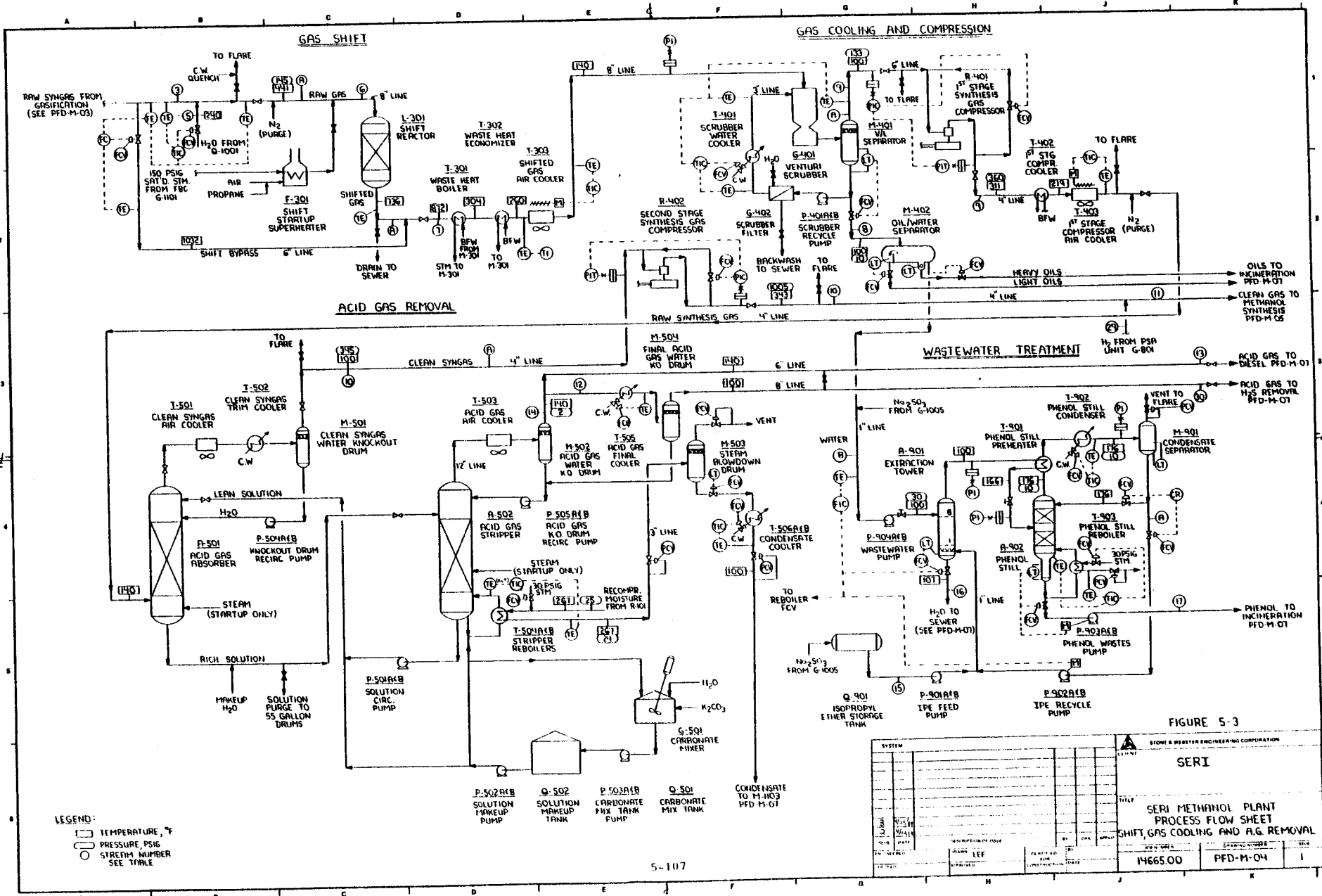


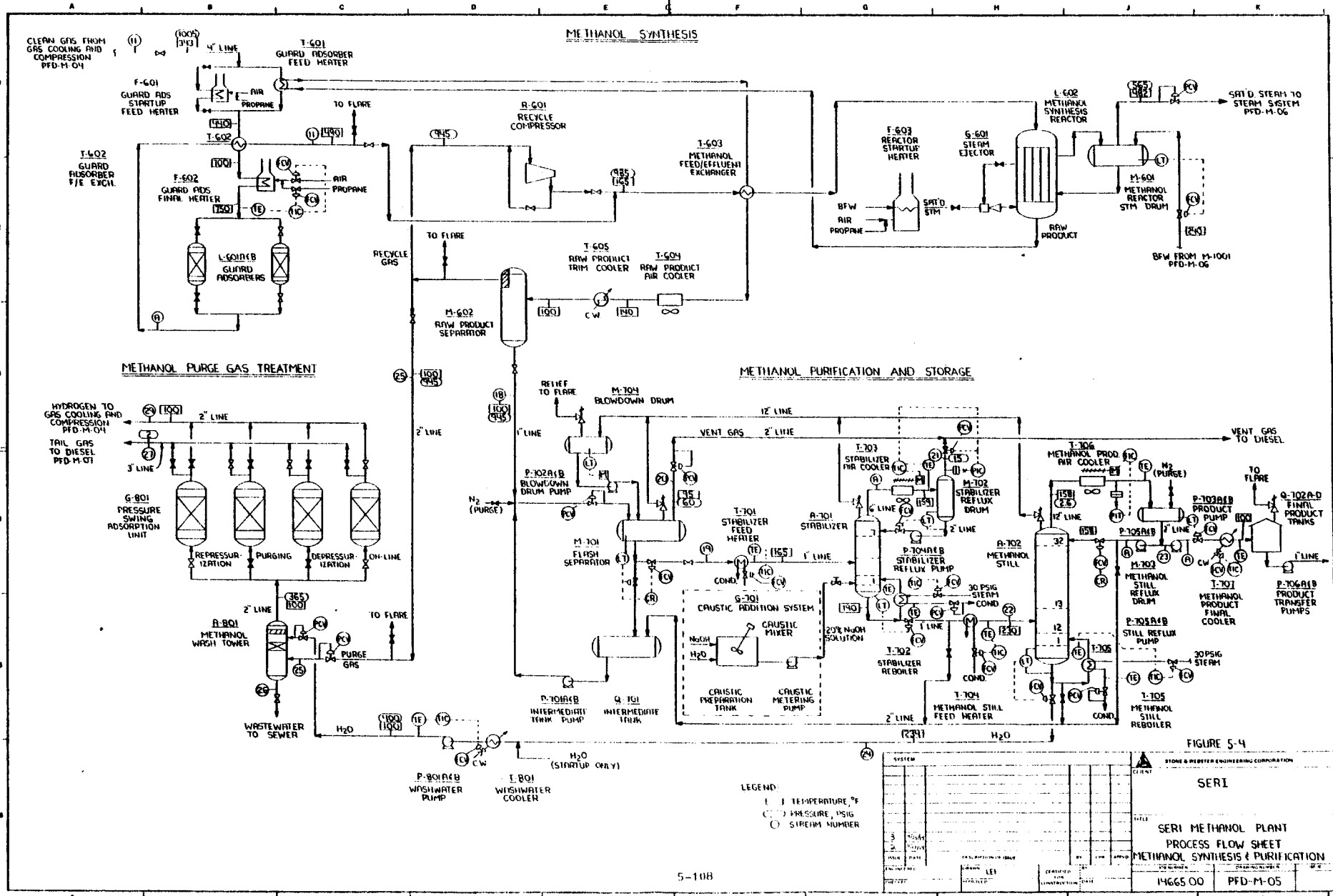
FIGURE 5-1
 SERI METHANOL PLANT
 PROCESS FLOW SHEET
 WOOD HANDLING

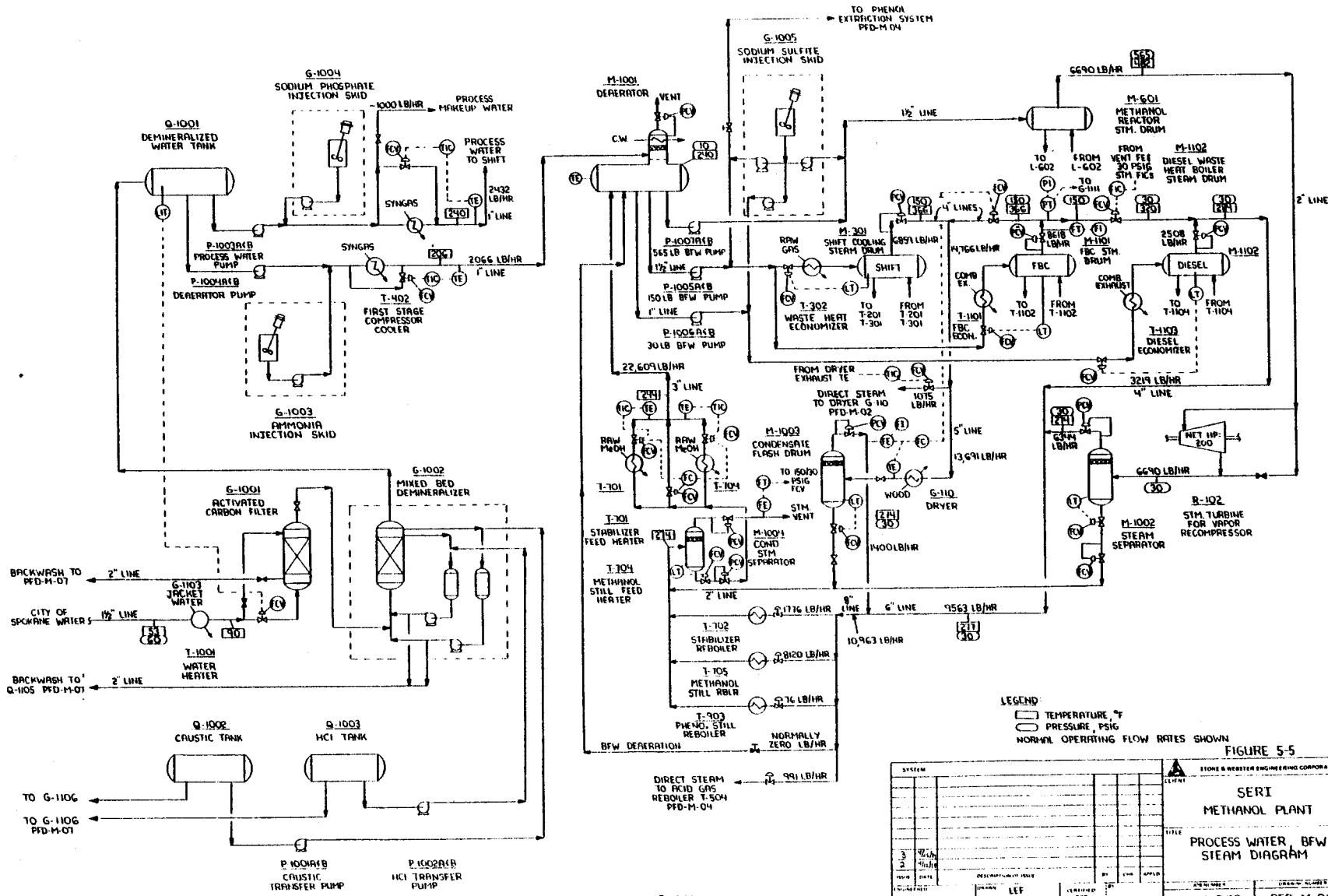
STONE & WEBSTER ENGINEERING CORPORATION
 J.O. 14665 PFD-M-02

7	6	5	4	3	2	1
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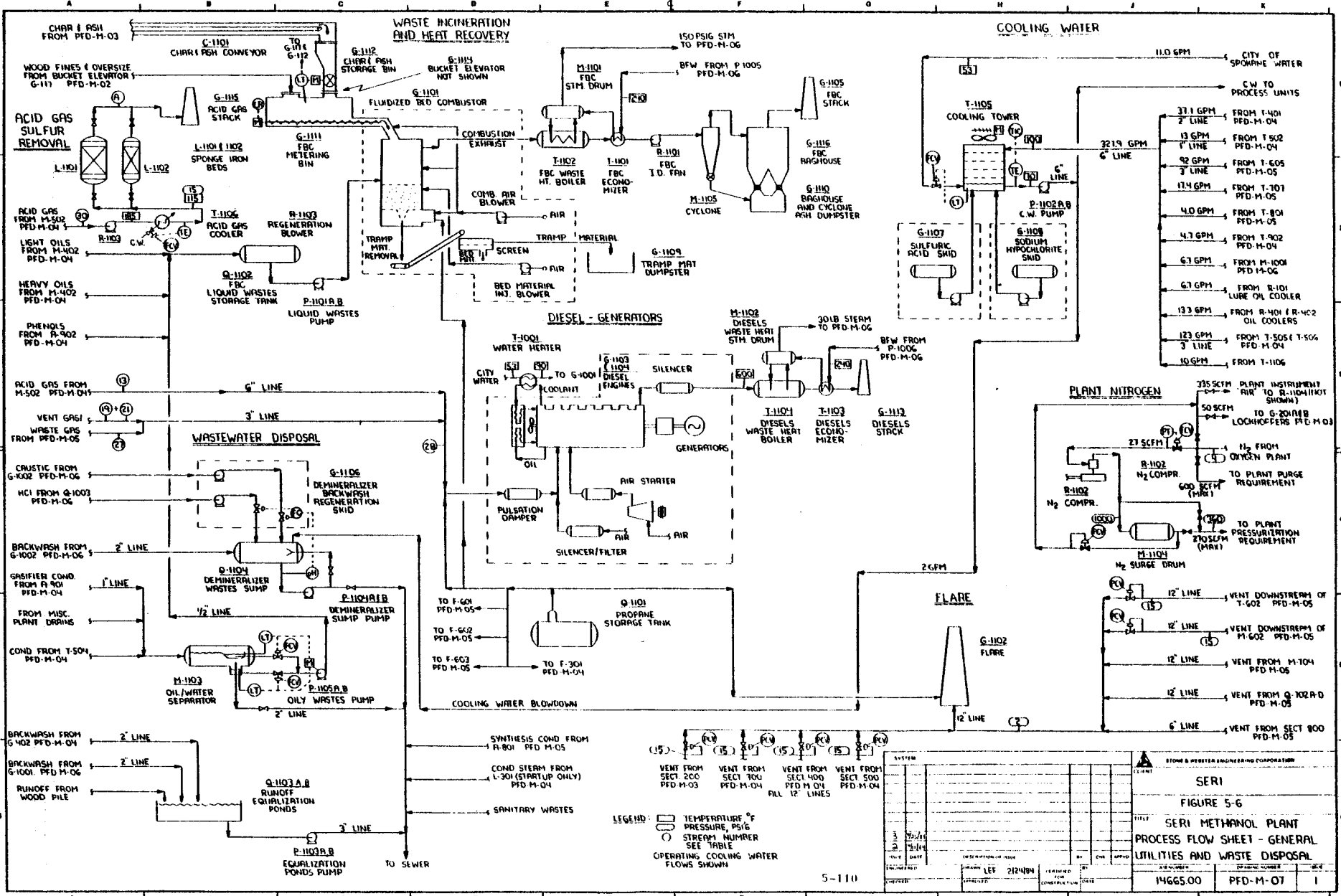


5-109

SYSTEM		DATE		BY		CHECKED		APPROVED	
1	DESIGNED								
2	DRAWN								
3	CHECKED								
4	APPROVED								

PROJECT	14665.00	DATE	PFD-M-06	NO.	1
TITLE	SERI METHANOL PLANT PROCESS WATER, BW, STEAM DIAGRAM				
DESIGNED BY	LEE	CHECKED BY		APPROVED BY	
DATE		DATE		DATE	

FIGURE 5-5
 STORE & NESTER ENGINEERING CORPORATION
 SERI
 METHANOL PLANT
 PROCESS WATER, BW, STEAM DIAGRAM
 14665.00 PFD-M-06



LEGEND: □ TEMPERATURE °F
 ○ PRESSURE, PSIG
 ○ STREAM NUMBER
 SEE TABLE
 OPERATING, COOLING WATER
 FLOWS SHOWN

(15)	(16)	(15)	(15)	(15)	(15)	(15)	(15)
VENT FROM SECT 200 PFD-M-05	VENT FROM SECT 300 PFD-M-04	VENT FROM SECT 400 PFD-M-04	VENT FROM SECT 500 PFD-M-04	ALL 12" LINES			

NO.	DATE	DESCRIPTION	BY	CHKD	APPROV
3	7/2/14				
2	7/2/14				
1	7/2/14				

STONE & WEBSTER ENGINEERING CORPORATION	
SERI	
FIGURE 5-6	
SERI METHANOL PLANT	
PROCESS FLOW SHEET - GENERAL	
UTILITIES AND WASTE DISPOSAL	
14665.00	PFD-M-07
	1

FIGURE 5-7
PROPOSED WASTEWATER
 FLOW DIAGRAM

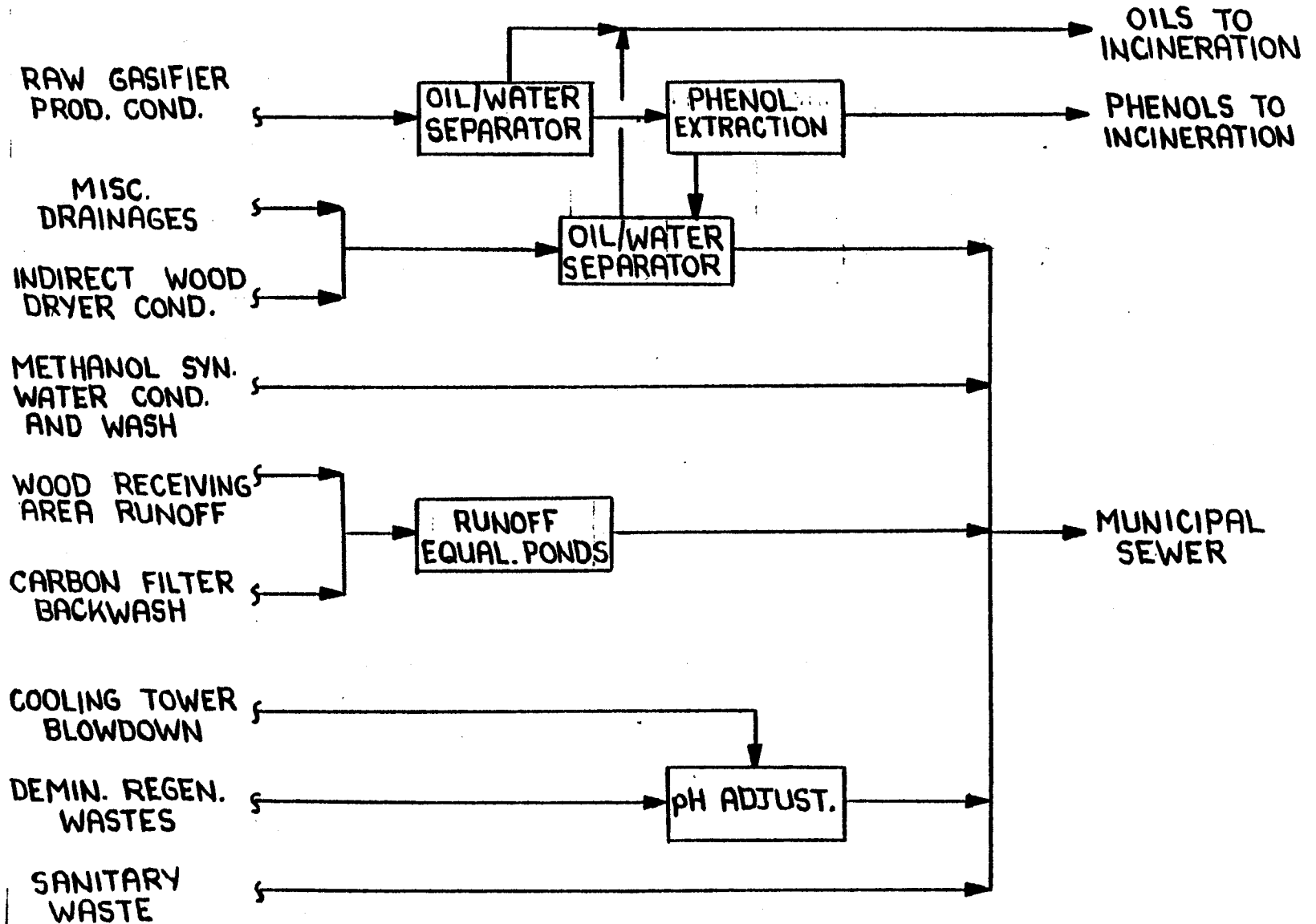


FIGURE 5-8
 BASE CASE MATERIAL BALANCE
 (ALL FLOWS IN TONS/DAY)

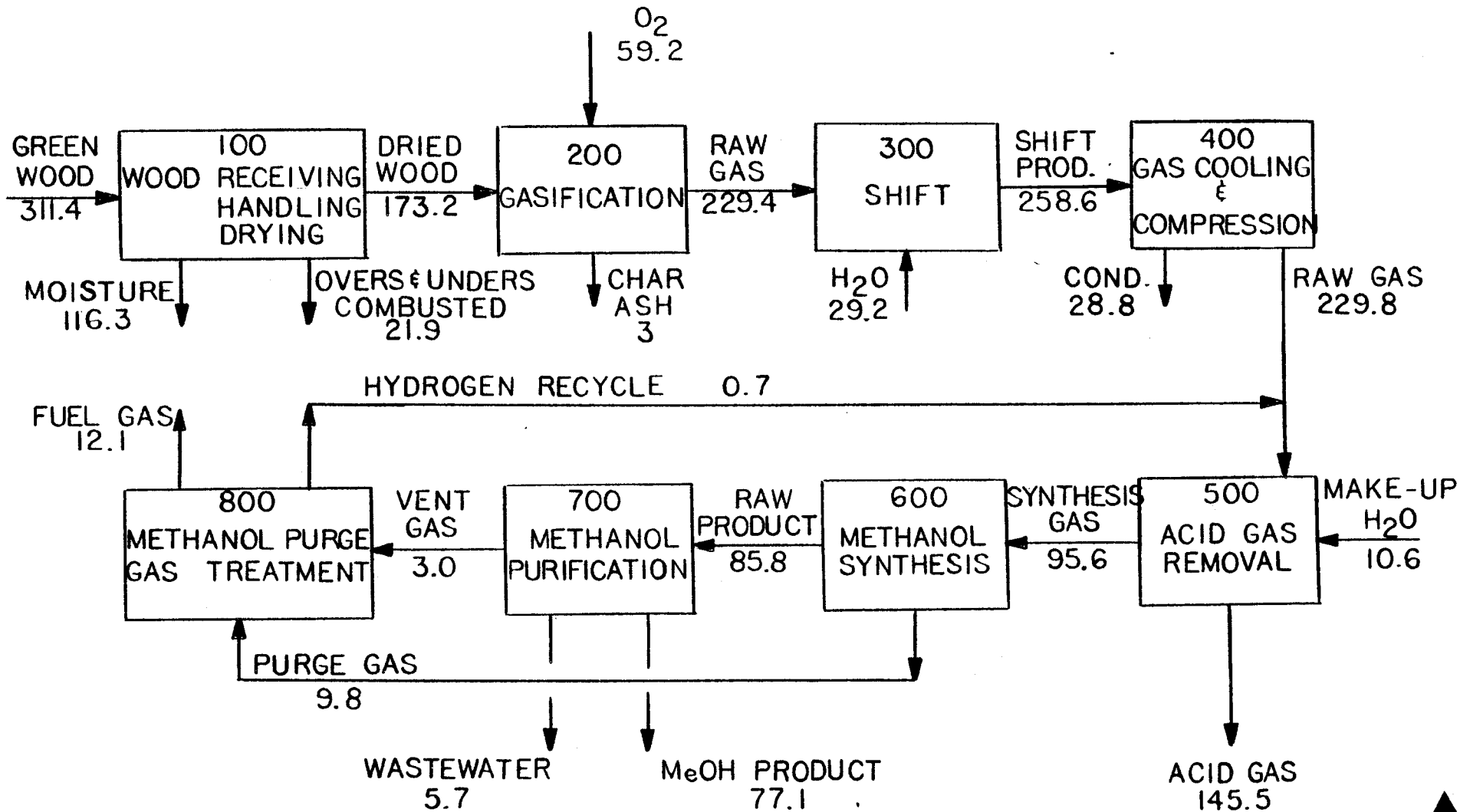


FIGURE 5-9 BASE CASE ENERGY BALANCE

ALL FLOWS IN MM BTU/DAY
REFERENCE STATE H₂O (LIQUID) AND 77°F

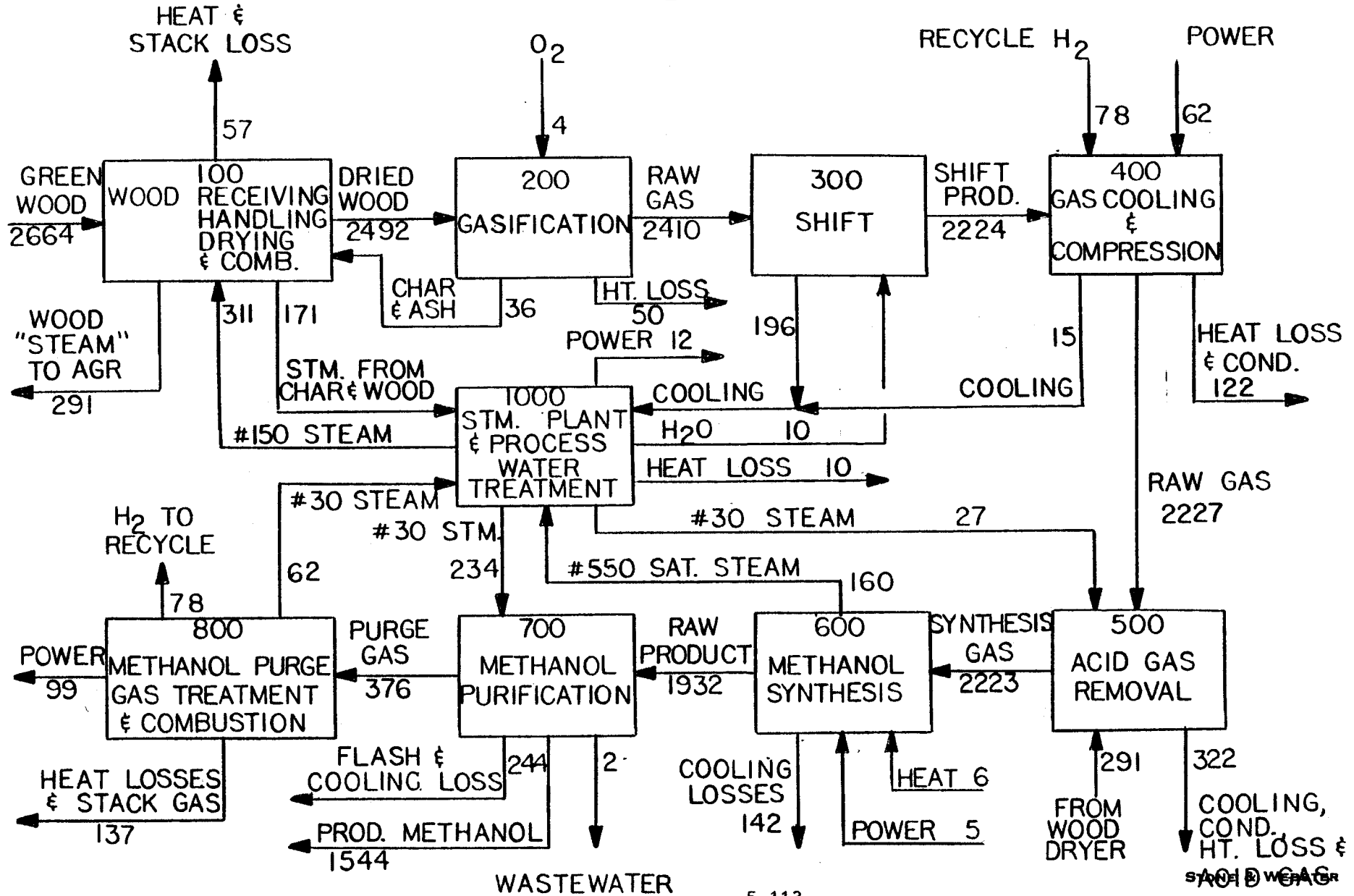
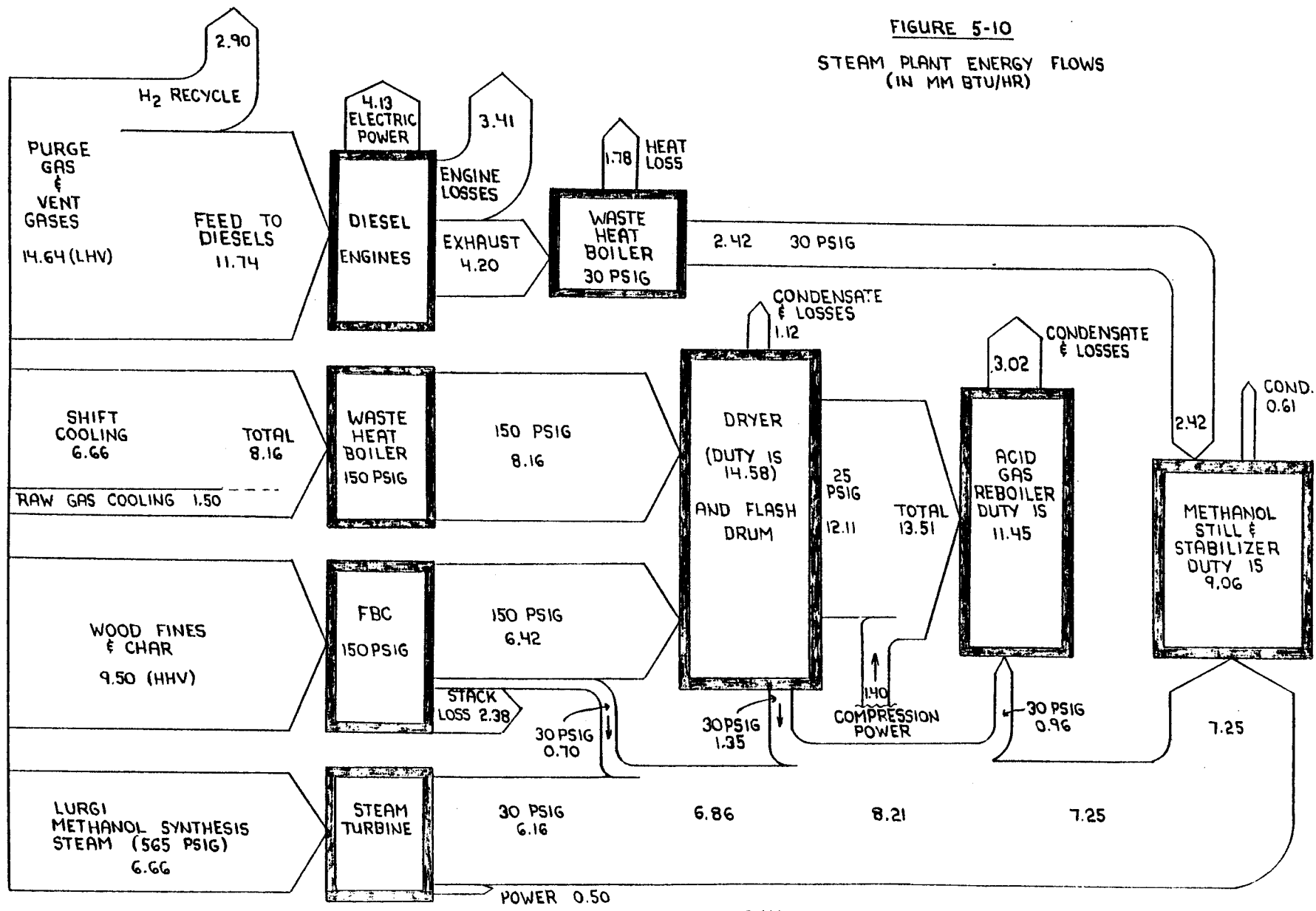
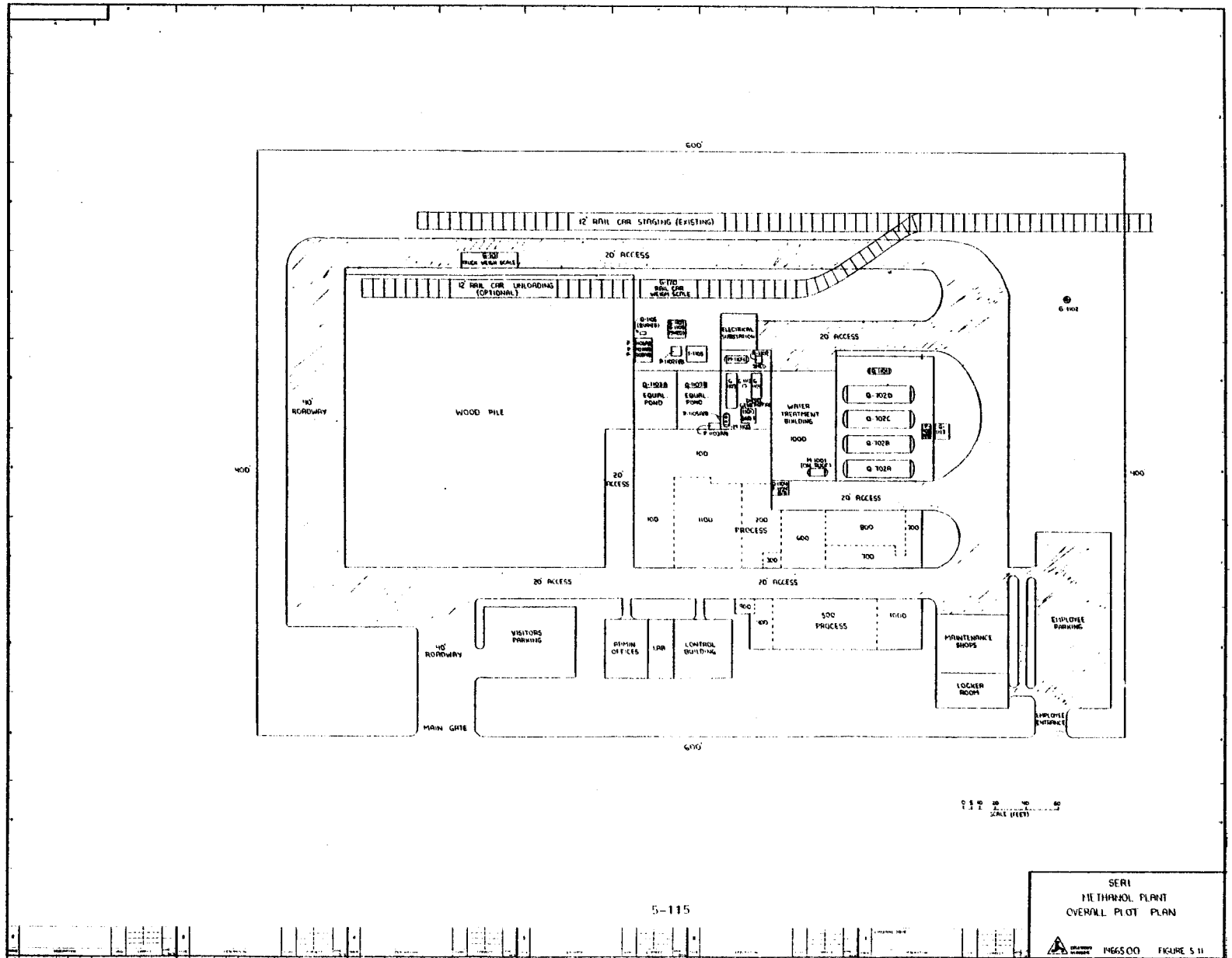
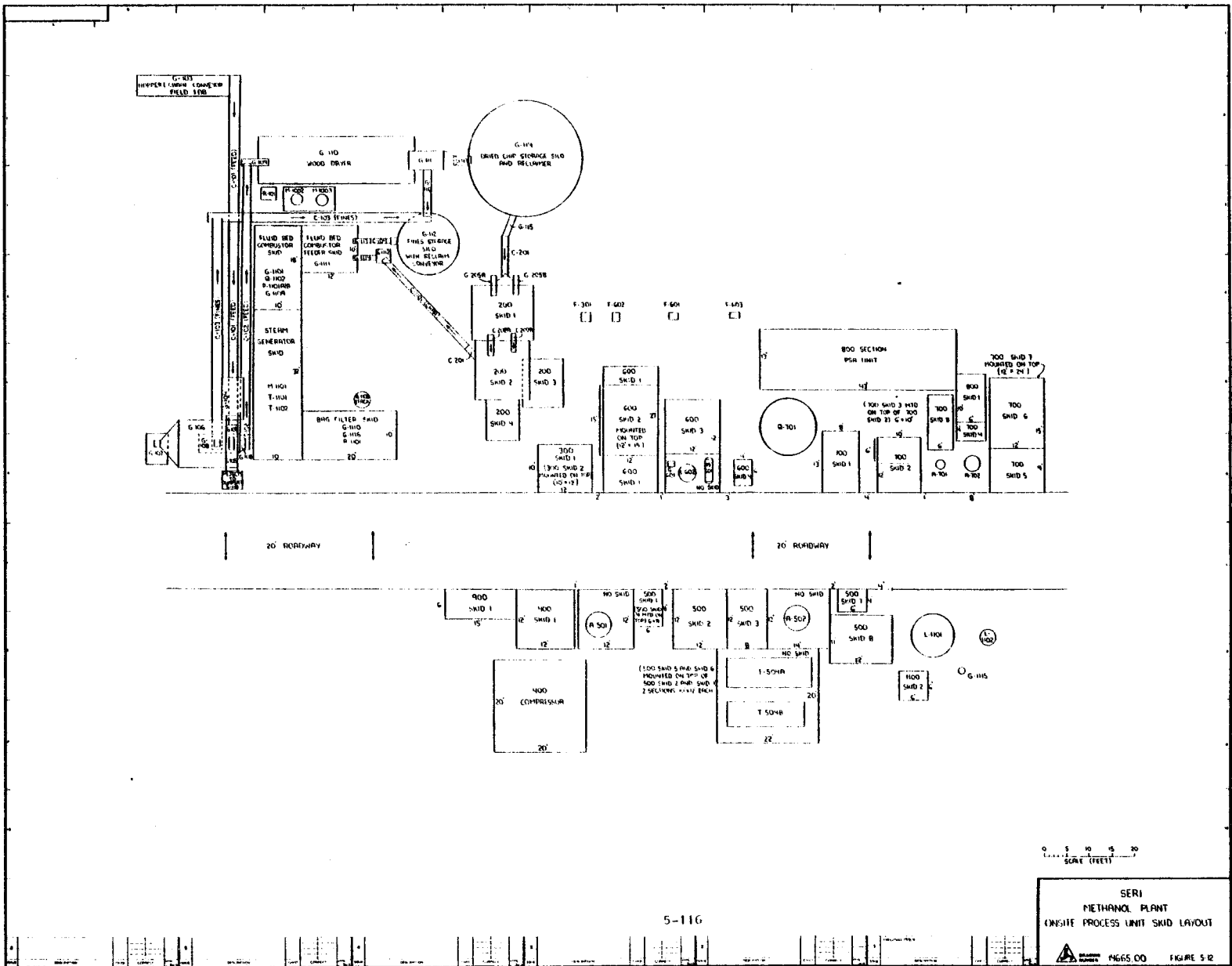
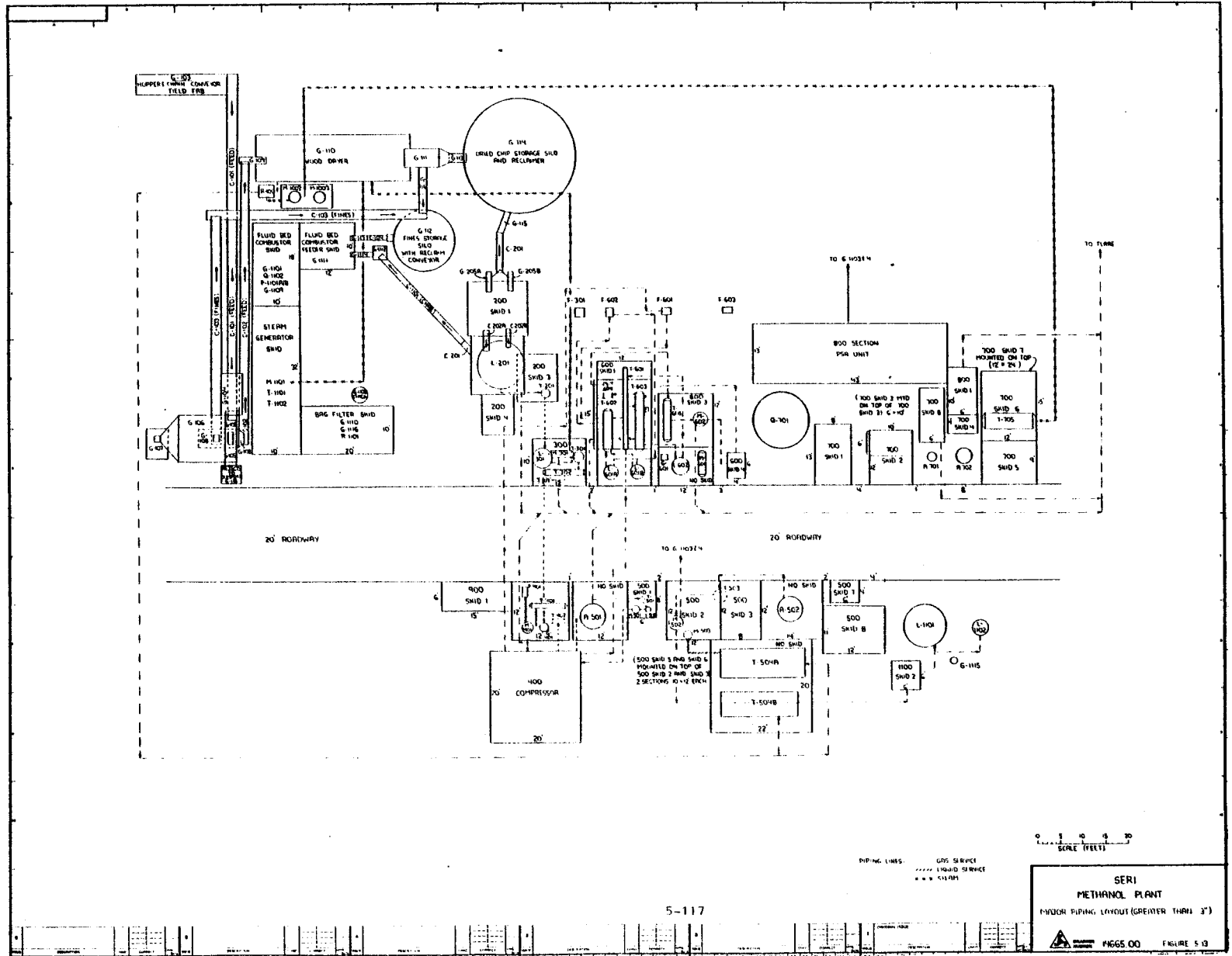


FIGURE 5-10
 STEAM PLANT ENERGY FLOWS
 (IN MM BTU/HR)









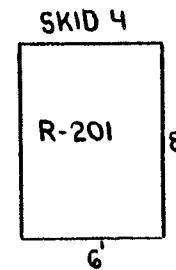
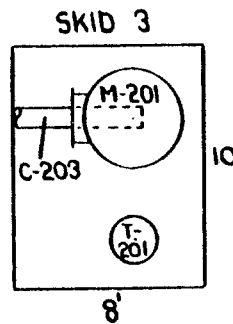
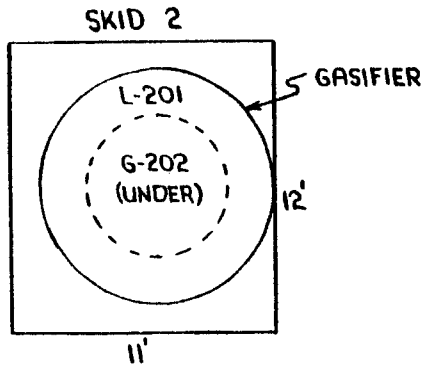
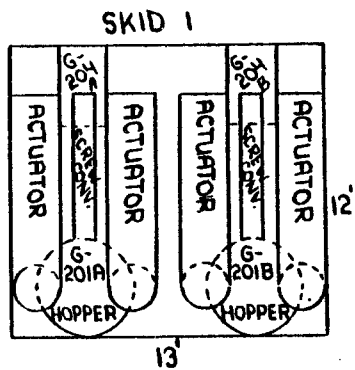
5-117

0 5 10 15 20
SCALE (FEET)

PIPING LINES: ——— GAS SERVICE
- - - - - LIQUID SERVICE
· · · · · CONTROL

SERI
METHANOL PLANT
MAJOR PIPING LAYOUT (GREATER THAN 3")
PIGGS OOO FIGURE 5 13

FIG. 5-14
200 SECTION
 GASIFICATION SYSTEM SKIDS



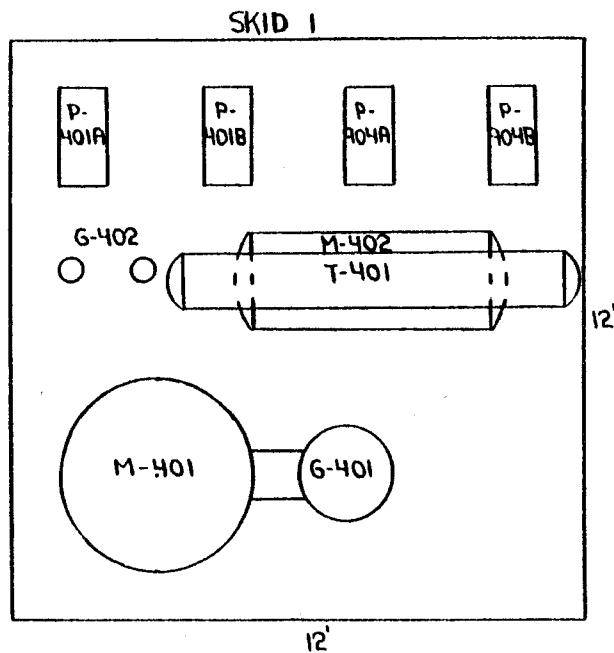
SCALE: $\overline{\text{---} 5'-0'' \text{---}}$

5-118

				SCALE		▲ STONE & WEBSTER ENGINEERING CORPORATION			
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				CHECKED					
NO	REVISION DESCRIPTION	BY	DATE	DRAWN	AREA	I.O. NO.	DRAWING NO.	ISSUE	
	CERT FOR FAB		CERT FOR CONRY	LEF			FIG. 5-14	1	
			APPROVED	CHECKED					

FIG. 5-16 400 SECTION

GAS COOLING + COMPRESSION SKID



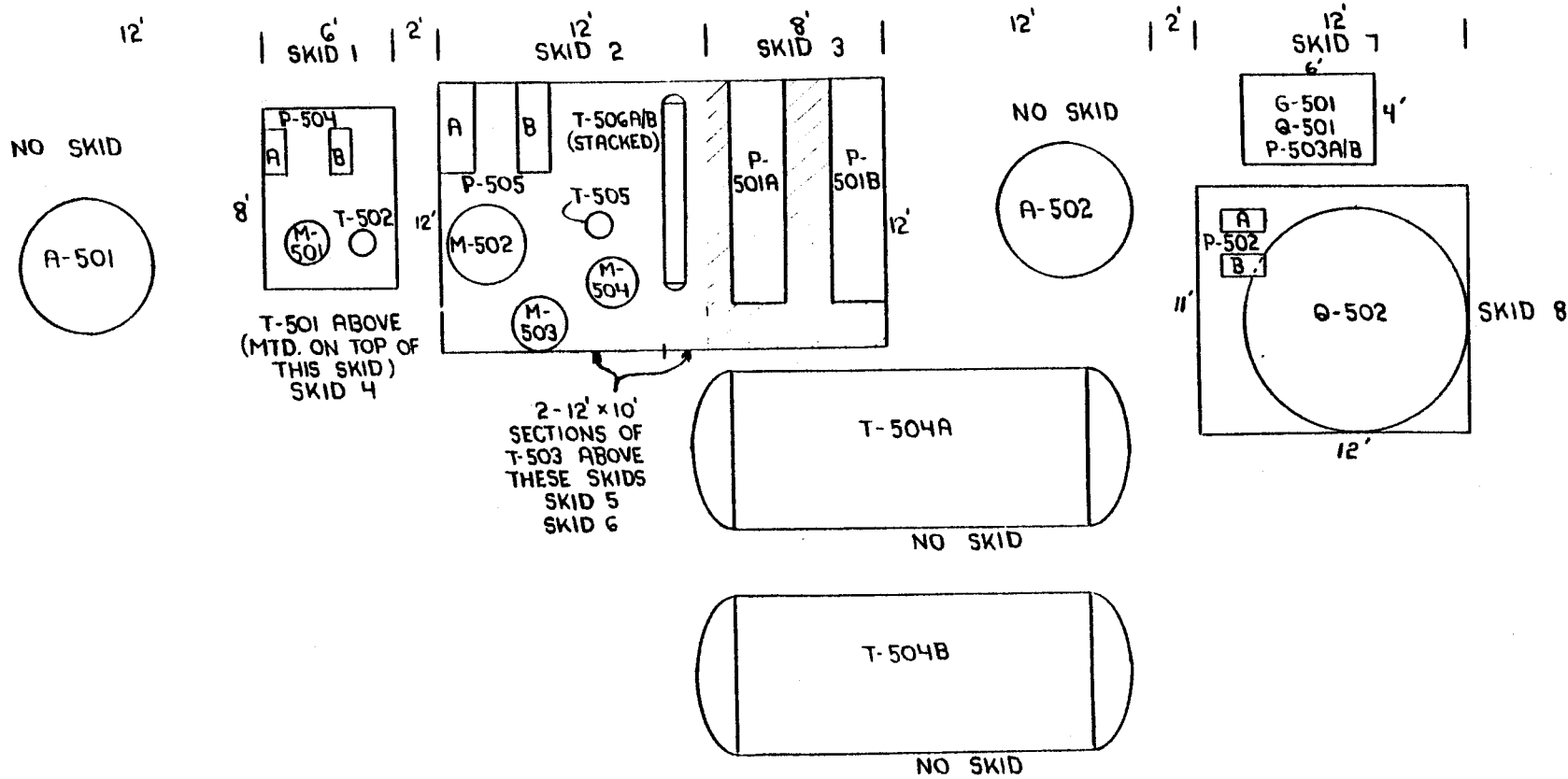
NOTE: 20' x 20' SYNTHESIS GAS COMPRESSOR
MODULE BY VENDOR.

SCALE: $\sim 5'-0''$

5-120

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				DESIGN		SERI METHANOL PLANT			
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NO	REVISION DESCRIPTION	BY	DATE	DRAWN	LEF	AREA	J.O. NO.	DRAWING NO.	ISSUE
	CERT FOR FAB		CERT FOR CONST	APPROVED	CHECKED				
						14665.00		FIG. 5-16	1

FIG. 5-17 500 SECTION
ACID GAS REMOVAL SKIDS



SCALE: 5'-0"

5-121

NO REVISION DESCRIPTION		BY	DATE	SCALE	STONE & WEBSTER ENGINEERING CORPORATION			
CERT FOR FAB		CERT FOR CONST	APPROVED	DESIGN	SERI METHANOL PLANT			
				CHECKED	AREA	J.O. NO.	DRAWING NO.	ISSUE
				LEF	14665.00	FIG. 5-17	1	
				CHECKED				

A 506418

FIG. 5-18
600 SECTION
METHANOL SYNTHESIS SKIDS

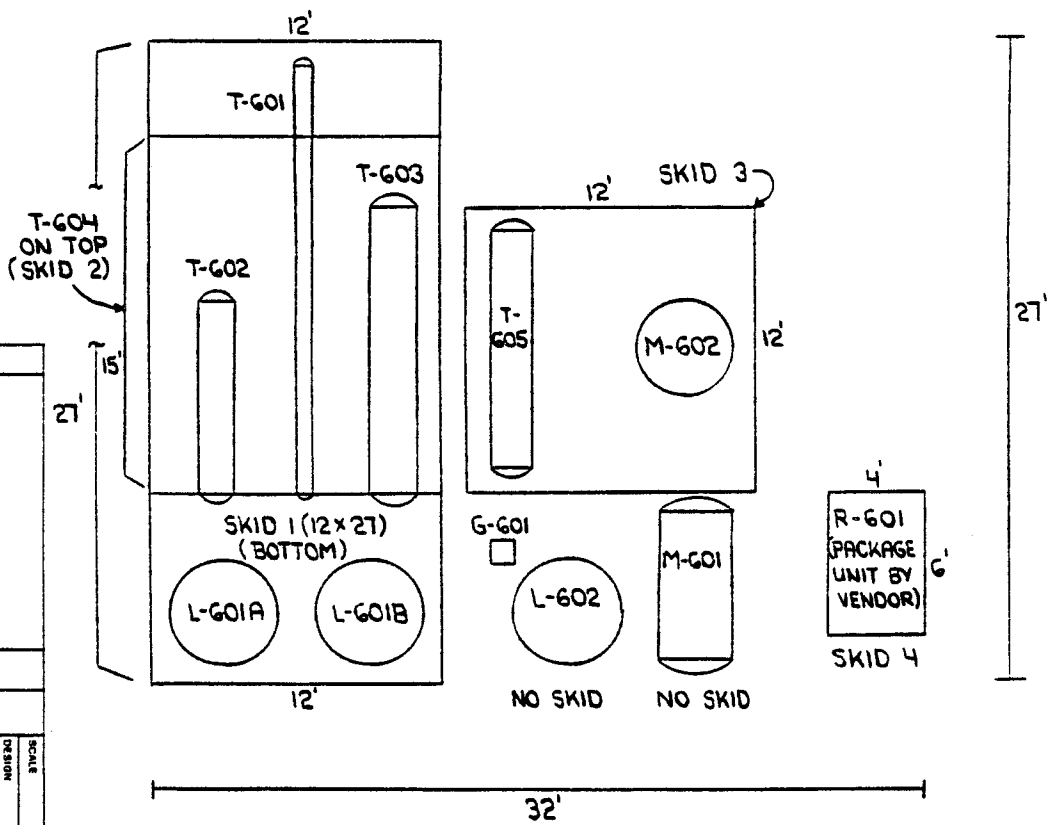
F-602

F-601

F-603



5-122



NO REVISION DESCRIPTION		SCALE	
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CERT FOR CONST		CHECKED	
APPROVED		DATE	
		CHECKED	
		CHECKED	
		DATE	
		ISSUE	
		FIG. 5-18	
		1	

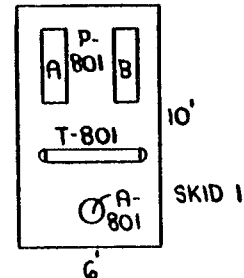
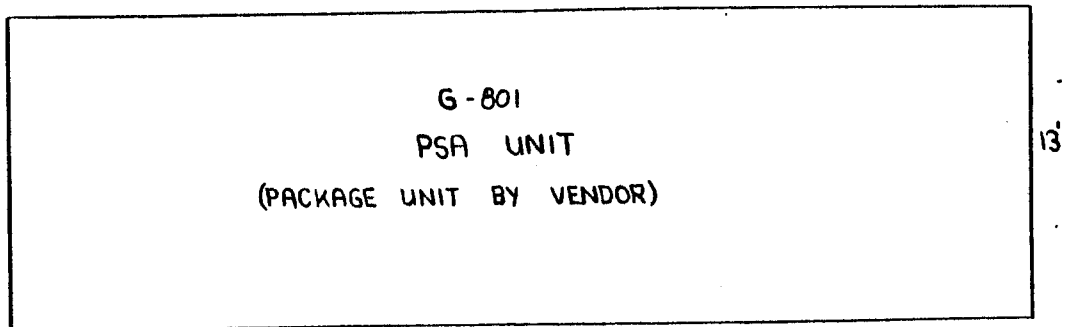
SCALE: 5'-0" = 1'

STONE & WEBSTER ENGINEERING CORPORATION

SERI METHANOL PLANT

ARCH: 1065.00 DRAWING NO: 1
 11/6/50 FIG. 5-18 1

FIG. 5-20
800 SECTION
PURGE GAS TREATMENT SKIDS

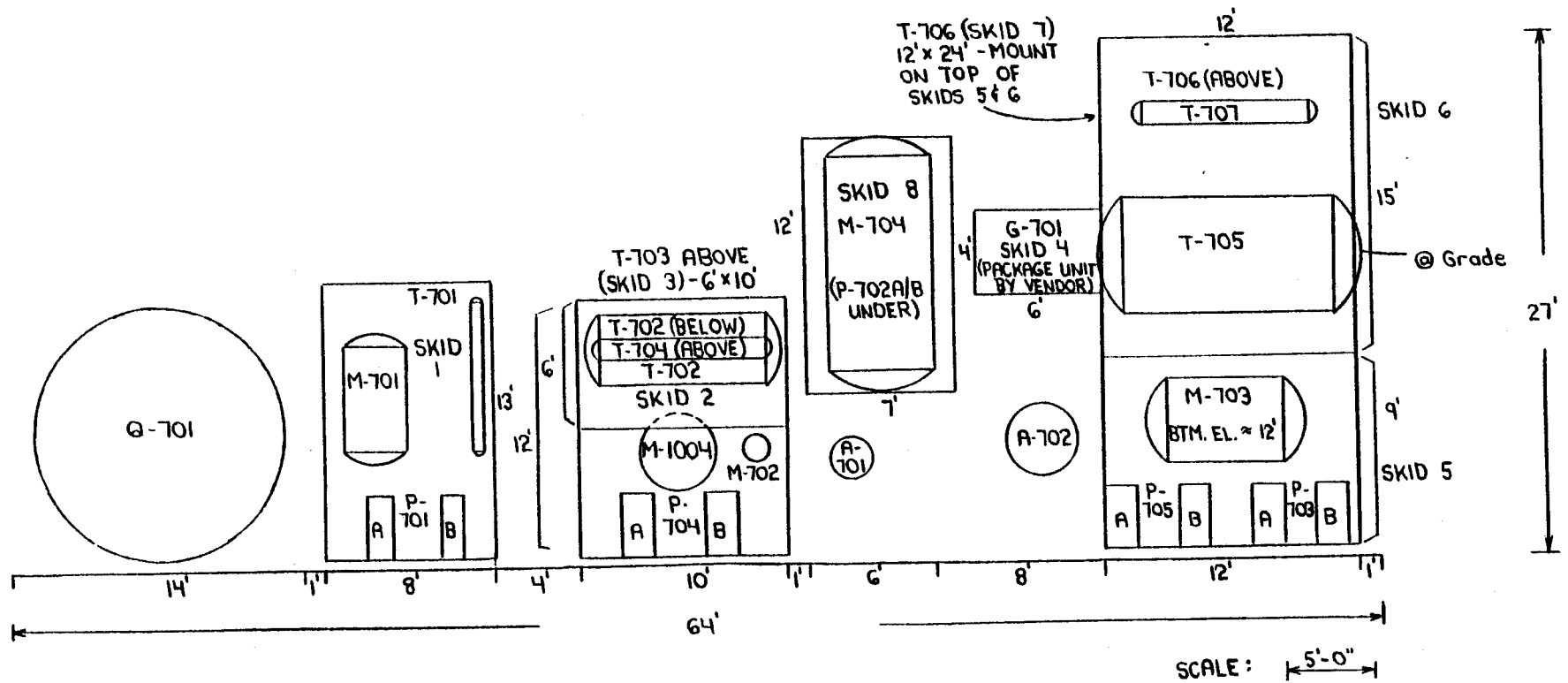


SCALE: 5'-0"

5-124

				SCALE		▲ STONE & WEBSTER ENGINEERING CORPORATION			
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				CHECKED					
NO	REVISION DESCRIPTION	BY	DATE	DRAWN	AREA	J.O. NO.	DRAWING NO.	ISSUE	
	CERT FOR FAB		CERT FOR CONST	LEF		14665.00	FIG. 5-20	1	
				APPROVED		CHECKED			

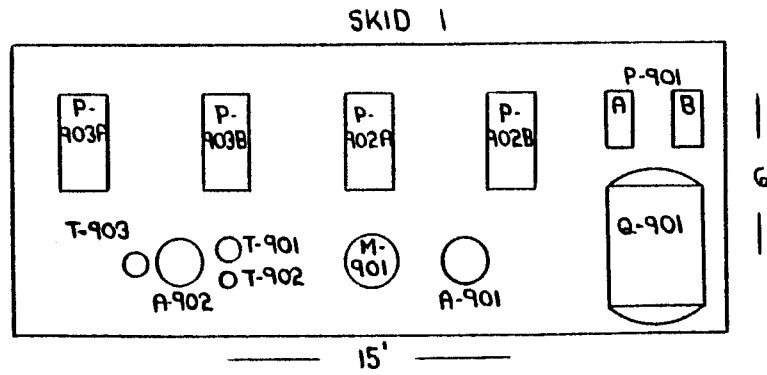
FIG. 5-19
700 SECTION
 METHANOL PURIFICATION SKIDS



5-123

NO REVISION DESCRIPTION		BY	DATE	SCALE	STONE & WEBSTER ENGINEERING CORPORATION		
CERT. FOR FAB				DESIGN	SERI METHANOL PLANT		
CERT. FOR CONST				CHECKED	AREA	J.O. NO.	DRAWING NO.
		APPROVED		DRAWN LEF	14665.00	FIG. 5-19	ISSUE
				CHECKED			1

FIG. 5-21
900 SECTION
 WASTEWATER TREATMENT SKID



NOTE: A-902 SHIPPED IN TWO PIECES.

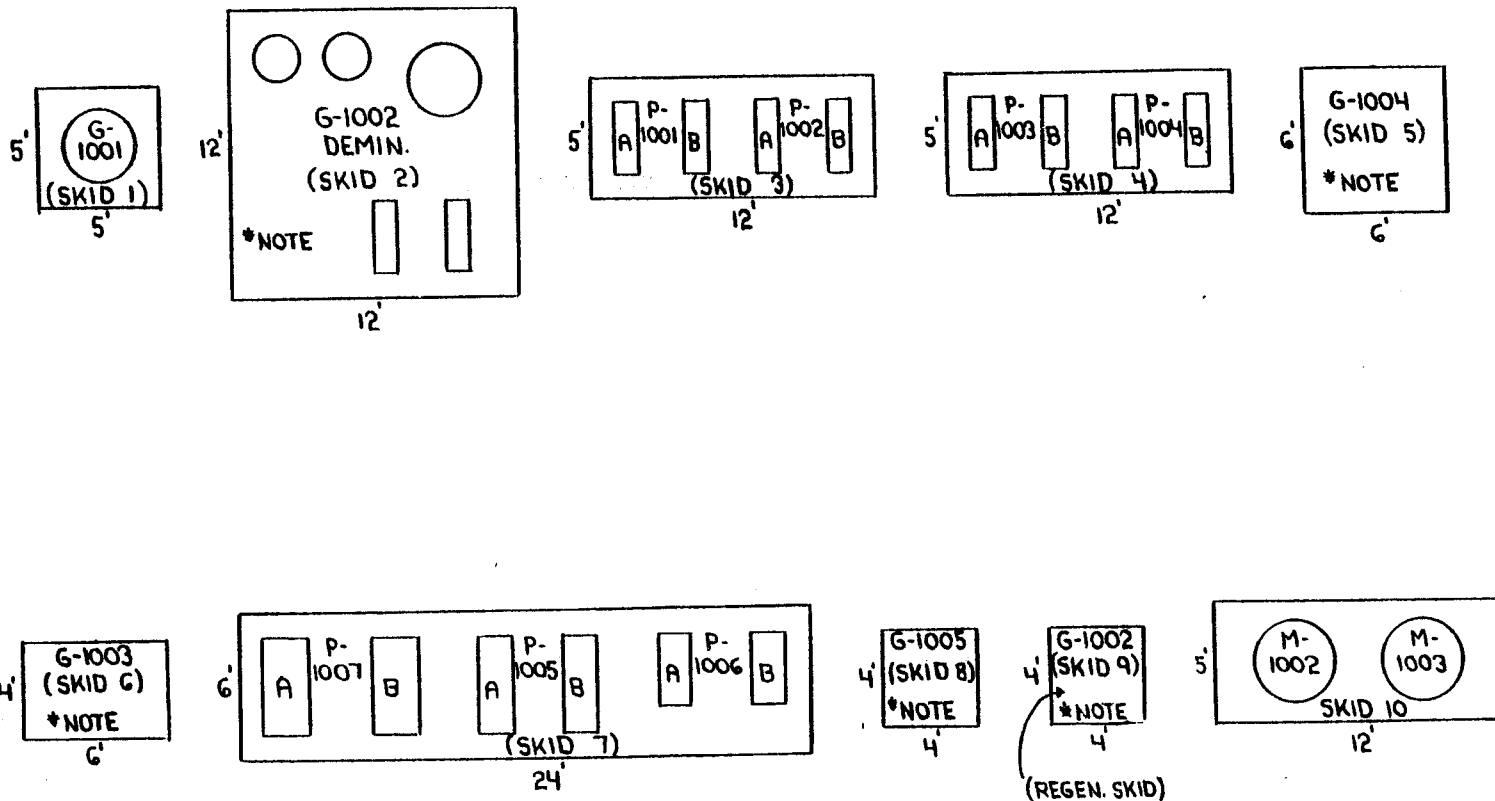
SCALE: $\leftarrow 5'-0'' \rightarrow$

5-125

				SCALE		STONE & WEBSTER ENGINEERING CORPORATION SERI METHANOL PLANT			
				DESIGN					
				CHECKED					
NO.	REVISION DESCRIPTION	BY	DATE	DRAWN	LEF	AREA	J.O. NO.	DRAWING NO.	ISSUE
	CERT. FOR FAB								
	CERT. FOR CONSY		APPROVED						
							1466500	FIG. 5-21	1

FIG. 5-22 1000 SECTION

PROCESS AND BOILER WATER TREATMENT AND STEAM PLANT SKIDS



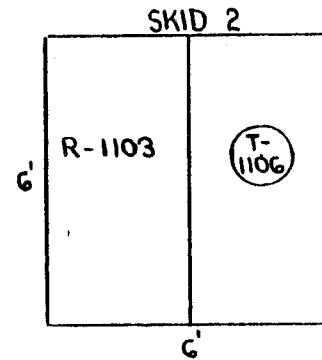
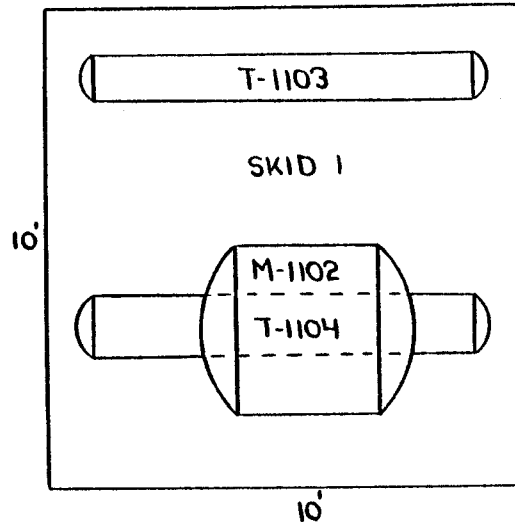
*NOTE: PACKAGE UNIT BY VENDOR

SCALE: 5'-0"

5-126

				SCALE		STONE & WEBSTER ENGINEERING CORPORATION			
				DESIGN		SERI METHANOL PLANT			
				CHECKED					
NO	REVISION DESCRIPTION	BY	DATE	DRAWN	LEF	AREA	I.O. NO.	DRAWING NO.	ISSUE
	CERT FOR FAB		CERT FOR CONST	APPROVED	CHECKED	14665.00		FIG. 5-22	1

FIG. 5-23
1100 SECTION
 GENERAL AREA SKIDS

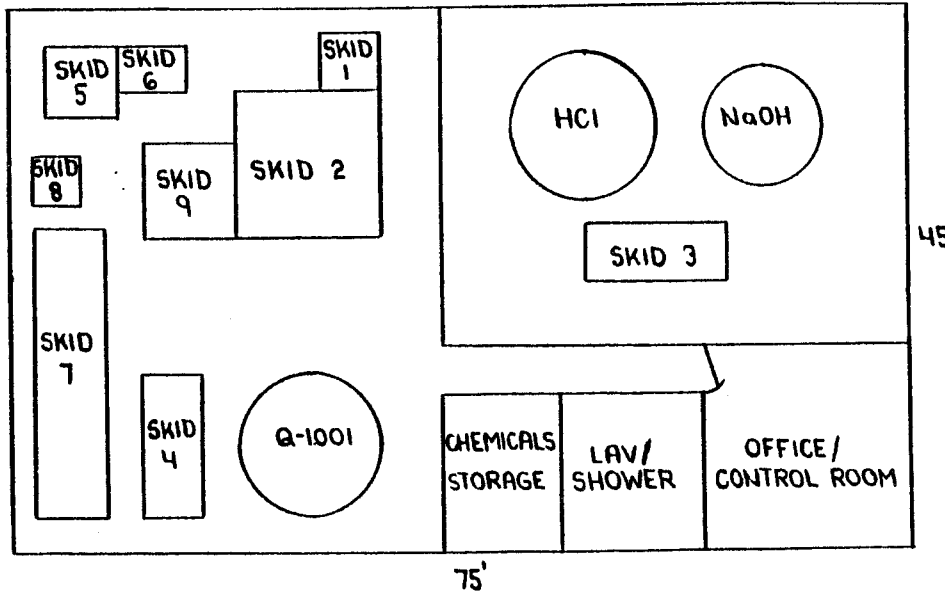


SCALE: $\leftarrow 5'-0'' \rightarrow$

5-127

				SCALE	STONE & WEBSTER ENGINEERING CORPORATION SERI METHANOL PLANT			
				DESIGN				
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NO	REVISION DESCRIPTION	BY	DATE	DRAWN	AREA	J.O. NO.	DRAWING NO.	ISSUE
	CERT FOR FAB		CERT FOR CONST.	LEF				
			APPROVED	CHECKED				
						14665.00	FIG. 5-23	1

FIG. 5-24
WATER TREATMENT BUILDING LAYOUT



SCALE : $\leftarrow 10'-0'' \rightarrow$

5-128

				SCALE		STONE & WEBSTER ENGINEERING CORPORATION			
				DESIGN		SERI METHANOL PLANT			
				CHECKED					
NO	REVISION DESCRIPTION	BY	DATE	DRAWN	AREA	J.O. NO.	DRAWING NO.	ISSUE	
	CERT FOR FAB		CERT FOR CONST	LEF			FIG. 5-24	1	
			APPROVED	CHECKED		14665.00			

FIGURE 5-25
 SERI METHANOL PLANT
 SECTION 100 - WOOD HANDLING
 ELEVATION VIEWING SOUTH

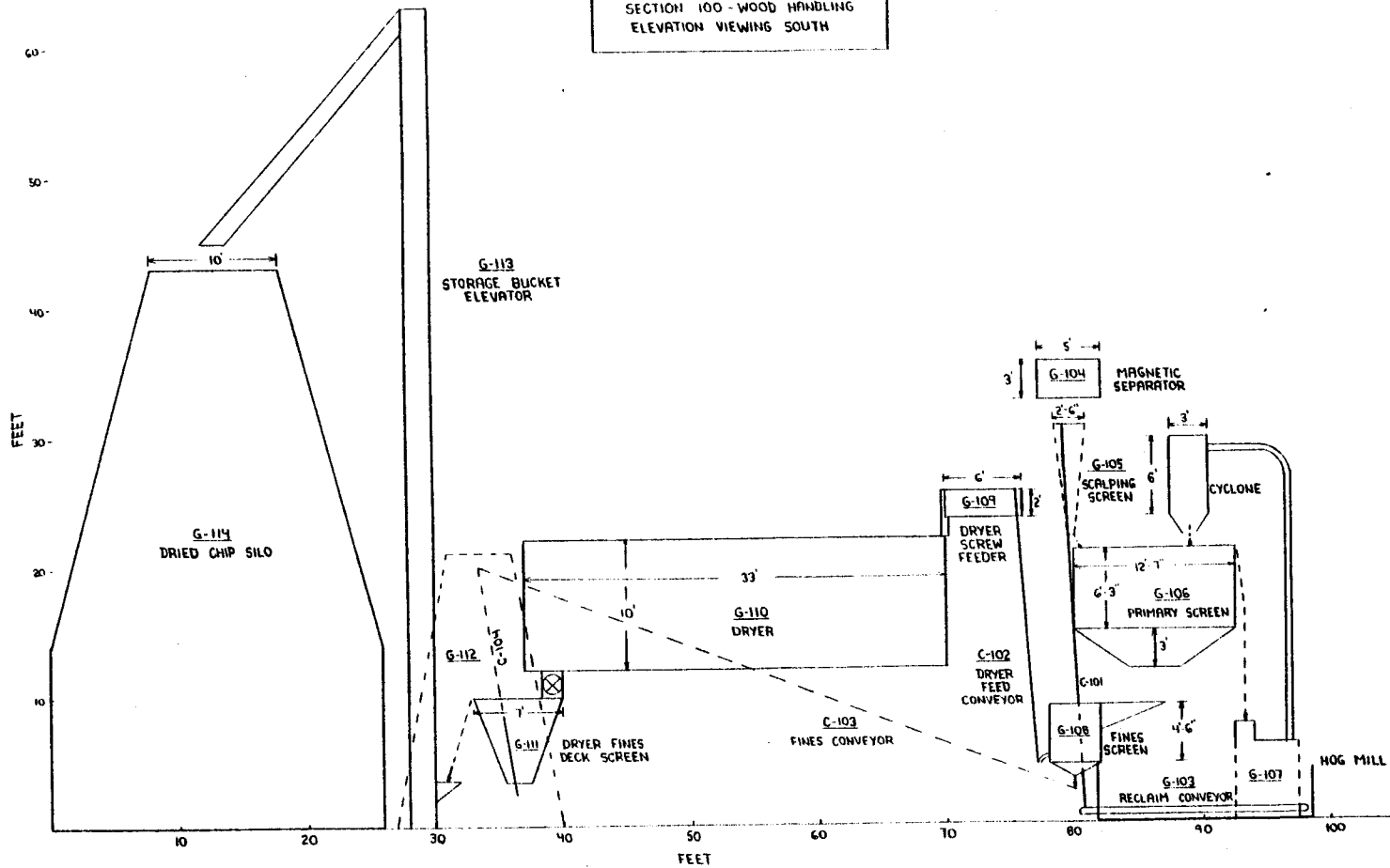
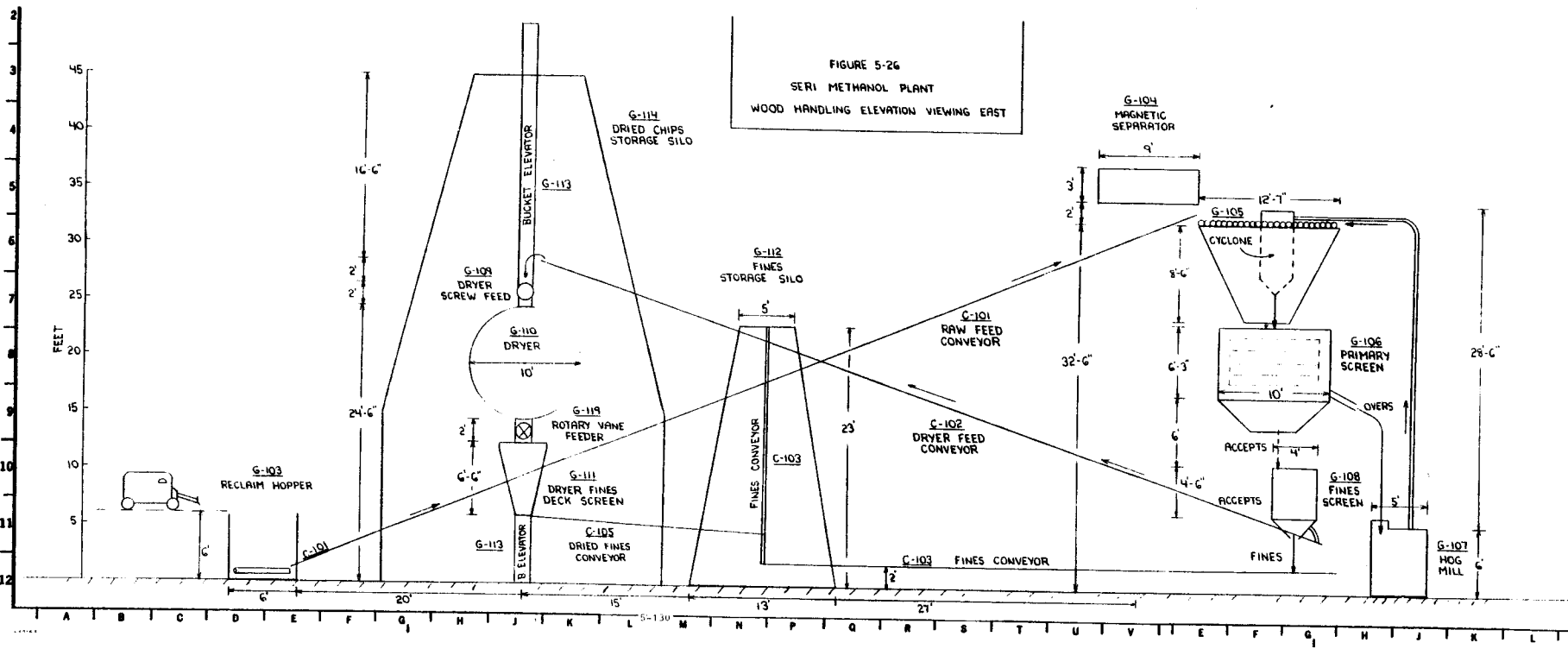
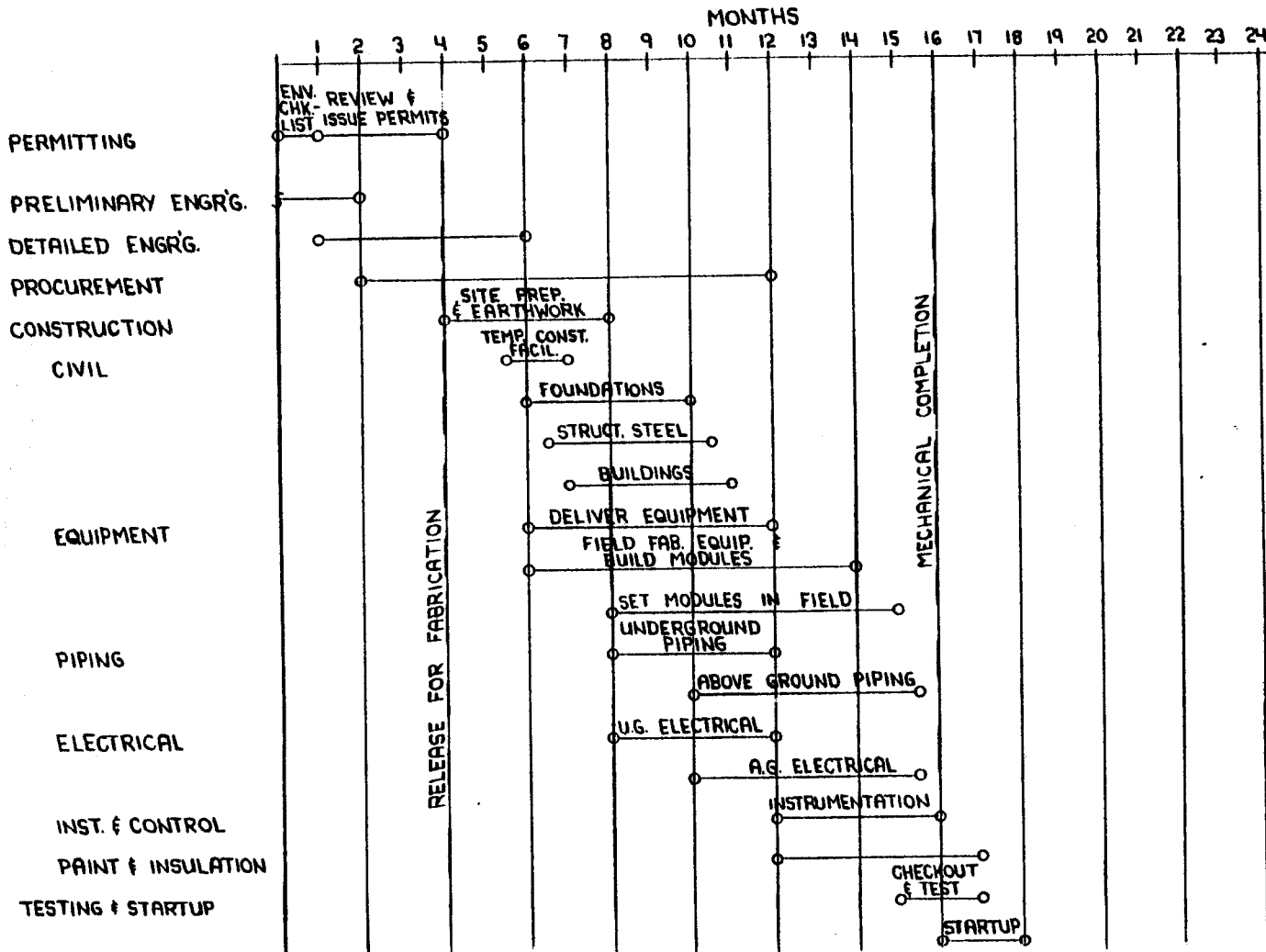


FIGURE 5-26
SERI METHANOL PLANT
WOOD HANDLING ELEVATION VIEWING EAST





5-132

NO.	REVISION DESCRIPTION	BY	DATE
	CERT FOR FAB		
	CERT FOR CONST		
	APPROVED		

FIGURE 5-28

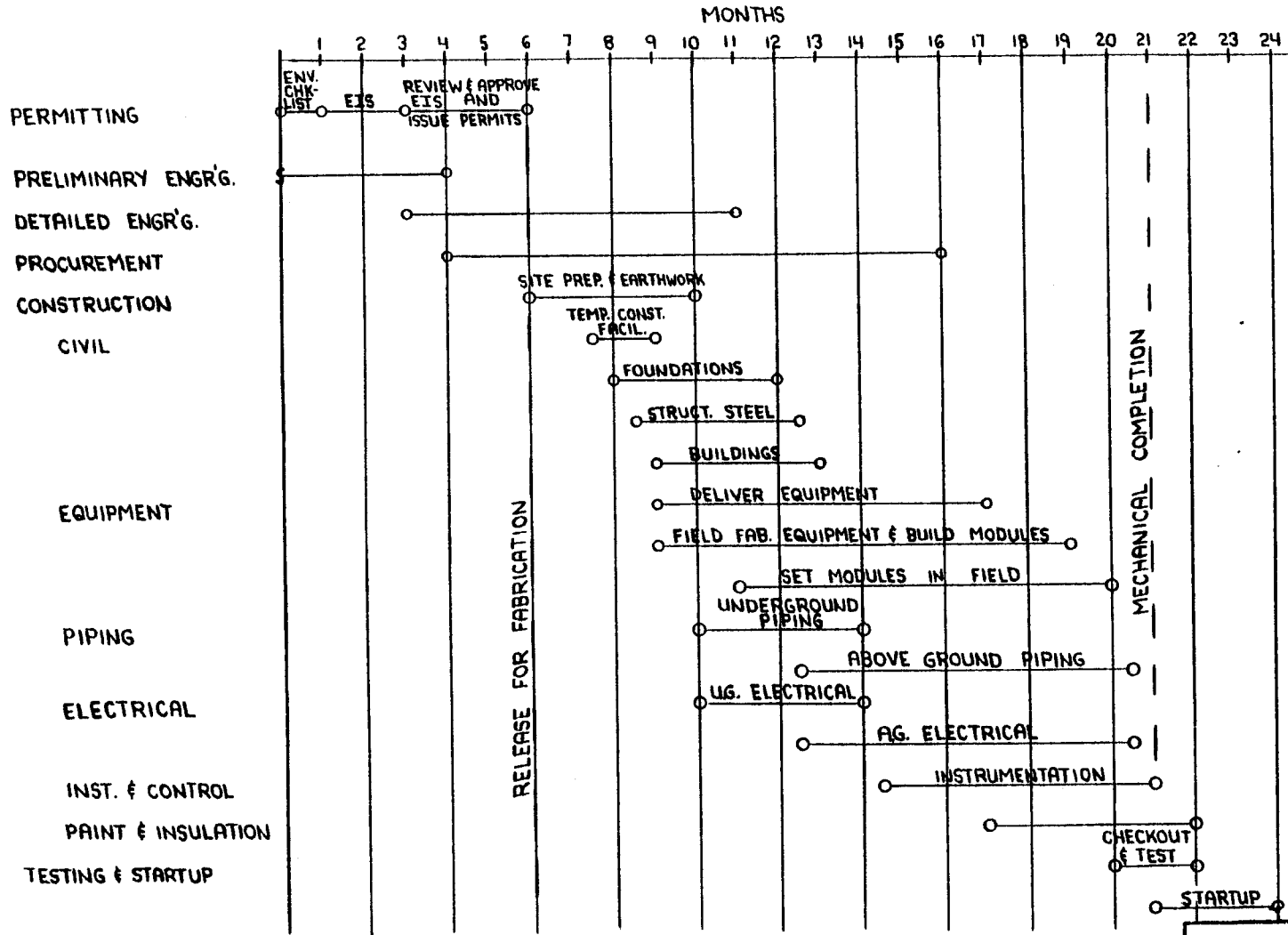
SRI METHANOL PLANT

PROJECT SCHEDULE

FOR "n" PLANT

STONE & WEBSTER ENGINEERING CORPORATION

J.O. 14665 SK



5-131

NO	REVISION DESCRIPTION	BY	DATE	SCALE
	CERT FOR FAB		APPROVED	DESIGN
	CERT FOR CONST		CHECKED	CHECKED
			DRAWN	LEF
			CHECKED	

FIGURE 5-21

SRI METHANOL PLANT

PROJECT SCHEDULE

FOR FIRST PLANT

STONE & WEBSTER ENGINEERING CORPORATION

JO. 14665
 SK-

SECTION 6
ECONOMIC ANALYSIS

The economic feasibility of the biomass based methanol plant (BMP) is presented based on the Nth plant sited in Spokane, Washington. The analysis includes the economic results of design improvements as well as a sensitivity analysis on critical parameters.

6.1 SUMMARY

The investment analysis of the biomass-based methanol plant (BMP) takes into consideration the variables in the operation of the plant, the price of raw materials and product selling price by performing a sensitivity analysis. The sensitivity analysis considers variation in operational availability (stream factor), raw materials cost (wood, oxygen, electricity), by-product credits, and capital costs. The total facilities investment is estimated to be \$22,153,000 including contingency.

The sensitivity analysis on the base case design is then extended by considering technical improvements in the design which improve the economics. In addition, changes in the method of financing the initial investment (i.e., the assumption of debt) are investigated.

The analysis is based on a constant dollar basis (1984 dollars), 100 percent equity financing and current prices for conventional methanol. It is recognized that the current supply/demand situation in the methanol market is resulting in a domestic price structure that is depressed from historical levels. It is unknown as to how long this pricing structure will persist, and the following analysis should be judged to be conservative.

It is assumed that the minimum acceptable after tax discounted cash flow rate of return (DCFROR) on a constant dollar basis is 15 percent. Taking full advantage of the Alcohol Fuel Tax Credit of 50¢/gal, the maximum selling price of the methanol product is \$1.16/gal in the Spokane area. This price results in the following range of DCFROR's (see Section 6.7.3.2).



<u>BMP (Case)</u>	<u>DCFRROR</u>
A) Pessimistic	11.1
B) Most likely	21.5
C) Optimistic	30.0

The required methanol selling price to achieve the minimum acceptable DCFRROR of 15 percent for these scenarios is:

<u>Case</u>	<u>Methanol Selling Price (¢/Gal)</u>
A) Pessimistic	131
B) Most likely	100
C) Optimistic	81

Figure 6-1 shows the DCFRROR vs methanol selling price for the base case. Changing the financing of the plant to 75 percent equity and 25 percent debt at real interest rates (net of inflation) of 4 and 8 percent result in required methanol selling prices as shown below for a DCFRROR of 15 percent, for Case B.

<u>Interest Rate</u>	<u>Methanol Selling Price (¢/Gal)</u>
4%	83
8%	86

The two design alternatives, doubling of the plant size and reforming of the purge gas result in the following required methanol selling prices, for Case B (at 100 percent equity, 15 percent DCFRROR).

Design AlternativeMethanol Selling Price (¢/Gal)

Doubling of the plant size	83
Reforming of the purge-gas	94

The analysis shows the need to utilize all or part of the Alcohol Fuel Tax Credit in order to attract investors to finance this type of facility. However, any rise in the price of domestic methanol in combination with debt financing and inclusion of the improvements identified in the alternative design cases would reduce the need to depend on the tax credit to attract investors to this concept.

Although the specific investment analysis is based on a Spokane, Washington site, other sites in the Tier III transportation region (see Figure 3-1) would be acceptable. Sites in other regions would become attractive as the price of methanol rises or under special circumstances, such as local incentives or inexpensive feedstocks.

6.2 ECONOMIC BASIS

The economic feasibility of the biomass based methanol plant (BMP) is based on the Nth plant in a series of identical BMP's. The only changes in the BMP's design would be dictated by site specific conditions, such as, climatic conditions, environmental requirements, and geotechnical limitations. A discounted cash flow (DCF) type of analysis is used to determine the rate of return available to investors for such a plant. In order to eliminate the need to estimate future inflation rates, the analysis is performed in constant 1984 dollars. The economic parameters for the BMP are given in Table 6-1.

The product methanol from the BMP will be competing with methanol that can be supplied from other domestic or international sources for the fuel/octane enhancer market. SWEC has assumed that methanol produced from the BMP can be sold at the U.S. Gulf Coast price plus a transportation differential. This analysis gives a current sales price of 66 cents per gallon (40¢/gal



plus 26¢/gal transportation). Additional revenue accrues to the facility through the Federal Alcohol Fuel Tax Credit. This credit is currently equivalent to 50¢/gal of alcohol produced from sources other than natural gas, coal, or peat, i.e., biomass. Therefore, the total sales price of methanol on a cash flow basis is \$1.16/gal. To keep the economic analysis on a more general basis, no state tax credits have been assumed. There are, however, several states which grant tax credits for fuel methanol.

6.2.1 Capital Cost Estimation Technique

The base case conceptual design of the BMP plant was costed to determine the base capital cost and to allow the economic evaluation to be performed. The capital costs for the BMP were determined from vendor quotes, cost data on file from previous engineering and construction projects, and estimates where direct determination was not possible (see Section 6.4). As discussed in Section 5, a maximum amount of equipment was shop fabricated to take advantage of higher labor productivity. The plant was assumed to be constructed by contractors using union labor. Wage rates have been supplied by the State of Washington for the Spokane area.

A contingency has been added to the capital cost estimate. The calculation of a plant contingency is made up of two factors. An Allowance for Indeterminates (AFI) is taken as a percentage of total materials and labor cost (not including subcontractor's fee and overhead) and is based on the known level of detail of the engineering design (e.g., the AFI goes down with a greater level of detail). The Process Development Allowance (PDA) is a method of recognizing the higher cost of designing and fabricating a first-of-a-kind piece of equipment and is applied to process units of the plant that are not presently available on a commercial level. In the BMP, only the gasifier is not a commercially proven unit. The other system to which a PDA is applied, because of limited commercial experience or a slightly different application, is the phenol extraction unit. The AFI has been set at 10 percent ($\$1.5 \times 10^6$) while the PDA is determined in accordance with the Gas Research Institute Procedure and is approximately \$200,000.



6.2.2 Economic Analysis Methods

The investment analysis of the biomass-based methanol plant (BMP) takes into consideration the variables in the operation of the plant, the price of raw materials and product selling price by performing a sensitivity analysis. The sensitivity analysis considers variation in operational availability (stream factor), raw materials cost (wood, oxygen, electricity), by-product credits, and capital costs.

The sensitivity analysis on the base case design is then extended by considering technical improvements in the design which improve the economics. In addition, changes in the method of financing the initial investment (i.e., the assumption of debt) are investigated. The base case considers 7,884 hours per year (90 percent stream factor) of plant operation. The plant has been designed to accommodate this level of operation via increased capacity in the solids handling areas and on-line sparing of critical process pumps.

6.3 ESTIMATED PRICE OF CONVENTIONAL METHANOL

The price of methanol produced from natural gas is presented for comparison purposes with the selling price of BMP methanol required to meet the desired DCFROR. The prices predicted in this section assume a cost of methane as projected by the DOE. The current market conditions of a worldwide methanol capacity surplus have resulted in methanol prices below that predicted by this analysis. At this time, domestic producers are covering only production costs. It is unknown at what point in time that these market conditions will return to a more normal historic level (if ever).

Table 6-2 shows the projected prices for natural gas used for industrial purposes as compiled by the Energy Information Agency. These prices were used to predict the selling price of methanol (Table 6-2) f.o.b. on the U.S. Gulf Coast from an existing methanol plant. The plant is assumed to come on-line in January 1984. The methanol price estimate is based on a methanol from natural gas facility using LURGI technology and producing 2,000 tons per day of methanol, with a 25 percent Rate of Return, before taxes.



6.4 CAPITAL COSTS

The capital costs of the Bimoass-Based Methanol Plant (BMP) are shown in Table 6-3. The total cost is \$22,153,000. The costs have been divided into materials and equipment, field installation labor, freight and tariffs, engineering and construction management, initial charge of catalysts and chemicals, interest during construction, and working capital. The working capital is based upon 7 days' storage of both raw materials and finished product plus 0.9 percent of the total base cost for spare parts and miscellaneous materials. The basis for the cost estimate is the base case engineering design performed as a part of this study. This cost estimate is based on the installation of the Nth plant in a series of BMPs and the skid-mounted nature of a majority of the equipment. The Nth BMP is assumed to be constructed over an 18-month schedule. During this period, interest during construction is calculated using an S curve (for the expenditure of funds) at a real interest rate (net of inflation) of 8 percent.

The construction of the first plant of this type would result in an increased capital cost of approximately $\$2.0 \times 10^6$. The increases result from the following areas:

- Additional one-time engineering required for the initial modular design and first of a kind project.
- Additional startup time for the first of a kind project.
- Increased financing costs due to a longer project schedule.

The cost of installation is based upon the prevailing union wage rates in the Spokane, Washington vicinity. The Spokane site is located in Tier III, a favorable region with respect to methanol transportation costs (see Section 3). Other potential site locations in Tier III included Salt Lake City, Utah, and Albuquerque, New Mexico. Table 6-4 shows the differential in base labor rates between the three potential Tier III sites. The potential savings in construction cost are shown in Table 6-5. The effect



of this capital cost reduction on the rate of return and how other local parameters (availability of raw materials) related to these costs are discussed in Section 6.7.2.

6.5 REVENUES

The revenues from the BMP come from two sources, the sale of methanol and the sale of carbon dioxide. In general, raw carbon dioxide can be sold to a distributor for purification and sale for approximately \$10/ton. Although there is not a sufficient market in Spokane to justify the sale of this product from the BMP, the effect of this revenue on plant economics for other sites is discussed in Section 6.7.

The selling price of the methanol product is calculated from three independent factors: the current U.S. Gulf Coast selling price, the transportation costs to Spokane, and the Federal Alcohol Fuel Tax Credit (Table 6-6). For the base case analysis it is assumed that the full effect of the Fuel Alcohol Tax Credit is reflected in the selling price of the methanol. The methanol from the BMP is eligible for the tax credit since the feedstock is biomass. At the present time (June 1984), the domestic methanol price is low due to excess methanol production capacity. The potential effect of an increase in the U.S. Gulf Coast price on the economics of the BMP is discussed in Section 6.7.

6.6 OPERATING AND MAINTENANCE COSTS

The operating costs for the BMP include wood, oxygen, manpower, electricity, chemicals, catalysts, water, fuel, and waste disposal costs. Tables 6-7 to 6-10 enumerate these various costs for the BMP. The manpower costs include 76 1/2 percent of wages for overhead. The overhead costs were specifically calculated for a small plant and include such items as taxes, benefits, security, janitorial services, sales cost, office supplies, payroll, accounting and legal services.

Maintenance costs are calculated as a percentage of plant section base cost. Table 6-11 gives the maintenance factor by plant section. Insurance and local taxes are taken as 1.5 percent of the BMP's total base cost. Table 6-12 is a summary of the annual operating and maintenance costs.

6.7 INVESTMENT ANALYSIS

The basis for the economic analysis is presented in Section 6.2. The projection of conventional methanol selling prices (Table 6-2) shows a steady increase for the projected period from present day costs. A majority of this cost increase is due to the projected price of natural gas. Current (June 1984) selling prices of methanol are substantially below the prices that would be projected using a reasonable rate of return on investment. This shows that current market conditions are forcing domestic methanol producers to absorb certain costs of production. It is unknown how long this situation will exist. The projected price increases established in Section 6.3 may not occur as stated or may exceed these rates if market conditions improve. This uncertainty in projected prices for conventional methanol directs the analysis to be conducted in constant 1984 dollars with the product price set at stated in Section 6.2 at \$1.16 per gallon. The effect of changes in this selling price is discussed below.

The minimum rate of return that would be acceptable to an investor is dependent upon the level of risk that is perceived. Two components of risk are apparent in this project:

- inability to sell the methanol project, and
- potential problems in operation.

The inability to sell the methanol product is of prime concern, since the technological problems would be known from the pilot plant and operation of previous plants. The risk in selling the product is concerned with market penetration and the availability of co-solvents. In the context of the constant dollar nature of this analysis and the present level of real rate

of return obtainable on secure debt securities (approximately 8 percent), an after tax DCFROR of 15 percent should be considered the minimum return that an investor would expect for this project.

6.7.1 Methanol Selling Price

The variation of selling price with DCFROR is shown in Figure 6-1. The minimum price required to achieve the desired DCFROR of 15 percent is seen to be \$1.00/gal. In the Spokane area, with the assumption of full tax credits (50¢/gal) over the life of the project, this minimum selling price is exceeded by 16¢/gal. A methanol selling price of \$1.16/gal results in a DCFROR (after tax) of 21.5 percent for the base case. It should be recognized that the elimination of the tax credit or the inability to use at least 68 percent of the credit (34¢/gal) will result in a BMP (in Spokane) with an insufficient return to attract investors (assuming a minimum DCFROR of 15 percent). The portion of the methanol selling price which is assumed to be contributed by the Federal tax credit may be contributed via price increases in conventional methanol, state subsidies for biomass based methanol, or higher transportation cost for conventional methanol.

The transportation costs to other areas are outlined in Section 3. The required methanol selling price in these areas is not simply based on transportation differentials, but must include other changes in costs such as wood, oxygen, electricity and by-product credits. The sensitivity analysis will identify the value of these several parameters.

6.7.2 Sensitivity Analysis

The change in required methanol selling price with variations in parameters such as plant capital cost, raw materials prices, plant operation, and by-product credits requires a sensitivity analysis to be performed in order to determine the range of potential rates of return. The effect of these parameters on the minimum methanol selling price to meet the required 15 percent DCFROR is also discussed. This information can also be used to evaluate sites other than Spokane, if the price of raw materials and the saleability of by-products are known.



6.7.2.1 Capital Costs

Figure 6-2 shows the effect of a change in capital cost on the DCFROR. A 1 percent change in DCFROR is caused by a change of approximately \$900,000 in capital costs. This change in capital costs is equivalent to a change of 2.75¢/gal of methanol selling price. One area of potential capital cost reduction is in field (installation) labor. As shown in Section 6.4, moving the site to Salt Lake City or Albuquerque would reduce the capital costs by approximately \$610,000 and \$820,000, respectively. In addition, if labor productivity is increased by 20 percent (i.e., nonunion labor) the capital costs of base case plant could be reduced by approximately \$900,000.

Variations in capital cost of approximately 10 percent ($\$2.1 \times 10^6$) change the DCFROR calculated by 2.3 percent and are equivalent to approximately a 6.3¢/gal change in methanol selling price.

The base case economics assumes the design and construction of the Nth plant. As stated in Section 6.4, the capital cost of the first plant would increase by approximately \$2.0 MM. This would decrease the DCFROR by approximately 2.2 percent and increase the minimum methanol selling price by approximately 6.0¢/gal.

6.7.2.2 Stream Factor

The sensitivity of the DCFROR to the on-line availability of the BMP (stream factor) to produce methanol is shown in Figure 6-3. The base case assumes that the plant operated 7,884 hours per year or 90 percent of the available time. A reduction in availability to 70 percent produces a DCFROR of below the 15 percent minimum required for investor participation in this venture. A 10 percent change in availability results in a 3.4 percent change in DCFROR. This corresponds to approximately a 9¢/gal change in methanol selling price. This calculation demonstrates the critical nature of the availability of the plant to return on investment.



6.7.2.3 Raw Material Costs and By-Product Credits

The sensitivities of raw material prices, i.e., wood, oxygen, electricity, and of by-product revenue, e.g., carbon dioxide, can be analyzed as direct changes in methanol selling price. Table 6-13 gives the results of a unit price change in each of these commodities. Any change in revenues of \$76,800 is equivalent to 1¢/gal in methanol price. These price changes can be related back to the DCFROR calculation through Figure 6-1. Figures 6-4 to 6-7 shows the DCFROR sensitivities to changes in the price of wood, electricity, oxygen, and carbon dioxide. Because of the slightly nonlinear nature of the DCFROR vs equivalent price curve (Figure 6-1), a 1 percent change in DCFROR is caused by a change in equivalent methanol price ranging from 2.48¢/gal to 2.74¢/gal in the range of concern (i.e., DCFROR 15-20 percent).

6.7.3 Discussion of Results

6.7.3.1 Alternative Sites

The sensitivity analysis can be used to estimate the appropriateness of an alternative site. For example, the ability to sell carbon dioxide at \$10/ton can offset an increase in wood costs of approximately \$7.40/ton, or can reduce the amount of revenue required from the tax credit by 4.9¢/gal, or can increase the DCFROR by 1.9 percent. In conjunction with the transportation cost map (Figure 3-1), a quick approximation can be made as to what additional sites are viable or what magnitude of incentives are required to establish a BMP in a specific location. For example, a BMP sited in the State of Hawaii with electricity priced at 7¢/kWh ($\Delta = +3\text{¢/kWh}$) and wood costing \$37/BDT ($\Delta = \7/BDT, and transportation costs of 20¢/gal ($\Delta = -6\text{¢}$) would require state incentives equivalent to 15.35¢/gal to maintain base case DCFROR (21 1/2 percent).

The sensitivities can also be used to evaluate the potential of a new feedstock. For example, rice hulls can currently be obtained at zero cost ($\Delta = -\$30\text{/ton}$) in the Gulf Coast region. This area also has oxygen



available at approximately \$40/ton ($\Delta = -\$12.82/\text{ton}$), and a ready market for carbon dioxide ($\Delta = +\$10/\text{ton}$). This combination of revenues and material savings of approximately 24¢/gal offsets the transportation differential to Spokane of 26¢/gal. This quick analysis indicates that rice hulls may make a reasonable alternative feedstock to test in the gasifier.

6.7.3.2 Combinations of Parameters

The sensitivity analysis has considered the effect of individual parameters on the economics of the plant. The question arises as to the effect of a combination of parameters of the site specific case of Spokane, Washington. Two cases are considered, a pessimistic case and an optimistic case. The pessimistic case considers a capital cost of \$24,000,000 (+10 percent), a stream factor of 70 percent (-20 percentage points), and a wood price of \$40/ton (+\$10/ton). The optimistic case considers a capital cost of \$20,000,000 (-10 percent), a stream factor of 95 percent (+5 percentage points), a wood price of \$20/ton (-10/ton), and the sale of CO₂ at \$10/ton (+\$10/ton).

The pessimistic case would occur with a tight supply of wood, some plant operational difficulties, and capital cost within the accuracy of the current estimate. This scenario results in a DCFROR of 11.1 percent at the methanol selling price of \$1.16/gal. This rate of return is below the minimum acceptable level even considering the full use of the Alcohol Fuel Tax Credit.

The optimistic case may occur by obtaining wood, establishing a new market for carbon dioxide, not encountering plant operational problems, and reducing the capital cost within the accuracy of the current estimate. This scenario results in a DCFROR of 30 percent at the methanol selling price of \$1.16/gal. This case allows the methanol selling price to be 81¢ at 15 percent DCFROR which requires the use of only 30 percent (15¢/gal) of the Alcohol Fuel Tax Credit.

6.7.3.3 Alternative Designs

The base case design of the BMP has defined a starting point in the economic analysis of the concept of a small-scale methanol production facility. Several technical changes to this design have been investigated in order to gain additional insight into plant operating conditions that can improve the overall economics. The details of the design changes are discussed in Section 7. Of the three design variations on the base case (higher gasifier operating pressure, higher wood moisture content, and reforming the purge gas methane) only the case of reforming the purge gas showed any significant improvement in the economics.

The alternative design of reforming the purge gas increases the methanol production to 8.77 MM gal/yr from 7.68 MM gal/yr. This design also increases total capital expenditures by approximately \$1,000,000, but reduces the methanol selling price by 8¢/gal to \$1.08/gal to achieve the base DCFROR (21 1/2 percent). To achieve the minimum 15 percent DCFROR, the methanol selling price would be reduced to 94¢/gal, thus requiring the use of only 56 percent of the current tax credit. This increase in production can easily be sold in the local market without any significant change in the market penetration assumptions. This design alternative would be recommended to be incorporated in any revision to the base case design.

The final design alternative investigated was to double the gasifier throughput and thus the methanol output. This case is also discussed in detail in Section 7. It is necessary to recognize that at a production rate of 15.36 MM gal/yr of methanol, a portion of the product may have to be shipped a considerable distance to market and thus reduce the revenue received by the BMP.

For this particular analysis the assumption of local sales will be retained. The reduction in required methanol selling price to maintain the base DCFROR of 21 1/2 percent is approximately 21¢/gal. This is due to economies of scale in both the BMP and the oxygen plant. The methanol selling price



required to achieve the minimum 15 percent DCFROR is reduced to approximately 83¢/gal. This price can be achieved by utilization of only 34 percent of the Federal Alcohol Tax Credit. This calculation demonstrates the importance of accurately defining both the gasifier throughput and the local market in order to size the BMP properly.

6.7.3.4 Debt Financing

The foregoing analysis was based on the assumption of total equity financing for the BMP. The investment in the BMP can be partially financed by debt. A reasonable level of debt for this type of facility to be financed by a long term (20 year) bond would be 25 percent. The interest rate on the debt is based on a zero inflation rate, since the economic calculations are performed in constant 1984 dollars. The historical real rate of return (net of inflation) on secured debt is 4 percent although the current rate is approximately 8 percent. The DCFROR for these two cases, at a methanol price of \$1.16/gal, are:

<u>Equity</u>	<u>Debt</u>	Net <u>Interest Rate</u>	<u>DCFROR (%)</u>
75%	25%	4%	36.01
75%	25%	8%	34.21

The methanol selling price required to obtain the minimum 15 percent return on equity is approximately 86¢/gal for the 8 percent interest case and 83¢/gal for the 4 percent interest rate case. Both of these price levels require use of the Alcohol Fuel Tax Credit (40 percent and 34 percent, respectively), to maintain investor interest.

TABLE 6-1

BASIS FOR INVESTMENT ANALYSIS^(a)

<u>Component</u>	<u>Amount</u>
Plant Life	20 years
Operating Factor	7,884 hr/yr (90%)
Equity	100%
Minimum Return on Equity Required	15% DCFROR (after tax)
Interest Rate During Construction	8% real
Methanol Production Rate	7.68 MM gal/yr
Depreciation	ACRS (13.5%, 19.8%, 18.9%, 18.9%, 18.9%)
Tax Rate	50%
Investment Tax Credit	10%
Renewable Energy Tax Credit	10%
Alcohol Fuel Tax Credit	50¢/gal
Construction Time	18 months
Wood Cost	\$30/BDT
Oxygen Cost	\$52.82/ton ^(b)

(a) The investment analysis is based on a discounted cash flow rate of return (DCFROR) in constant 1984 dollars. The DCFROR analysis assumes that the ITC and RETC tax credits are used to reduce the first year capital investment. The five (5) year accelerated cost recovery system method is applied to the total depreciable plant and the DCFROR calculation is applied on a beginning of the year basis. The cost of capital (Rate of Return) and interest rates is net of inflation.

(b) The cost of oxygen is based on a 60 TPD dedicated facility operated by an independent vendor, with the methanol plant having a take-or-pay contract.



TABLE 6-2

PROJECTED PRICES

<u>Commodity</u>	<u>1985</u>	<u>1986</u>	<u>1987</u>	<u>1988</u>	<u>1989</u>	<u>1990</u>
Natural Gas (\$/MM Btu) ^(a)	5.62	6.45	7.00	7.13	7.17	7.23
Methanol (¢/gal) ^(b)	65.9	71.3	75.0	76.3	77.0	78.1

(a) From "1982 Annual Energy Outlook" DOE/EIA 0383, 82, in constant dollars.

(b) Assuming an existing 2,000 TPD conventional methanol plant at 25 percent return on investment before taxes.

TABLE 6-3

BIOMASS BASED METHANOL PLANT

Capital Cost - Base Case

<u>Component</u>	<u>Cost</u>
Material and Equipment	\$11,354,900
Field Labor (including Contractor's Fee)	4,610,000
Freight and Tariffs	<u>189,100</u>
Total Base Cost	\$16,154,000
Land Cost	30,000
Engineering and Construction Management	1,870,000
Contingency and Process Development Allowance	<u>\$ 1,700,000</u>
Total Installed Cost	\$19,754,000
Initial Catalysts and Chemicals	207,000
Startup Expenses (2 months)	751,000
Interest During Construction	1,156,000
Working Capital	<u>285,000</u>
Total Facilities Investment	\$22,153,000



TABLE 6-4

DIFFERENTIAL OF LABOR RATES IN SELECTED TIER III CITIES

<u>Category</u>	<u>Spokane</u>	<u>Salt Lake City</u>	<u>Albuquerque</u>
Pipefitters	Base = 1.00	0.74	0.78
Ironworkers	Base	0.90	0.82
Electricians	Base	0.90	0.84
Laborers	Base	0.81	0.68
Cement Finishers	Base	0.81	0.75

TABLE 6-5

POTENTIAL SAVINGS IN CONSTRUCTION COSTS DUE TO LABOR COST DIFFERENTIALS

<u>Site</u>	<u>Savings</u>
Spokane	Base = 0
Salt Lake City	\$610,000
Albuquerque	\$820,000

TABLE 6-6

METHANOL SELLING PRICE

<u>Component</u>	<u>Value (¢/Gal)</u>
U.S. Gulf Coast Price	40
Transportation Costs to Spokane	26
Alcohol Fuel Tax Credit	<u>50</u>
Equivalent Methanol Selling Price	116

Methanol Revenue $(7.68 \times 10^6 \text{ gal/yr}) \times (\$1.16) = \$8,908,800/\text{yr}$



TABLE 6-7

PLANT INPUT OPERATING COSTS

<u>Component</u>	<u>Unit Cost</u>	<u>Amount (per year)</u>	<u>Cost (\$)</u>
Wood	\$30/ton (dry)	102,295 tons	1,534,424
Oxygen ^(a)	\$52.82/ton	19,111 tons	1,009,443
Electricity	4.07¢/kWh	4.346 MM kWh	176,894

(a) Oxygen cost is made up of \$36.53/ton for production and \$16.29/ton for electricity.



TABLE 6-8

CHEMICALS AND CATALYSTS

<u>Component</u>	<u>Unit Cost</u>	<u>Amount (per year)</u>	<u>Cost (\$)</u>
50% NaOH	\$230/ton	35.48 tons	8,160
30% HCL	\$105/ton	23.65 tons	2,483
Sodium Sulfite	\$0.22/lb	6,718 lb	1,478
30% NH ₃	\$0.21/lb	28 lb	6
Sodium Phosphate	\$0.25/lb	9,655 lb	2,414
100% Sulfuric Acid	\$80.30/ton	2.31 tons	185
15% Sodium Hypochlorite	\$0.55/gal	236 gal	130
Isopropyl Ether	\$0.37/lb	40,000 lb	14,800
Shift Catalyst	\$216/ft ³	42 ft ³	9,072
Guard Absorber Catalyst	\$135/ft ³	171 ft ³	23,085
Methanol Catalyst	\$478/ft ³	69.3 ft ³	33,128
Benfield Solution	\$1.65/gal	78,840 gal	130,086
Sponge Iron	\$7/Bushel	1371.4 bushel	9,600
Ion Exchange Resin	\$160/ft ³	10 ft ³	1,600
Activated Carbon	\$30/ft ³	11.3 ft ³	340
			<u>\$236,567</u>



TABLE 6-9

MISCELLANEOUS OPERATING COSTS

<u>Component</u>	<u>Unit Cost</u>	<u>Amount (per year)</u>	<u>Cost (\$)</u>
Water	\$0.28/100 ft ³	1,473,877 ft ³	4,127
Nitrogen ^(a)	\$3/ton	6,701 tons	20,103
Propane	\$0.50/gal	29,520	14,760
Sewer Charge	\$0.357/100 ft ³	2,284,200 ft ³	8,148
Solid Waste Disposal	\$20/ton	1,183 tons	23,660
Diesel Fuel	\$1/gal	9,000 gal	9,000
General Supplies	--	--	<u>20,000</u>
			\$99,798

(a) Available from the oxygen plant



TABLE 6-10
MANPOWER COSTS

<u>Personnel</u>	<u>Hr/Yr</u>	<u>\$/Hr</u> ^(a)	<u>Total Cost (\$)</u>
Control Room Operators	7,884	11.00	86,724
Process Laborers	13,140	11.00	144,540
Wood Pile Operators	7,884	9.38	73,952
Shipper/Receivers	2,628	6.64	17,450
Lab Technicians	2,628	7.50	19,710
Clerk	2,080	6.21	12,917
Manager	2,628	20.00	<u>52,560</u>
			407,852
	Overhead at 76 1/2%		<u>312,007</u>
	Total		\$719,859

(a) Base salary 1984\$

TABLE 6-11

MAINTENANCE FACTOR BY PLANT SECTIONS

<u>Plant Section</u>	<u>Maintenance Factor (% of Capital)</u>
100 Wood Handling and Preparation	6.0
200 Gasification	6.0
300 Shift	4.0
400 Gas Cooling and Compression	5.0
500 Acid Gas Removal	4.0
600 Methanol Synthesis	3.0
700 Methanol Purification and Storage	3.0
800 Methanol Purge Gas Treatment	2.0
900 Wastewater Treatment	3.0
1000 Steam Plant and Process Water Treatment	2.0
1100 Utilities and Buildings	2.0
1200 General Facilities	2.0



TABLE 6-12
BMP ANNUAL OPERATING COSTS SUMMARY^(a)

<u>Component</u>	<u>Cost (\$)</u>	<u>Methanol Component Cost ¢/gal</u>
Wood (\$30/ton)	1,534,424	19.98
Oxygen (\$52.82/ton)	1,009,443	13.14
Electricity (4.07¢/kWh)	176,894	2.30
Catalysts and Chemicals	236,567	3.08
Miscellaneous	99,798	1.30
Labor	720,000	9.38
Maintenance	507,000	6.60
Insurance and Taxes	<u>242,310</u>	<u>3.16</u>
	4,526,436	58.94

(a) Methanol production rate is 7.68×10^6 gal/yr

TABLE 6-13

METHANOL SELLING PRICE CHANGES DUE TO COMMODITY PRICE CHANGES

<u>Commodity</u>	<u>Unit Price Change</u>	<u>Equivalent Price Change (¢/gal)</u>
Wood	+\$10/dry ton	+6.66
Carbon dioxide	+\$10/ton	-4.91
Oxygen	+\$10/ton	+2.49
Electricity ^(a)	+1¢/kWh	+1.56

(a) The change in electricity includes electricity for oxygen production at 0.995¢/gal and electricity for the BMP at 0.566¢/gal.



FIGURE 6-1

METHANOL SELLING PRICE VS. DCFROR

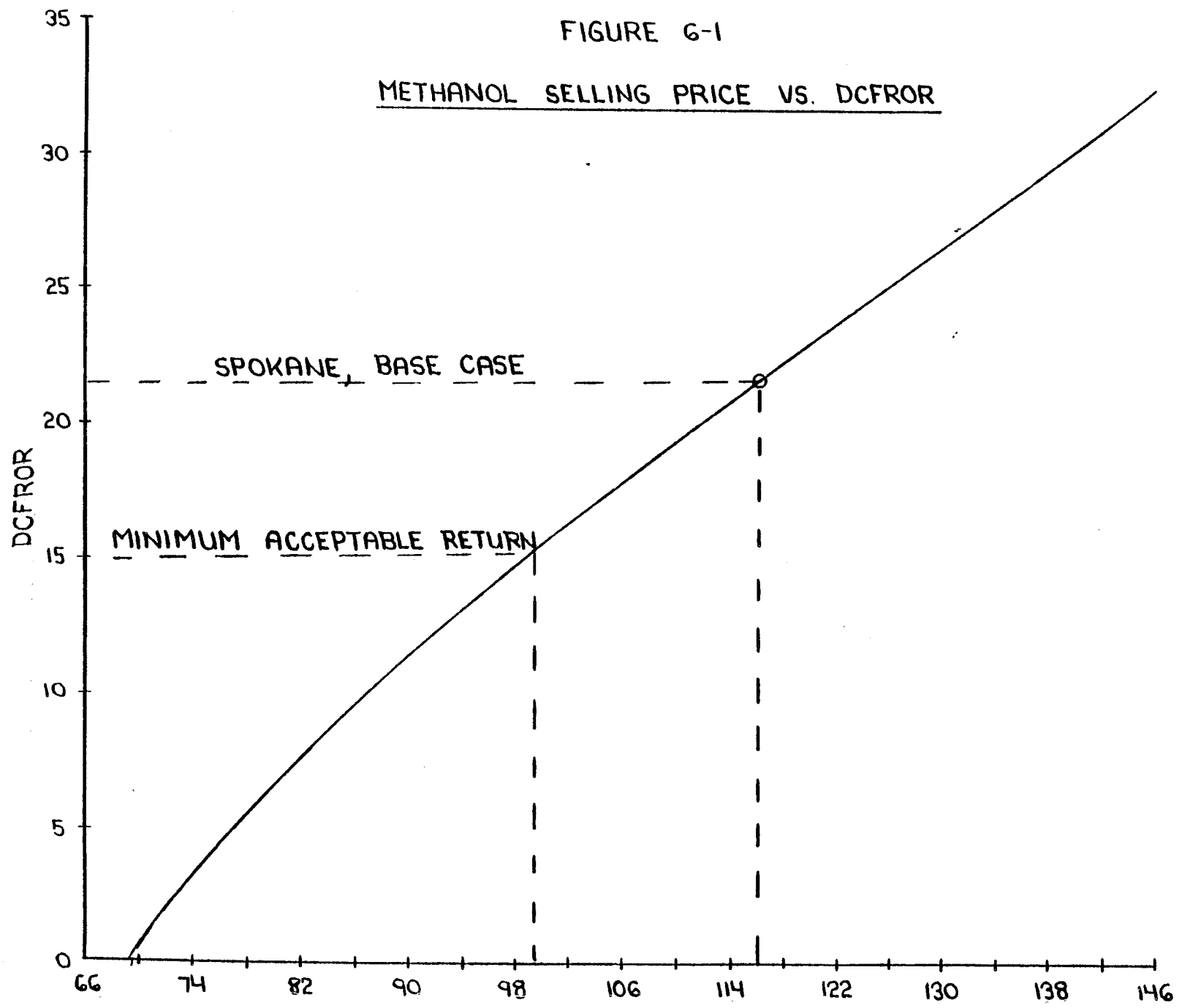


FIGURE 6-2
DCFRR VS. CAPITAL COST

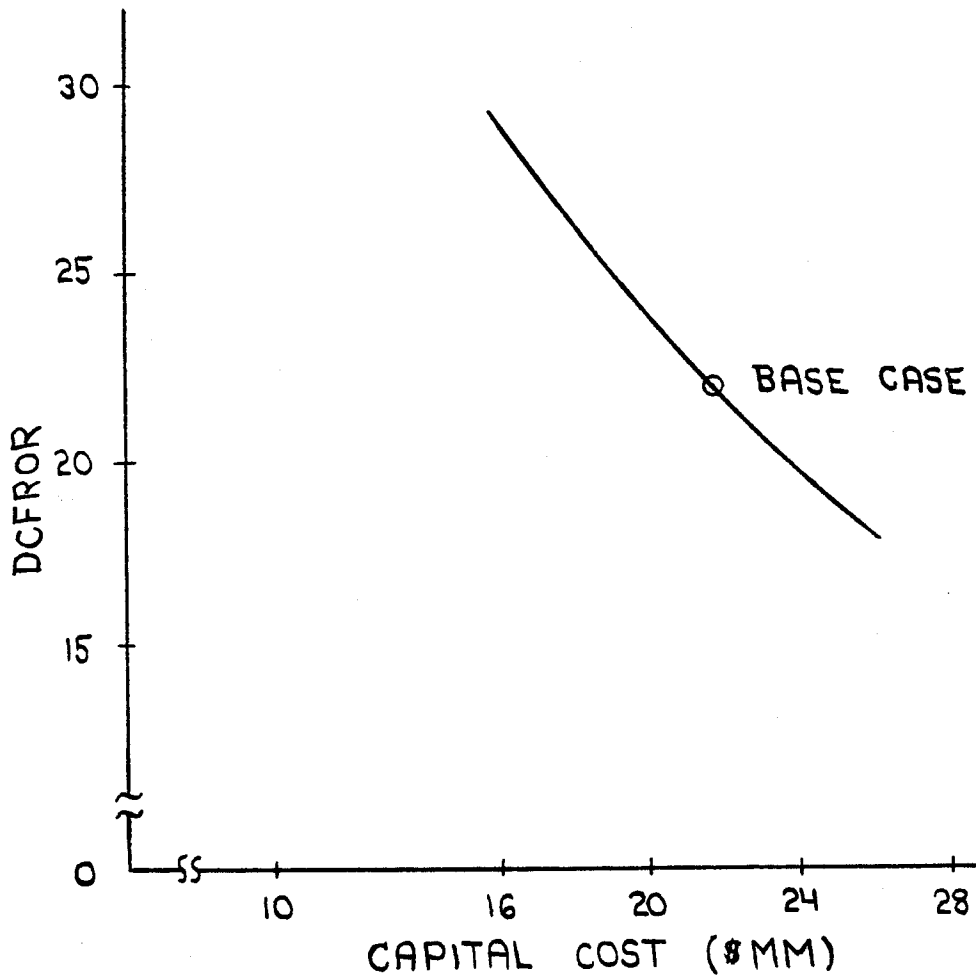


FIGURE 6-3
DCFROR VS. STREAM FACTOR

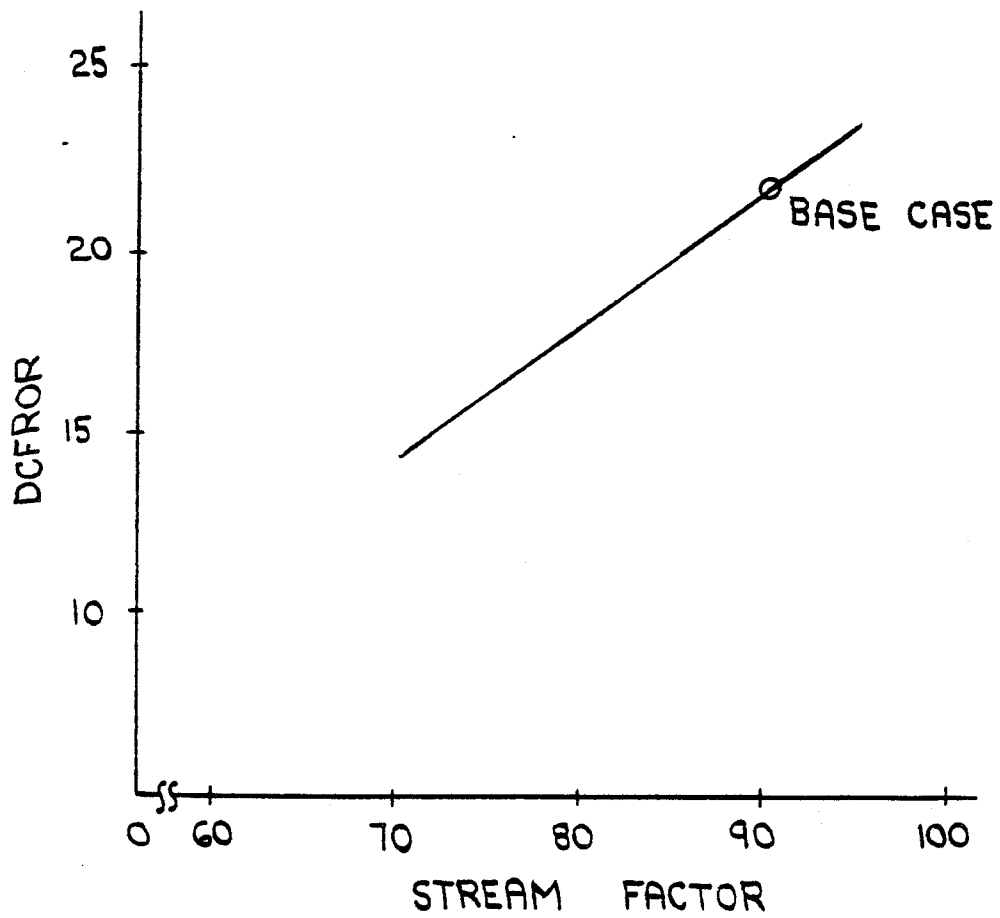


FIGURE 6-4
DCFROR VS. WOOD PRICE

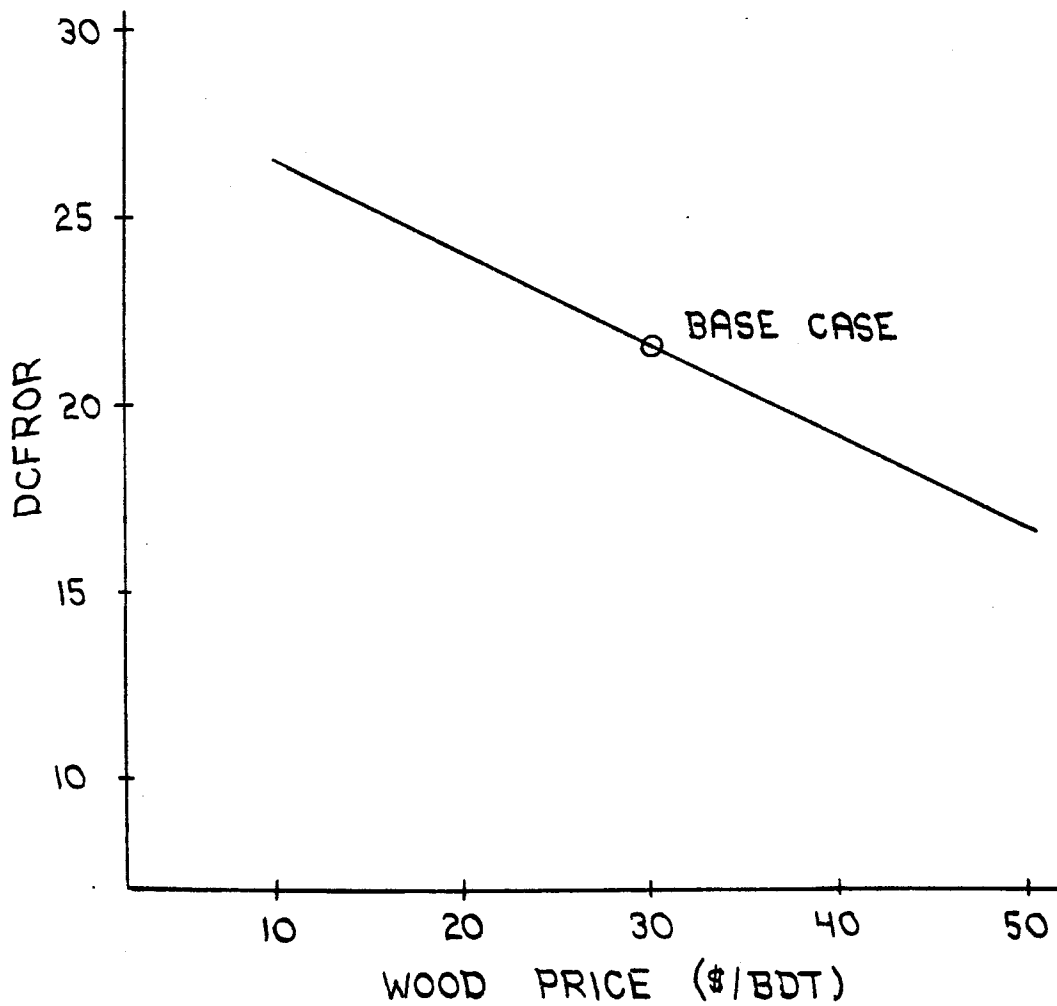
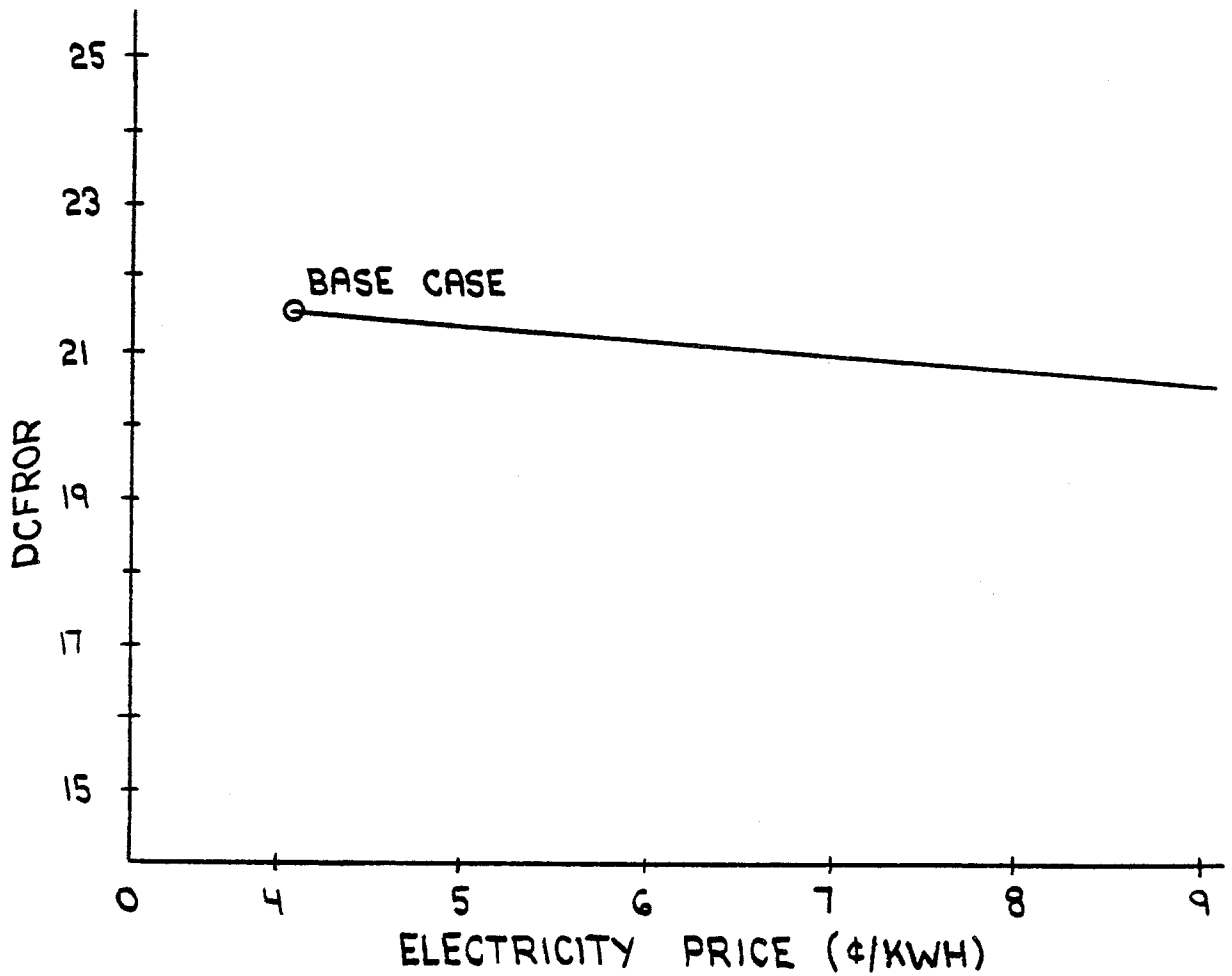


FIGURE 6-5
DCFROR^(a) VS. ELECTRICITY PRICE



(a) Based On Methanol Plant Electricity Only
(O₂ Plant Electricity Not Included)

FIGURE 6-6
DCFROR VS. OXYGEN PRICE

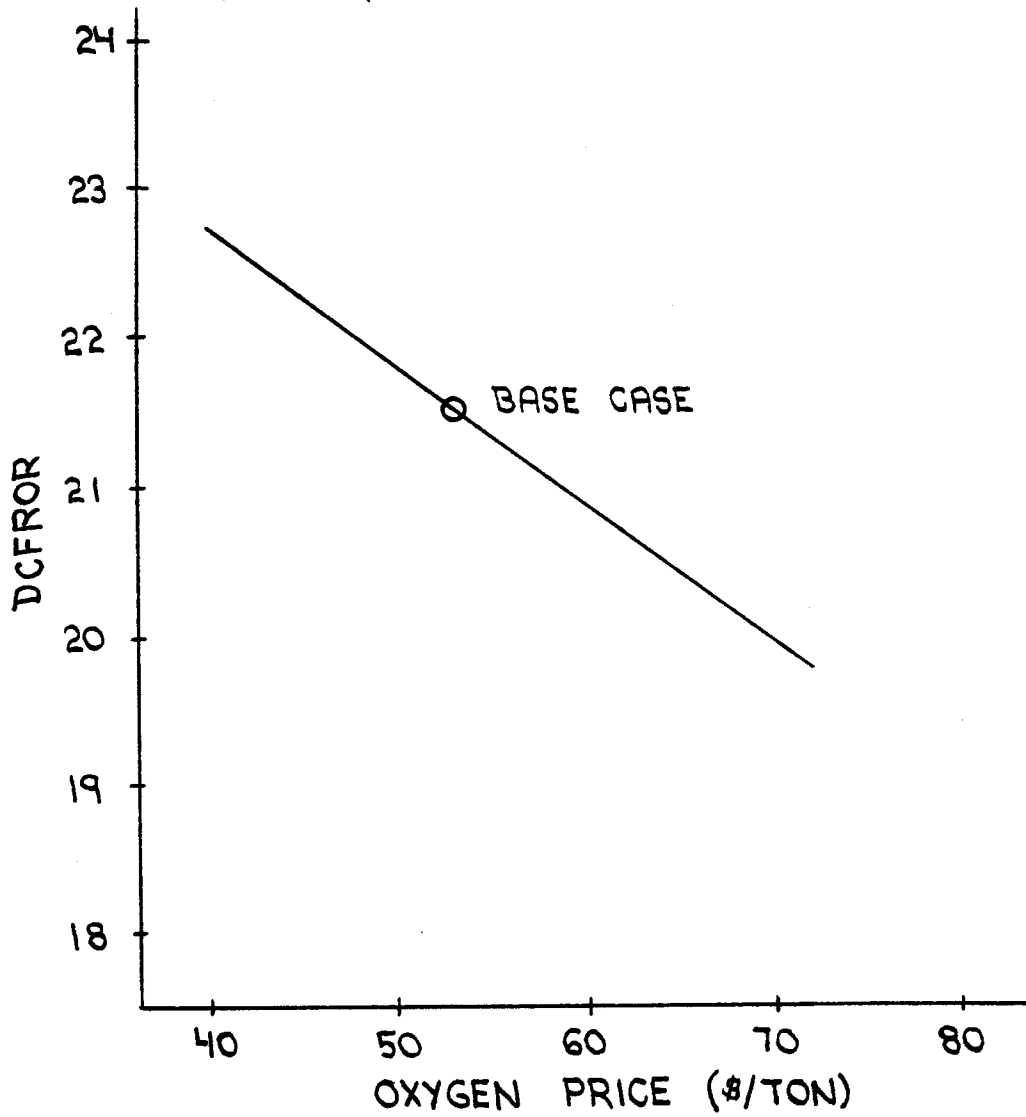
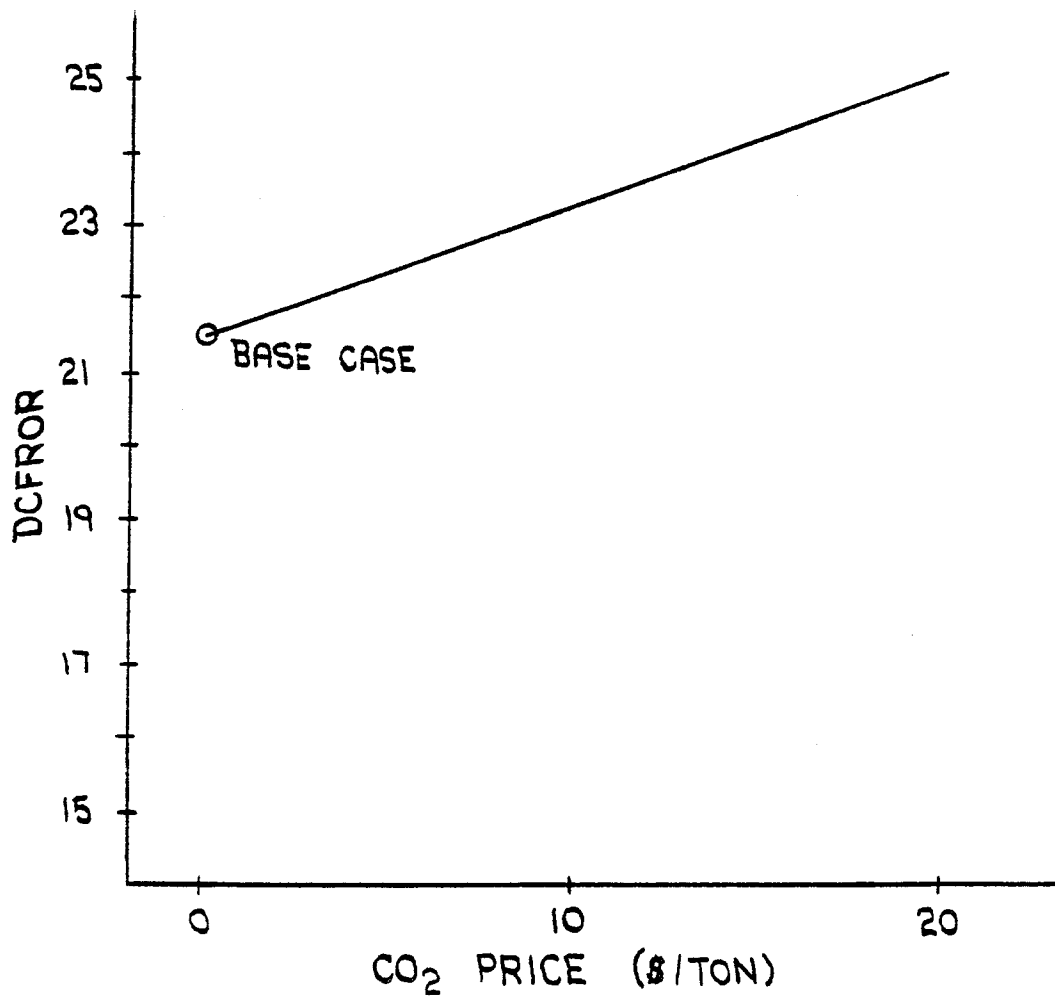


FIGURE 6-7
DCFRROR VS. CARBON DIOXIDE PRICE



SECTION 7

ALTERNATIVE DESIGNS TO THE BASE CASE AND COMPARATIVE ECONOMICS

The base case design of the biomass-based methanol plant (BMP) has defined a starting point in the analysis of the BMP's economic potential and design. Different basis assumptions will cause changes to the base case design. These changes and the corresponding capital and operating costs provide insight into the gasifier operating conditions that improve the economics of producing methanol from biomass. These conditions, in turn, can be used to identify gasifier development needs (see Section 8). Three alternative conditions, greater throughput or multiple gasifiers, increased moisture content wood feed, and higher gasifier operating pressure, have been examined. Additionally, the conversion of methanol synthesis purge gas methane to hydrogen for additional carbon conversion to methanol has been examined.

7.1 HIGHER OPERATING PRESSURE

Increasing the gasifier operating pressure from 150 psig to approximately 360 psig eliminates one stage of compression prior to acid gas removal and methanol synthesis. Because the gasifier yield model is insensitive to pressure in this range (see Section 4.2), the gas composition is not changed from the base case. Except for a reduction in power consumption during synthesis gas compression, the base case energy and material balances are unchanged. Table 7-1 summarizes capital cost changes. The major savings is due to eliminating one stage of compression. Cost increases in Sections 200 (gasification) and 300 (shift) are due to the increased pressure ratings required. Operating cost changes and methanol production costs are shown in Table 7-2. The major change is due to decreased power consumption.

Importation of electricity is reduced to about 100 kW, but the cost savings of approximately 1.5¢/gal is not a significant reduction in operating cost. Raising the gasifier pressure to methanol synthesis pressure (~1000 psia)



would probably increase capital cost because of the additional high pressure piping and vessel costs. The lock hopper system used is not proven at 800-1000 psig. Therefore, increasing gasifier operating pressure significantly beyond 150 psig should not be a priority development goal.

7.2 INCREASED PLANT SCALE

A larger BMP can be obtained by increasing the throughput of the gasifier or by installing two gasifiers with the currently assumed capacity. Although it may not be attainable, for this discussion, the throughput of the gasifier has been doubled from that of the base case to 3.56 MM Btu/hr-ft². Energy and material balance numbers are also doubled.

The savings in this case accrue to economies of scale in larger equipment, since wood, chemical and utilities consumption per unit of methanol produced is not changed. There is also a savings on operating labor; only 6 additional personnel (a 40 percent increase) are required. Table 7-3 compares the capital cost for the larger plant with the base case on a section-by-section basis. The increased cost/noneconomy of scale in Section 400 is due to switching from a reciprocating compressor to an axial compressor which is more reliable and avoids vibrational stresses. In Section 100, two wood dryers and vapor compressors of the same size as the base case design are required, because single-sized units are not available. Note that the scale factors used in this analysis cannot be extrapolated much further because multiple trains would be used.

Table 7-4 compares operating costs. Oxygen costs are reduced due to economies of scale. Because the capital savings are significant, doubling the size of the BMP can reduce methanol production costs. Wood resources for doubling green wood feed to 627 tpd are available in the Spokane area, but further enlargement of the plant may increase the cost of the wood resources required. Because the population of the Spokane-Coeur d'Alene area is only 401,605 (1980), use of 144 tpd of methanol would require that almost all area automobiles use a 10 percent blend of methanol in gasoline. Shop



fabrication of the enlarged modules is still feasible, but special means of transportation would be required.

Conversely, for the doubled throughput rate for the SERI gasifier, reducing the gasifier diameter to accommodate only the base case throughput reduces capital cost by only \$37,000 and therefore methanol production cost savings are negligible.

7.3 HIGHER MOISTURE CONTENT FEED

Increasing the allowable moisture content of the wood feed to the gasifier provides the benefits listed below:

- Reduced amount of steam required for shifting
- More shifting in the gasifier
- Decreased dryer energy consumption and size
- Smaller vapor compressor

Because the gasifier is in shift equilibrium and wood must be combusted to evaporate the additional wood moisture, the gasifier product composition and outlet temperature (for the same char conversion) are different. With lower gasifier temperatures, methane production is higher than for the base case. The estimated raw gasifier composition is shown in Table 7-5. Overall energy and material balances are shown in Figures 7-1 and 7-2, respectively, using the base case, dry wood feed rate. (Actual wood feed rate to the gasifier is higher due to higher wood moisture content.) While the dryer energy requirement is reduced, the use of a direct dryer fired with (hydrogen stripped) purge gas would necessitate burning green chips in the FBC to balance the steam demand. Therefore, the cascading use of dryer waste heat via vapor compression is still beneficial and should be used.

Methanol yield is slightly lower than the base case due to higher methane and carbon dioxide (from wood combustion) production in the gasifier. Because less wood moisture latent heat is recovered, a small amount of green



wood (about 200 lb/hr) must be burned in the FBC to balance the steam demand. Because more methane is produced, electricity generation from the diesels is greater.

Table 7-6 compares the capital costs by plant section between the base case and higher moisture feed case. Capital costs are lower for the alternative case in the wood handling section because the dryer and vapor compressor are smaller. Since more condensate is produced, the phenol extraction (Section 900) equipment is larger. The diesel engine and exhaust waste heat recovery capacity is increased about 20 percent. A smaller shift reactor is needed and less process quality water is required. Overall capital costs are essentially unchanged.

Operating costs for the base case and the higher moisture feed case are shown in Table 7-7. Slightly more oxygen is consumed to evaporate the extra moisture in the feed. Imported power is reduced because more electricity is produced from the methanol purge gas and less power is needed to recompress the reduced dryer vapor flow. The net change in methanol cost is +3¢/gal.

7.4 REFORMING PURGE GAS METHANE

The methane in the methanol synthesis purge gas represents a source of hydrogen and carbon that can be consequence (reformed) to a synthesis gas to produce additional methanol. The consequence is a decrease in electricity generation. The small tubular reformer(s) necessary are commercially available. Figure 7-3 is a simplified process flow diagram. Fuel for the reformer is vent gas from methanol purification and a portion of the purge gas. The reformed product is combined with the gasifier shift bypass to minimize the cost of cooling and heat recovery. Slightly less carbon dioxide removal will be required in the acid gas section because of the extra hydrogen production.

Because reformer products are in shift equilibrium, more methane can be converted if purge gas hydrogen is removed via the PSA unit before the



methane is reformed. Therefore, the PSA unit is unchanged except for the increased purge gas volume due to methane recycle (unconverted methane in the reformer product). Block flow energy and material balances are shown in Figures 7-4 and 7-5, respectively. Methanol yield is increased by 11 tpd or 14 percent.

Capital cost changes are summarized and compared in Table 7-8 with the base case. Capital costs are slightly higher because of the increased flow downstream of the gasifier. The reformer cost is offset partially by eliminating the diesel engines. Operating costs shown in Table 7-9 for the reformer case are higher due to increased power costs and higher capital-related charges. The increased operating costs are offset by the increased methanol yield. The reduction in methanol price by reforming the purge gas is 8¢/gal.



Table 7-1

Capital Cost Changes for Higher Gasifier Operating Pressure

<u>Section Changed</u>	<u>Equipment Cost Change</u>
Section 200 - Gasification	+\$ 88,000
Section 300 - Gas Shift	+ 10,000
Section 400 - Gas Compression and Cooling	- 170,000
Total Plant Installed Cost	-\$100,000

Table 7-2

Changes in Operating Costs for Higher Gasifier Operating Pressure

	Cost Change	
	<u>\$/yr</u>	<u>¢/gal</u>
Electricity	-\$149,000	-1.94
Maintenance Charges	- 5,000	-0.06
Oxygen (extra compression)	+ 38,000	+0.49
Total	-\$116,000	-1.51

Table 7-3

Capital Cost for Doubled Plant Size

<u>Section</u>	<u>Section Scale Factor</u>	<u>Section Direct Cost</u>
100 - Wood Handling	0.875	\$ 3,378,000
200 - Gasification	n.a.	895,000
300 - Gas Shift	0.625	258,000
400 - Gas Cooling and Compression	n.a.	1,259,000
500 - Acid Gas Removal	0.69	1,881,000
600 - Methanol Synthesis	0.65	5,718,000
700 - Methanol Purification	0.75	1,116,000
800 - Methanol Purge Gas Treatment	0.60	552,000
900 - Wastewater Treatment	0.59	93,000
1000 - Water Treatment and Steam Plant	0.56	674,000
1100 - General Utilities and Waste Disposal	0.64	3,215,000
1200 - General Facilities	0.47	6,309,000
Total Base Cost ^(a)		\$25,348,000

(a) Not including engineering, construction management, contingency, etc.



Table 7-4

Methanol Production Costs with Plant Capacity Doubled
(in ¢/gal)

<u>Component</u>	<u>7.68 MM gpy</u>	<u>15.36 MM gpy</u>
Wood	19.98	19.98
Oxygen	13.14 a	11.43 b
Electricity	2.30	2.30
Catalyst and Chemicals	3.08	3.08
Miscellaneous	1.30	1.30
Labor	9.38	6.70
Maintenance	6.62	5.39
Insurance and Local Taxes	3.16	2.48
Capital Recovery ^c	<u>57.06</u>	<u>42.34</u>
Total	116.00	94.90

^aOxygen cost for 60 tpd includes \$36.53/ton for capital recovery and \$16.29/ton for electricity.

^bOxygen cost for 120 tpd includes \$30.14/ton for capital recovery and \$15.82/ton for electricity.

^cAt a 21.5 percent annual rate of return (base case).

Table 7-5

Gasifier Product for the Higher Moisture Wood Feedstock Case
 (Basis - 144 dry tpd of wood to gasification)

<u>Species</u>	<u>Higher Wood Moisture Case</u> (27 wt percent)		<u>Base Case</u> (17 wt percent)	
	<u>mph</u>	<u>mole percent</u>	<u>mph</u>	<u>mole percent</u>
H ₂	383.8	36.39	332.3	35.34
CO	214.4	20.33	287.8	30.60
CO ₂	248.1	23.52	178.0	18.93
CH ₄	33.8	3.20	30.0	3.19
H ₂ O	171.2	16.23	108.8	11.57
N ₂	<u>3.5</u>	<u>0.33</u>	<u>3.5</u>	<u>0.37</u>
	1054.8	100.00	940.4	100.00



Table 7-6

Capital Cost Savings by Gasifying Wetter Wood Feed

<u>Section with Cost Change</u>	<u>Direct Cost Savings</u> ^b
100 - Wood Handling and Receiving	\$ 184,000
400 - Gas Cooling and Compression	(10,000) ^a
900 - Wastewater Treatment	(8,000) ^a
1000 - Water Treatment and Steam Plant	39,000
1100 - General Utilities and Waste Disposal	<u>(135,000)</u> ^a
Total Capital Savings	\$ 70,000

^aCost Increase

^bCompared to the Base Case

Table 7-7

Higher Moisture Wood Case Operating Costs
(7.41 MM gpy)

<u>Component</u>	<u>\$/yr.</u>	<u>¢/gal</u>
Wood (\$30/BDT)	1,544,279	20.84
Oxygen (\$52.82/ton) ^a	1,033,735	13.95
Electricity (4.07¢/kwh)	70,914	0.96
Catalysts and Chemicals	232,101	3.13
Miscellaneous	79,798	1.08
Labor	720,000	9.72
Maintenance	500,300	6.75
Insurance and Local Taxes	241,260	3.26
Capital Recovery ^b	<u>4,401,408</u>	<u>59.40</u>
Total	8,823,795	119.09

^aOxygen cost consists of \$36.53/ton for capital recovery and \$16.29/ton for electricity consumed.

^bAt a 21.5 percent annual rate of return (base case).



Table 7-8

Capital Cost Changes for Reforming the Purge Gas

<u>Section Changed</u>	<u>Cost Change from Base Case</u>
300 - Gas Shift	+\$ 5,000
400 - Gas Cooling and Compression	+ 57,000
600 - Methanol Synthesis	+ 233,000
700 - Methanol Purification	+ 66,000
800 - Purge Gas Treatment	+ 72,000
900 - Wastewater Treatment	+ 12,000
1100 - General Utilities	- 454,000
1200 - General Facilities	+ 150,000
1300 - Reforming Plant (new)	+ <u>672,000</u>
Total Section Changes	+\$813,000
Total Installed Cost Change ^a	+\$996,000

^aIncluding indirect capital charges such as engineering, contingency, etc.



Table 7-9

Operating Costs for Reforming Purge Gas Case
(8.77 MM gpy)

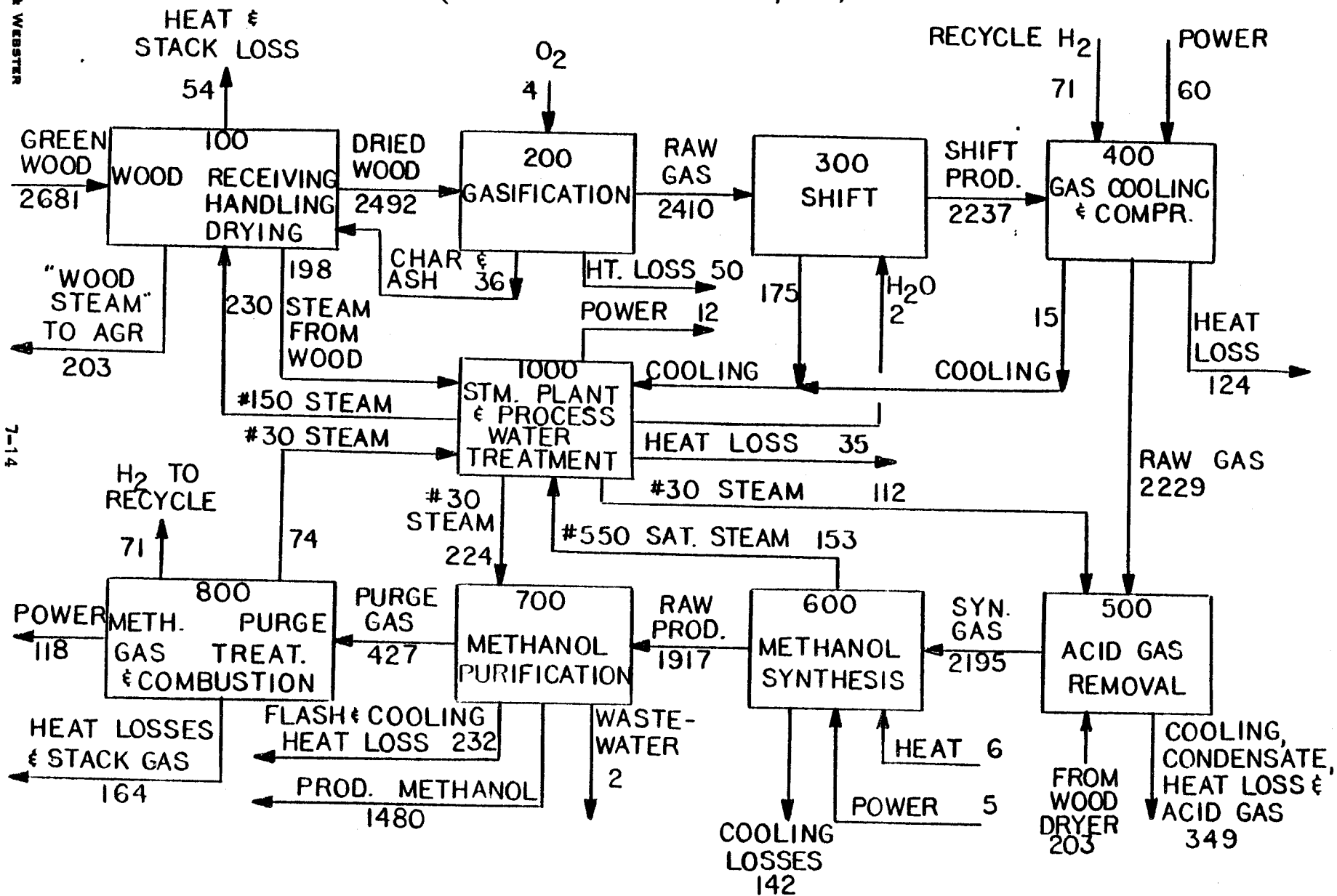
<u>Component</u>	<u>\$/yr.</u>	<u>¢/gal</u>
Wood (\$30/BDT)	1,534,424	17.49
Oxygen (\$52.82/ton)	1,009,443	11.50
Electricity (\$0.0407/kWh)	532,813	6.07
Catalysts and Chemicals	250,000	2.85
Miscellaneous	99,768	1.14
Labor	720,000	8.21
Maintenance	533,000	6.08
Insurance and Local Taxes	254,505	2.90
Capital Recovery ^a	<u>4,601,015</u>	<u>52.23</u>
Total	9,534,968	108.47

^aAt a 21.5 percent annual rate of return (base case).



HIGHER MOISTURE WOOD CASE ENERGY BALANCE

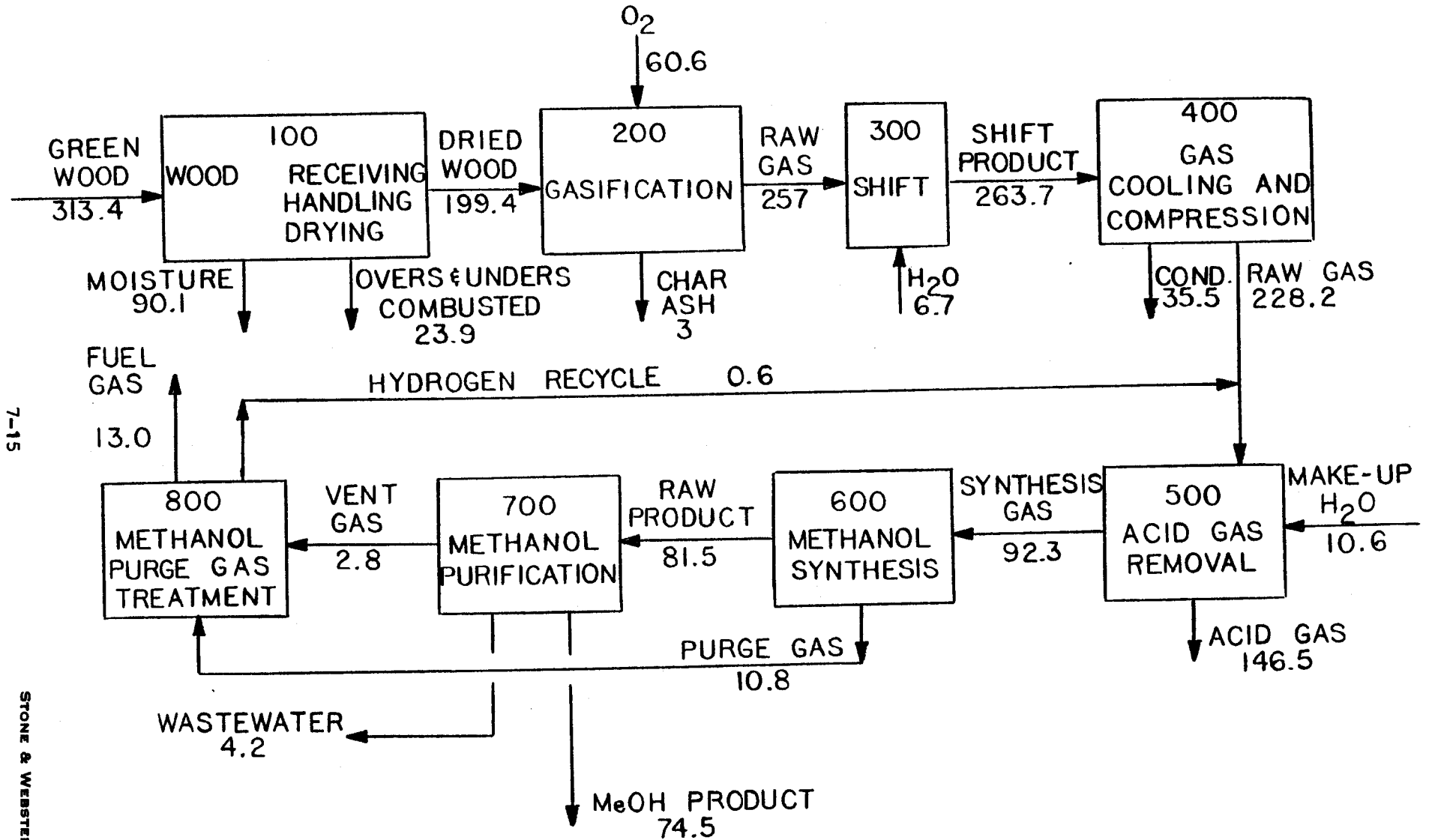
(ALL FLOWS IN MM BTU/DAY)



STONE & WEBSTER

7-14

HIGH MOISTURE WOOD CASE MATERIAL BALANCE (ALL FLOWS IN TONS PER DAY)



7-15

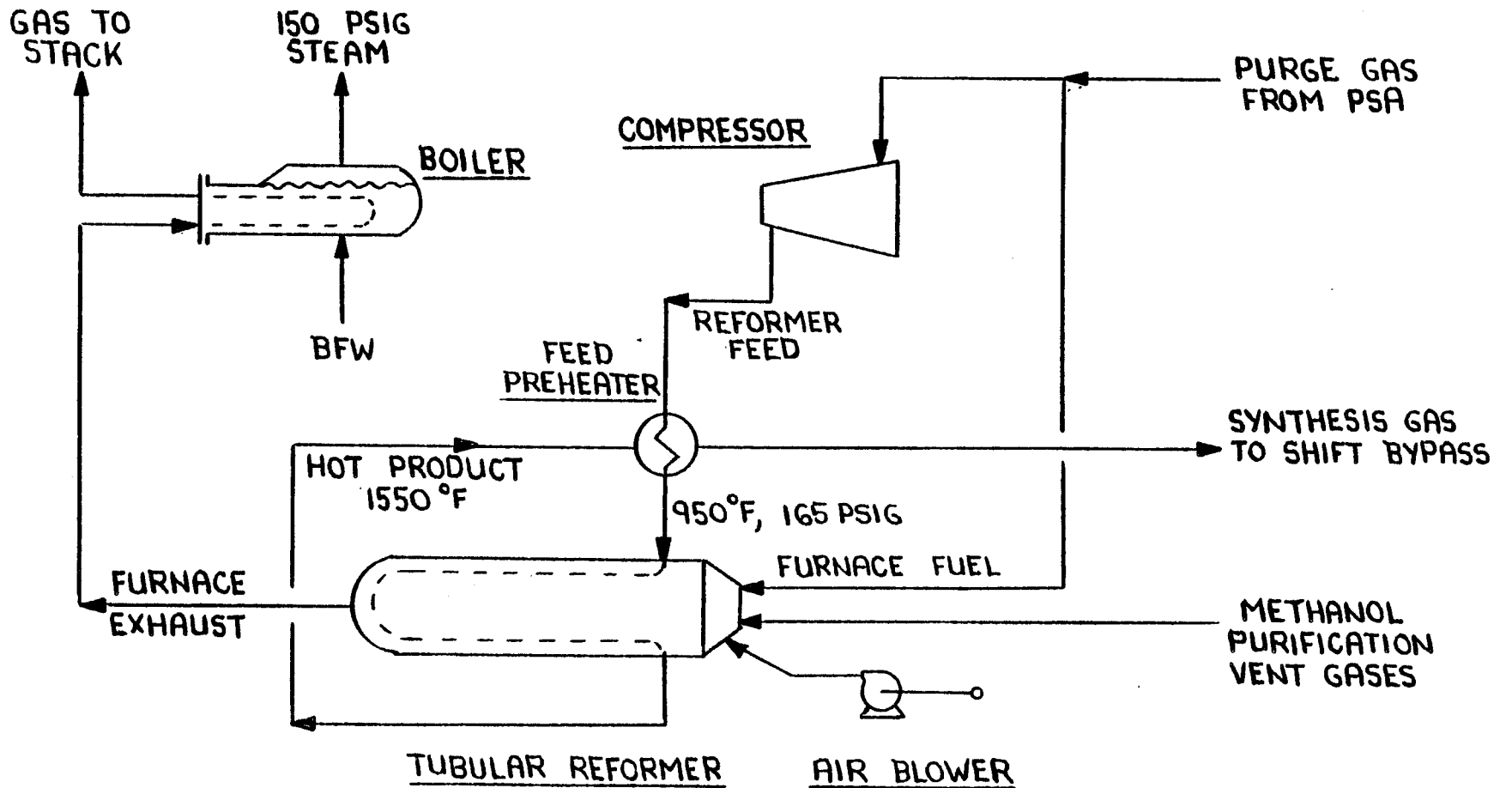
STONE & WEBSTER



FIG. 7-2

FIGURE 7-3

ADDITION OF PURGE GAS REFORMING TO THE BASE CASE METHANOL PLANT



REFORMER CASE ENERGY BALANCE

(ALL FLOWS IN MM BTU/DAY)

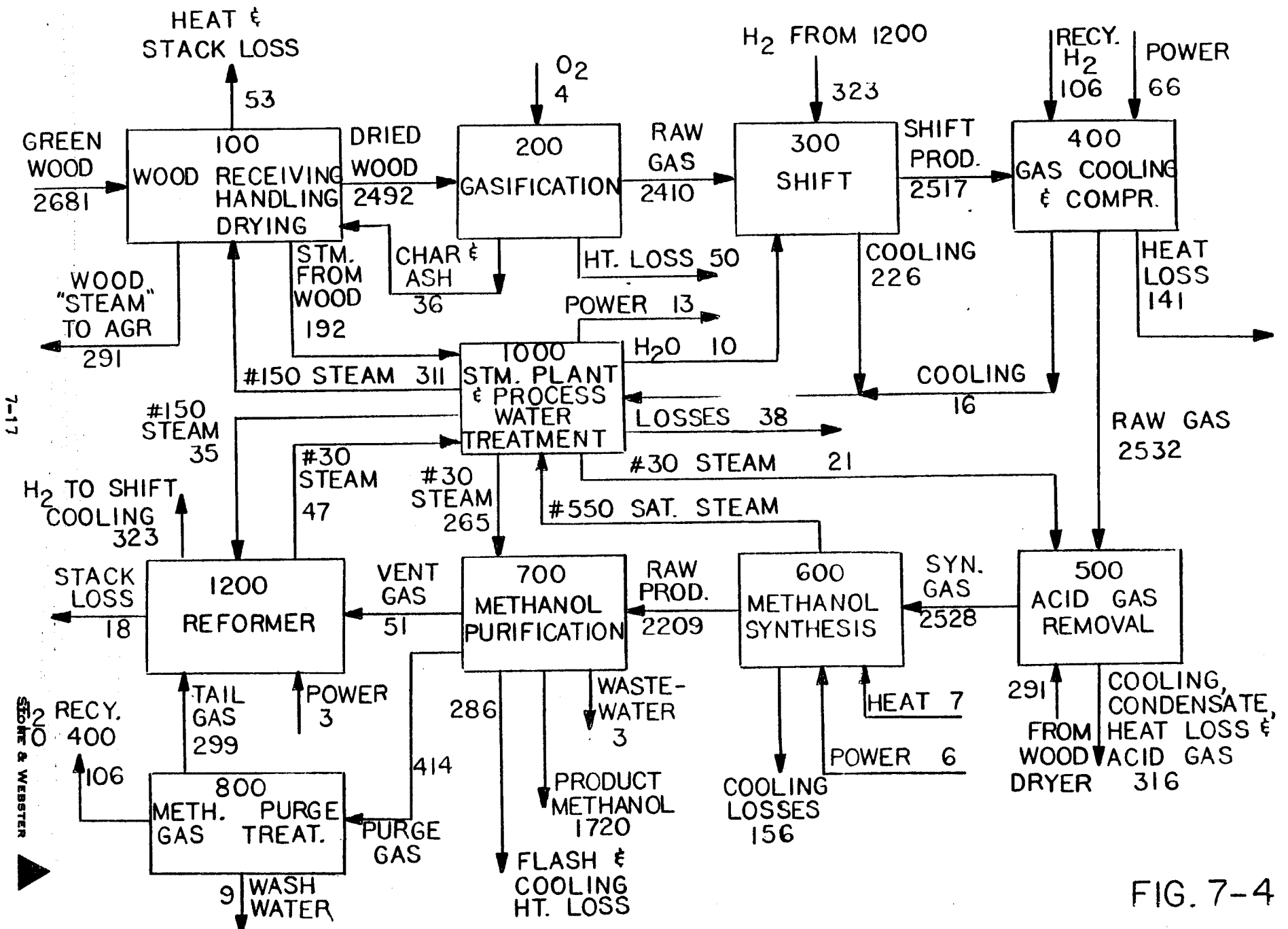
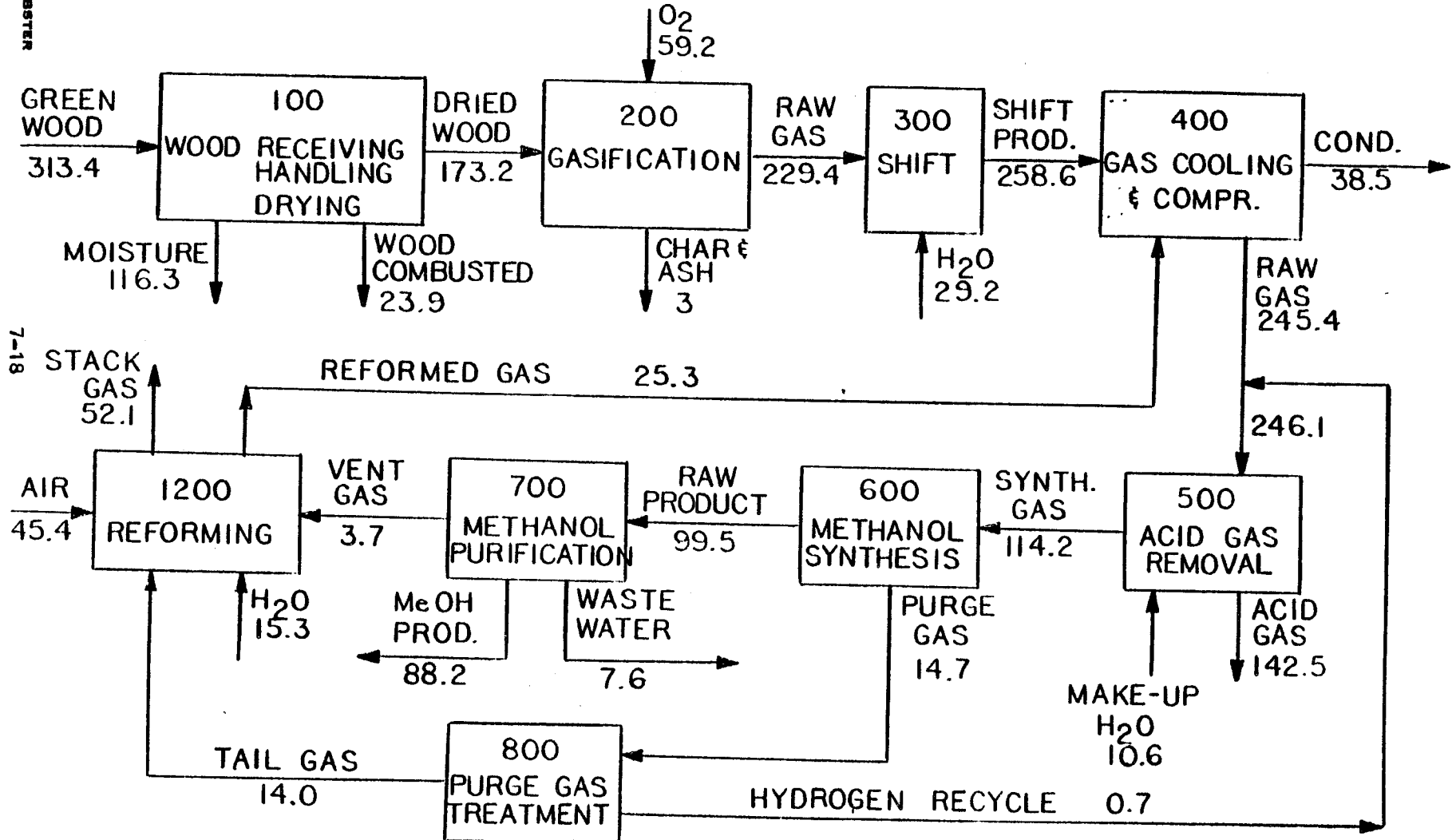


FIG. 7-4

REFORMER CASE MATERIAL BALANCE

(ALL FLOWS IN TONS PER DAY)



7-18

SECTION 8

RESEARCH AND DEVELOPMENT NEEDS

The next milestone in the commercialization of the biomass-to-methanol concept is the scale-up of the SERI gasifier. The SERI gasifier has been tested only at a scale approximately 1/400 of the base case commercial design used in this study. An intermediate sized pilot plant should be built to evaluate the gasifier at a scale compatible with confirming all operating concepts and assumptions, but still small enough to build and operate without great expense. Some of the required information should first be obtained in the existing pilot facility to ensure that the commercial biomass from methanol plant designs are achievable and to incorporate this information into the design of the larger pilot unit.

The following information can be obtained with the current SERI pilot plant gasifier and is needed to provide process and specification information for an actual plant design:

- Reproducible, tightly closed energy and material balances to accurately verify gas yields and composition. Elemental balances (C, H, O, S, and N) should also be closed. Operating conditions should be at 150 psig and with oxygen.
- Determination of the sulfur species present in the gas phase to properly design sulfur removal facilities prior to methanol synthesis, and establish acceptable acid gas disposal methods.
- Analysis of raw gas condensate from scrubbing (especially tar compounds condensed) to design a condensate treatment system.
- Operation with wood feed at higher moisture contents (especially wood with only surface moisture removed) to establish yields and gas composition.



- Operation with waste product feedstocks such as rice hulls to determine the suitability of gasifying low cost materials.
- Determination of particle size distribution and disposition (grate, cyclone, or scrubber) of the char and ash.

Concurrently, a test of indirect drying of wood should be arranged to determine the heat transfer coefficients and conditions needed for indirect drying of wood chips to 15-17 percent moisture content in commercial equipment. Mechanical designs for oxygen distribution and solids agitation should be made.

As discussed in Section 4.2, SERI pilot plant data to date (January 1984) do not simultaneously fulfill closure of the material and elemental balances. The actual pilot plant data available were simulated by SWEC with a computer model that does simultaneously close these balances. The computer model was then used to simulate commercial plant operating conditions. Pilot plant data that produce closure and are reproducible should be obtained to confirm or modify the design basis used in this study.

Currently, sulfur dioxide is being produced by the gasifier, probably due to burning of reduced species above the bed level in the gasifier (i.e., in the freeboard). Oxidized sulfur is not reduced as it passes through the bed; the equilibrium constant is unfavorable. Sulfur oxides will be captured by the Benfield acid gas scrubbing system, but the potassium sulfate formed must be purged from the system. The cost of replacing spent Benfield solution is approximately \$130,000 per year or 1.75 cents per gallon of methanol. Prevention or reduction of sulfur oxides formation can reduce this cost. In addition, the distribution of sulfur species affects the design of the acid gas treatment and disposal system.

The composition and yield of the tar compounds are needed. If tar composition is less phenolic and/or tar yields are different, the treatment of the condensate from raw gas scrubbing will be changed. For example, if phenol production is much lower than estimated, the condensate treatment system (Section 900) might be unnecessary. If phenol production is higher, the size of the extraction equipment is increased.



The current pilot plant has been operated with wood containing less than 5 weight percent moisture. This is not feasible in a commercial dryer without expending a large amount of energy for drying and using more exotic dryers than those discussed in the process design and energy integration. The pilot plant should be operated using wood at a moisture content that can be achieved in an indirect dryer (this value must be confirmed in a dryer test). These data should be used to make the required material and elemental balances.

Agricultural waste product feedstocks should be tried in the gasifier. If suitable for the SERI gasifier, low or zero cost material such as rice hulls can expand the geographic range of sites for a profitable biomass-based methanol plant. For example, if rice hulls are suitable, they can be obtained on average at zero cost in the Houston area where oxygen is available at lower cost (than Spokane) and carbon dioxide should be saleable. The savings on feedstock (20¢/gal.) and oxygen (3¢/gal.) and the sale of carbon dioxide (5¢/gal.) offset the transportation advantage in Spokane (26¢/gal.) to still produce a 21.5 percent after-tax rate of return for the base case.

Additionally, testing of rice hulls will help evaluate the hypothesis that gas phase temperatures are higher than solid particle temperatures in the gasifier. Because the gasifier flaming pyrolysis zone operates above the fusion temperature of rice hull ash, the hypothesis should be confirmed if the hulls are not fused.

Once the above information has been obtained, the gasifier should be scaled up to a size at which the oxygen distribution system and mechanical agitation can be realistically tested for mechanical integrity and suitability. Maximum gasifier throughput should also be studied at this phase. Gasifier feed size should also be re-evaluated and variations (if any) in yield, gas composition, throughput or gasifier operability with feedstock size should be noted. The pilot work should also reconfirm the design assumptions and energy and material balances produced from the data taken in the smaller pilot plant. A continuous 30 day run should be a minimum goal to evaluate reliability and integrity of materials.



Any changes in gasifier yield and throughput with feedstock (wood) size are important information in specifying wood chips to the plant and the detailed requirements of the wood handling system. The variation in char/ash particle size distribution and char yield as a function of wood feed size and gasifier throughput are important for designing the char recovery facilities - the char hoppers, char quench, gasifier cyclone and venturi scrubber system. Minimizing char production is important to maximize synthesis gas and methanol yields.

The size of the pilot gasifier should be approximately 3 ft, inside diameter or 4.5 ft, outside diameter. At the current assumed throughput, approximately 2 tph of green wood would be required. Because downstream processing technology is commercial and does not need to be integrated directly with the gasifier, the raw gasifier product can be burned or flared. This larger pilot plant should be near a source of oxygen, because oxygen storage would be a considerable problem for the pilot facility. A chipped wood source nearby would also be useful to reduce the amount of wood handling and preparation equipment required for the pilot plant.

Table 8-1 lists the major equipment required to be purchased for a larger pilot facility, assuming it will have ready access to both oxygen and wood chip supplies.



Table 8-1

Major Equipment Needed for Large Scale Pilot Testing

Reclaim Hopper and Conveyor
Magnetic Separator
Overs Screen
Dryer Feeder
Direct Rotary Dryer (with pollution controls
for exhaust, if necessary)
Fines Screen
Storage Conveyor/Elevator
Storage Hoppers (Live Bottom)
Gravimetric Feeder
Bucket Elevator
Lock Hopper
Feed Screw
Gasifier
Gasifier Cyclone
Char and Ash Lock Hopper
Char and Ash Removal Screw
Air Blower
Flare



APPENDIX A

EQUIPMENT LIST

WITH WEIGHTS AND DIMENSIONS



APPENDIX A
EQUIPMENT LIST WITH DIMENSIONS AND WEIGHTS

Section 100

Wood Handling

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
C-101	Scalper Feed Conveyor	11,200	24" □ x 83' L
C-102	Dryer Feed Conveyor	6,600	24" □ x 72" L
C-103	Fines Conveyor	7,700	24" □ x 103' L
C-104	Stored Fines Conveyor	3,800	18" □ x 20' L
C-105	Dried Fines Conveyor	3,800	18" □ x 20' L
G-101	Truck Scale	--	10' x 35'
G-102	Tractor Shovel	--	--
G-103	Reclaim Hopper and Chain Conveyor	--	6' H x 20' L x 6' W
G-104	Magnetic Separator	3,900	32" H x 106" L x 59" W
G-105	Scalping Screen	6,800	42" H x 151" L x 30" W ^a
G-106	Primary Screen	20,000	75" H x 151" L x 121" W ^a
G-107	Hog Mill and Blower	8,500	6' H x 76" L x 5' W
G-108	Fines Screen	7,000	54" H x 91" L x 48" W
G-109	Dryer Screen Feeder	2,300	20" □ x 6' L
G-110	Wood Dryer	215,000	12' H x 33' L x 10' W
G-111	Dryer Fines Deck Screen	1,500	3' H x 7' L x 52" W ^a
G-112	Fines Storage Silo and Reclaimer	26,000	13' ∅ x 23' H
G-113	Storage Bucket Elevator	6,700	65' H x 4' L x 20" W
G-114	Dried Chips Storage Silo and Reclaimer	69,000	25' ∅ x 45' H
G-115	Gravimetric Feeder	3,400	57" H x 80" L x 35" W
G-116	Rail Car Weigh Scale (optional)	--	--
G-117	Fines Discharge Bucket Elevator	5,500	40' H x 4' L x 20" W
G-118	Disposal Dumpster	--	--
G-119	Rotary Vane Feeder	with G-110	with G-110
R-101	Dryer Vapor Recompressor	8,000	60" H x 160" L x 67" W
R-102	Steam Turbine	900	28" H x 45" L x 26" W

^a Screen dimensions only, does not include discharge hopper.

Section 200

Gasification

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
C-201	Wood Conveyor	5,400	24" □ x 32' L
C-202A&B	Top Screen Feeders	@ 700 @	12" Ø x 6'
C-203	Cyclone Char Removal Screw	900	10" Ø x 10' L
G-201A&B	Lock Hopper Feeders	not available	as per vendor
G-202	Char and Ash Lock Hopper	1,600	see sketch
G-203	Lock Hopper Char and Ash Removal Quench Screw	800	9" Ø x 10' L
G-204A&B	Pressurized Bucket Elevators	5,100	35' H x 2' L x 2' W
G-205A&B	Bucket Elevators	4,200	35' H x 2' L x 2' W
L-201	Gasifier	50,000	see sketch
M-201	Cyclone	4,800	30" Ø x 10' H
R-201	Startup and Shutdown Blower	8,000	35" H x 60" L x 31" W
T-201	Gas Cooler/Waste Heat Boiler	600	20" Ø x 8' TT, vert.



Section 300

Gas Shift

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
F-301	Shift Startup Superheater	1,250	8' H x 2' L x 2' W
L-301	Shift Reactor	23,500	6' \emptyset x 8' TT, vertical
M-301	Shift Cooling Steam Drum	2,200	42" \emptyset x 42" TT
T-301	Waste Heat Boiler	3,600	33" \emptyset x 8' TT, vertical
T-302	Waste Heat Economizer	2,800	16" \emptyset x 8' TT, vertical
T-303	Shifted Gas Air Cooler	12,500	6' H x 12' L x 10' W



Section 400

Gas Cooling and Compression

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
G-401	Venturi Scrubber	5,200	2' \emptyset x 6'9" H
G-402	Scrubber Filter	500	4' H x 2' L x 1' W
M-401	Vapor/Liquid Separator	with G-401	4'2" \emptyset x 8' TT, vert.
M-402	Oil/Water Separator	1,000	2' \emptyset x 5' TT, horiz.
P-401A&B	Scrubber Recycle Pump and Spare	@ 750	@ 24" H x 60" L x 28" W
R-401	First Stage, Synthesis Gas Compressor	61,000	9' H x 17' L x 16' W
R-402	Second Stage, Synthesis Gas Compressor	with R-401	with R-401
T-401	Scrubber Water Cooler	1,600	14" \emptyset x 8' TT, horizontal
T-402	First Stage Compressor Cooler/ Process and BFW Heater	1,300	10" \emptyset x 8' TT, horizontal
T-403	First Stage Compressor Air Cooler	3,000	6' H x 6' L x 6' W



Section 500

Acid Gas Removal

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
A-501	Acid Gas Absorber	not available	not available
A-502	Acid Gas Stripper	not available	not available
G-501	Carbonate Mixer	with Q-501	with Q-501
M-501	Clean Syngas Water K.O. Drum	1,000	2' Ø x 61" TT, vert.
M-502	Acid Gas Water K.O. Drum	3,200	42" Ø x 106" TT, vert.
M-503	Steam Blowdown Drum	1,000	30" Ø x 90" TT, vert.
M-504	Final Acid Gas Water K.O. Drum	800	28" Ø x 72" TT, vert.
P-501A&B	Solution Circulation Pump and Spare	@ 2,700	@ 36" H x 118" L x 30" W
P-502A&B	Solution Makeup Pump and Spare	with Q-501	with Q-501
P-503A&B	Carbonate Mix Tank Pump and Spare	with Q-501	with Q-501
P-504A&B	Knockout Drum Recirculation Pump and Spare	200	1' H x 2' L x 1' W
P-505A&B	Acid Gas Knockout Drum Recirculation Pump and Spare	500	15" H x 52" L x 15" W
Q-501	Carbonate Mix Tank	200	4' H x 6' L x 6' W
Q-502	Solution Makeup Tank	9,400	7' Ø x 16' TT, horiz.
T-501	Clean Syngas Air Cooler	5,500	6' H x 8' L x 6' W
T-502	Clean Syngas Trim Cooler	not available	not available
T-503	Acid Gas Air Cooler	57,000	6' H x 20' L x 12' W
T-504A&B	Stripper Reboiler and Spare	@ 9,000	@ 7' Ø x 16' TT, horiz.
T-505	Acid Gas Final Cooler	2,600	14' Ø x 8' TT, horiz.
T-506A&B	Condensate Cooler & Spare	@ 1,500	@ 12" Ø x 8' TT, horiz.



Section 600

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
F-601	Guard Adsorber Startup Feed Heater	1,500	9' H x 2' L x 2' W
F-602	Guard Adsorber Final Heater	1,250	5'6" H x 2' L x 2' W
F-603	Reactor Startup Heater	not available	not available
G-601	Steam Ejector	not available	not available
L-601A&B	Guard Adsorber and Spare	@ 25,200	@ 4'6" \emptyset x 7'5" TT, vert.
L-602	Methanol Synthesis Reactor	not available	not available
M-601	Methanol Reactor Steam Drum	not available	not available
M-602	Raw Product Separator	not available	not available
R-601	Recycle Compressor	not available	not available
T-601	Guard Adsorber Feed Heater	not available	not available
T-602	Guard Adsorber Feed/Effluent Exchanger	3,200	20" \emptyset x 8' TT, horiz.
T-603	Methanol Feed/Effluent Exchanger	not available	not available
T-604	Raw Product Air Cooler	not available	not available
T-605	Raw Product Trim Cooler	not available	not available



Section 700

Methanol Purification and Product Storage

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
A-701	Stabilizer	3,300	2' \emptyset x 18'6" TT
A-702	Methanol Still	18,000	3'6" \emptyset x 60' TT
G-701	Caustic Addition System	200	not available
M-701	Flash Separator	1,600	3' \emptyset x 5' TT, horiz.
M-702	Stabilizer Reflux Drum	500	18" \emptyset x 41" TT, vert.
M-703	Methanol Still Reflux Drum	2,400	48" \emptyset x 64" TT, horiz.
M-704	Blowdown Drum	4,400	4'6" x 11'3" TT, horiz.
P-701A&B	Intermediate Tank Pump and Spare	@ 500	@ 12" H x 32" L x 12" W
P-702A&B	Blowdown Drum Pump and Spare	500	12" H x 32" L x 12" W
P-703A&B	Product Pump and Spare	500	20" H x 35" L x 19" W
P-704A&B	Stabilizer Reflux Pump and Spare	500	20" H x 35" L x 19" W
P-705A&B	Still Reflux Pump and Spare	500	20" H x 35" L x 19" W
P-706A&B	Product Transfer Pumps	500	25" H x 60" L x 25" W
Q-701	Intermediate Tank	13,300	12' \emptyset x 10' H
Q-701A-D	Final Product Tanks	@ 38,800	@ 12' \emptyset x 45' TT, horiz.
T-701	Stabilizer Feed Heater	300	6" \emptyset x 8' TT, horiz.
T-702	Stabilizer Reboiler	1,600	3'6" \emptyset x 8' TT
T-703	Stabilizer Air Cooler	7,500	6' H x 10' L x 6' W
T-704	Methanol Still Feed Heater	300	8" \emptyset x 8' TT, horiz
T-705	Methanol Still Reboiler	7,200	66" \emptyset x 10' TT, horiz.
T-706	Methanol Product Air Cooler	36,000	6' H x 24' L x 12' W
T-707	Methanol Product Final Cooler	500	10" \emptyset x 8' TT, horiz.

Section 800

Purge Gas Treatment

Account No.

Item Name

Empty Weight (lb)

Dimensions

A-801	Methanol Wash Tower	900	12" \emptyset x 90" TT, vert.
G-801	PSA Unit	not available	25' H x 42'10" L x 13'2" W
P-801A&B	Washwater Pump and Spare	< 200	small
T-801	Washwater Cooler	< 200	small



Section 900

Washwater Treatment

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
A-901	Extraction Tower	600	1' \emptyset x 8' TT
A-902	Phenol Still	800	1' \emptyset x 9' TT
M-901	Condensate Separator	400	18" \emptyset x 3' TT, vert.
P-901A&B	IPE ^a Feed Pump and Spare	< 200	small
P-902A&B	IPE ^a Recycle Pump and Spare	< 200	small
P-903A&B	Phenol Wastes Pump and Spare	< 200	small
P-904A&B	Wastewater Pump and Spare	400	12" H x 36" L x 12" W
Q-901	IPE ^a Storage Tank	1,000	2' \emptyset x 9' TT, vert.
T-901	Phenol Still Preheater	300	8" \emptyset x 8' TT, vert.
T-902	Phenol Still Condenser	< 200	4" \emptyset x 4' TT, vert.
T-903	Phenol Still Reboiler	< 200	8" \emptyset x 8' TT, vert.

^a isopropyl ether



Section 1000

Washwater Treatment

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
G-1001	Activated Carbon Filter	5,100	36" Ø x 96" TT, vert.
G-1002	Mixed Bed Demineralizer	4,800	36" Ø x 96" TT, vert.
G-1003	Ammonia Injection Skid	500	4' H x 6' L x 4' W
G-1004	Sodium Phosphate Injection Skid	500	6' H x 6' L x 6' W
G-1005	Sodium Sulfite Injection Skid	500	4' H x 6' L x 4' W
M-1001	Deaerator	2,300	10' H x 10' L x 5' W
M-1002	Steam Separator	2,100	42" Ø x 80" TT, vert.
M-1003	Condensate Flash Drum	2,200	42" Ø x 92" TT, vert.
M-1004	Condensate Flash Drum	2,100	40" Ø x 84" TT, vertical
Q-1001	Demineralized Water Tank	4,600	7' Ø x 28' TT, horiz.
Q-1002	Caustic Tank	9,500	7' Ø x 21' TT, horiz.
Q-1003	Hydrochloric Acid Tank	12,600	8' Ø x 26'6" TT, horiz.
P-1001A&B	Caustic Transfer Pump Spare	< 200	small
P-1002A&B	Hydrochloric Acid Transfer Pump and Spare	< 200	small
P-1003A&B	Process Water Pump and Spare	@ < 200	@ 12" H x 32" L x 12" W
P-1004A&B	Deaerator Pump and Spare	< 200	12" H x 32" L x 12" W
P-1005A&B	150 Psig BFW Pump and Spare	500	16" H x 47" L x 15" W
P-1006A&B	30 Psig BFW Pump and Spare	< 200	12" H x 32" L x 12" W
P-1007A&B	565 Psig BFW Pump and Spare	800	20" H x 46" L x 20" W
T-1001	Water Heater	< 200	small



Section 1100

General Utilities, Waste Disposal and Offsites

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
C-1101	Char and Ash Conveyor	4,200	18" □ x 30' L
G-1101	Fluidized Bed Combustion (FBC)	--	7.5 ft dia, height by vendor
G-1102	Flare	--	24" ∅ x 50' H
G-1103	Diesel Engine	35,600	113" H x 286" L x 80" W
G-1104	Diesel Engine	28,500	90' H x 206" L x 80" W
G-1105	FBC Stack	--	--
G-1106	Demineralizer Backwash Regeneration Skid	1,000	4' H x 6' L x 6' W
G-1107	Sulfuric Acid Skid	500	4' H x 3' L x 3' W
G-1108	Sodium Hypochlorite Skid	500	4' H x 3' L x 3' W
G-1109	Tramp Material Dumpster	--	8' H x 30' L x 8' W
G-1110	Baghouse and Cyclone Ash Dumpster	--	6' H x 8' L x 6' W
G-1111	FBC Metering Bin	--	by G-1101 vendor
G-1112	Char and Ash Storage Bin	3,800	39" ∅ x 13' H
G-1113	Diesels Stack	--	--
G-1114	Bucket Elevator	4,000	14" □ x 35' H
G-1115	Acid Gas Stack	--	--
G-1116	FBC Baghouse	--	by G-1101 vendor
G-1117	Fixed Foam Fire Protection System and Building	--	--
G-1118	Halon Fire Extinguishing System	--	--
L-1101	Sponge Iron Bed (Main)	14,000	10' ∅ x 18' TT, vert.
L-1102	Sponge Iron Bed (Spare)	3,100	3'6" ∅ x 12' TT, vert.
M-1101	FBC Steam Drum	2,400	42" ∅ x 54" TT, horiz.
M-1102	Diesels Waste Heat Boiler Steam Drum	1,700	42" ∅ x 42" TT, horiz.
M-1103	Oil/Water Separator	2,000	42" ∅ x 90" TT, horiz.
M-1104	Nitrogen Surge Drum	21,500	5' ∅ x 14'6" TT, horiz.
M-1105	Plant Instrument Nitrogen Receiver	2,700	4' ∅ x 8' TT, horiz.
M-1106	Plant Air Receiver	2,700	4' ∅ x 8' TT, horiz.
P-1101A&B	Liquid Wastes Pump and Spare	@ < 200	small
P-1102A&B	Cooling Water Pump and Spare	@ 600	@ 24" H x 53" L x 25" W

Section 1100

General Utilities, Waste Disposal and Offsites (Cont)

<u>Account No.</u>	<u>Item Name</u>	<u>Empty Weight (lb)</u>	<u>Dimensions</u>
P-1103A&B	Equalization Ponds Pump and Spare	@ 600	@ 12" H x 32" L x 13" W
P-1104A&B	Demineralizer Sump Pump and Spare	400	24" H x 54" L x 21" W
P-1105A&B	Oily Wastes Pump and Spare	< 200	small
P-1106A&B	Fire Water Pumps	@ 1,600	@ 30" H x 84" L x 30" W
P-1107A&B	Jockey Pump and Spare	1,500	25" H x 66" L x 25" W
P-1108A&B	Diesel Fuel Oil Pumps	< 200	small
Q-1101	Propane Storage Tank	8,100	5' \emptyset x 12'3" TT
Q-1102	FBC Liquid Wastes Storage Tank	< 200	55 gallon drums
Q-1103A&B	Runoff Equalization Ponds	--	@ 7,500 gallons
Q-1104	Demineralizer Wastes Sump	2,000	7.5' \emptyset x 15' TT, horiz.
Q-1105	Fuel Oil Tank	300	270 gallons
R-1101	FBC I.D. Fan	--	by G-1101 vendor
R-1102	Nitrogen Compressor	1,200	34" H x 59" L x 28" W
R-1103	Regeneration Blower	2,000	25" H x 65" L x 25" W
R-1104	Plant Instrument Nitrogen Compressor	7,500	99" H x 79" L x 41" W
R-1105	Plant Air Compressor	7,500	99" H x 79" L x 41" W
T-1101	FBC Economizer	--	by G-1101 vendor
T-1102	FBC Waste Heat Boiler	--	by G-1101 Vendor
T-1103	Diesels Economizer	600	10" \emptyset x 8' TT, horiz.
T-1104	Diesels Waste Heat Boiler	2,300	16" \emptyset x 8' TT, horiz.
T-1105	Cooling Tower	6,500	16' H x 12' L x 10' W
T-1106	Acid Gas Cooler	2,300	16" \emptyset x 8' TT, vert.



APPENDIX B

LISTING AND DOCUMENTATION

OF THE CHAR ZONE COMPUTER MODEL



```

C
C *****
C **          PROCESS,IAH,SERIMEOH(CHARZ)          **
C **                                               **
C **                                               **
C **                                               **
C **                                               **
C *****
C
0001 READ IN DATA
0002 DIMENSION CASE(20)
0003 REAL KP
0004 12 CONTINUE
0005 READ(5,11,END=5010)CASE,TIN,YCH4,YN2,YCO2,YH2O,YCO,YH2,GMASS,
0006 1CMASS,TOUT,I
0005 11 FORMAT(20A4/10(F10.5/),I2)
0006 90 CONTINUE
C 90'S LABELS ARE FOR DEBUG. TO USE, REMOVE STATEMENTS AT THE
C END FROM COMMENT.
0007 WRITE(6,1006)CASE
0008 1006 FORMAT(1H1// ' STONE AND WEBSTER ENGINEERING CORPORATION' /
0009 1' SERI METHANOL - CHAR ZONE MODEL' /1X,20A4//)
K=0
C
C HEATS OF FORMATION
C
0010 HFCO=-45233.
0011 HFCO2=-165115.
0012 HFH2O=-98530.
0013 HFH2=3643.
0014 HFCARB=453.
0015 GMW=16.042*YCH4+28.*YN2+44.01*YCO2+18.016*YH2O+28.01*YCO+2.016*YH2
0016 GMOLES=GMASS/GMW
0017 3 XCH4=GMOLES*YCH4
0018 XN2=GMOLES*YN2
0019 XCO2=GMOLES*YCO2
0020 XCO=GMOLES*YCO
0021 XH2=GMOLES*YH2
0022 XH2O=GMOLES*YH2O
0023 XC=CMASS/12.
0024 91 CONTINUE
0025 M=0
0026 JJ=0
0027 K=K+1
C
C INCREMENT TEMPERATURE CHANGE AND CHANGE (REDUCE) TEMPERATURE
C
0028 IF (K.EQ.1) GO TO 5
0029 DELT=(TIN-TOUT)/I
0030 T=TIN-K*DELT
0031 GO TO 10
0032 5 T=TOUT
C

```

B-1



```

C   CALCULATE SENSIBLE HEAT CHANGE FROM INLET CONDITIONS
C
0033 10 CALL ENTH(TIN,HCH4,HN2,HCO2,HCO,HH2,HH2O)
0034      H1=XCH4*16.042*HCH4+XN2*28.*HN2+XCO2*44.01*HCO2+XCO*28.01*HCO+
1      XH2*2.016*HH2+XH2O*18.016*HH2O
0035 CALL ENTH(T,HCH4,HN2,HCO2,HCO,HH2,HH2O)
0036      H2=XCH4*16.042*HCH4+XN2*28.*HN2+XCO2*44.01*HCO2+XCO*28.01*HCO+
1      XH2*2.016*HH2+XH2O*18.016*HH2O
0037 92 CONTINUE
0038      DHSENS=H1-H2
0039 94 CONTINUE
C
C   CALCULATE HEAT OF STEAM CARBON REACTION AND MOLES REACTED
C
0040      HCARB=305+2.1*TOUT+0.0008*TOUT**2+1.1E-6*TOUT**3+4.64E-10*TOUT**4
0041      HRSC=HCO*28.01+HH2*2.016-HCARB-HH2O*18.016+HFCO+HFH2-HFCARB-HFH2O
0042 95 CONTINUE
0043      HRSC=ABS(HRSC)
0044      RMOLES=DHSENS/HRSC
C
C   CHECK TO SEE IF SUFFICIENT CARBON AND WATER WERE AVAILABLE
C
0045      IF (RMOLES.GT.XH2O) GO TO 700
0046      IF (RMOLES.GE.XC) GO TO 600
0047      XCO=XCO+RMOLES
0048      XH2=XH2+RMOLES
0049      XH2O=XH2O-RMOLES
0050      XC=XC-RMOLES
0051 93 CONTINUE
C
C   CORRECT FOR SHIFT
C
0052      TK=(T+459)/1.8
0053      KP=EXP(5639.5/TK+1.077*ALOG(TK)+.000544*TK-1.125E-07*TK**2-49170.
1      /TK**2-13.148)
0054 20 A=1.-KP
0055      B=XCO2+XH2+XH2O*KP+XCO*KP
0056      C=XCO2*XH2-KP*XCO*XH2O
0057      TKP=XCO2*XH2/XCO/XH2O
0058 96 CONTINUE
C
C   CHECKING QUADRATIC ROOTS
C
0059      IF (TKP.EQ.KP) GO TO 1000
0060      IF (TKP.GT.KP) GO TO 800
0061      IF (B**2-4.*A*C.LT.0.0) GO TO 900
0062      IF (KP.EQ.1) GO TO 60
0063      IF (A.GT.0.0.AND.C.GT.0.0) GO TO 800
0064      IF (A.LT.0.0) GO TO 40
0065 25 XX = (-B+SQRT(B**2-4.*A*C))/2./A
0066      GO TO 100
C
C   SELECT SMALLER ROOT
    
```

B-2



```

C
0067 40 XXF=(-B+SQRT(B**2-4.*A*C))/2./A
0068 XXM=(-B-SQRT(B**2-4.*A*C))/2./A
0069 IF (XXF.LE.XXM) GO TO 50
0070 XX=XXM
0071 GO TO 100
0072 50 XX=XXF
0073 GO TO 100
0074 60 XX=(XH2*XCO-XCO2*XH2)/(XH2+XH2O+XCO+XCO2)
0075 100 CONTINUE

C
C CHECK TO SEE IF SUFFICIENT WATER WAS AVAILABLE FOR SHIFTING
C
0076 IF (XX.GE.XH2O) GO TO 500

C
C REVISE GAS AND CALC SHIFT ENTHALPY AVAILABLE
C
0077 XH2=XH2+XX
0078 XCO2=XCO2+XX
0079 XH2O=XH2O-XX
0080 XCO=XCO-XX
0081 HWGS=XX*(HCO2*44.01+HH2*2.016-HCO*28.01-HH2O*18.016+HFCO2+
1HFH2-HFCO-HFH2O)
0082 HWGS=ABS(HWGS)
0083 97 CONTINUE

C
C TOLERANCE FOR CLOSURE IS ONE PERCENT, TEN ITERATIONS
C ARE ALLOWED.
C
0084 IF ((HWGS/DHSENS).LE.,01) GO TO 1000
0085 IF (M.GT.10) GO TO 400

C
C CORRECT FOR SHIFT HEAT WITH CARBON-CO2 REACTION
C
0086 HCCO2=HCO*2.*28.01-HCO2*44.01-HCARB+HFCO*2.-HFCO2-HFCARB
0087 RMOLES=HWGS/ABS(HCCO2)
0088 98 CONTINUE
0089 M=M+1
0090 XCO2=XCO2-RMOLES
0091 XCO=XCO+2.*RMOLES

C
C CHECK FOR AVAILABILITY OF CARBON TO REACT
C
0092 IF (XC.LE.RMOLES) GO TO 600
0093 XC=XC-RMOLES
0094 GO TO 20

C
C INTERMEDIATE TEMPERATURE OUTPUT
C
0095 1000 WRITE(6,1001)T,XC,XH2,XH2O,XCO,XCO2
0096 1001 FORMAT(' TEMPERATURE (F) = ',F10.4/
1 ' MOLES CARBON = ',F10.4/
2 ' H2 = ',F10.4/

```

B-3

```

3      ' H2O          = ',F10.4/
4      ' CO           = ',F10.4/
5      ' CO2          = ',F10.4///)
0097   IF (I.EQ.1) GO TO 1100
0098   IF (K.LT.1) GO TO 3
0099   1100 TMOLES=XCH4+XN2+XH2+XH2O+XCO+XCO2
0100     YN2=XN2/TMOLES
0101     YCH4=XCH4/TMOLES
0102     YH2=XH2/TMOLES
0103     YH2O=XH2O/TMOLES
0104     YCO=XCO/TMOLES
0105     YCO2=XCO2/TMOLES
0106     GO TO 1105

C
C C   OUTPUT
C
C C   ERROR MESSAGES
C
0107   400 WRITE(6,401)T,XCO,XCO2,XH2,XH2O,XC,XX,HMGS,DHSENS
0108   401 FORMAT(/' *****'/2X,'NOT CONVERGING...           '/2X,
1'T,XCO,XCO2,XH2,XH2O,XC,XX,HMGS,DHSENS =' /9(F10.4,'|'))
GO TO 5000
0109   500 WRITE(6,501)T,XCO,XCO2,XH2,XH2O,XC,XX
0110   501 FORMAT(/' *****'/2X,'NOT ENOUGH MOLES H2O FOR SHIFT'/2X,
1'T,XCO,XCO2,XH2,XH2O,XC,XX =' /7(F10.4,'|'))
GO TO 5000
0111   600 WRITE(6,601)T,XCO,XCO2,XH2,XH2O,XC,RMOLES
0112   601 FORMAT(/' *****'/2X,'NOT ENOUGH CARBON FOR REACTION'/2X,
1'T,XCO,XCO2,XH2,XH2O,XC,RMOLES =' /7(F10.4,'|'))
GO TO 5000
0113   700 WRITE(6,701)T,XCO,XCO2,XH2,XH2O,XC,RMOLES
0114   701 FORMAT(/' *****'/2X,'NOT ENOUGH WATER FOR STM/CARBON RXN.'/2X,
1'T,XCO,XCO2,XH2,XH2O,XC,RMOLES =' /7(F10.4,'|'))
GO TO 5000
0115   800 WRITE(6,801)T,XCO,XCO2,XH2,XH2O,XC,KP,TKP,A,C
0116   801 FORMAT(/' *****'/2X,'NEGATIVE ROOT FOR SHIFT'/2X,
1'T,XCO,XCO2,XH2,XH2O,XC,KP,TKP,A,C =' /10(F10.4,'|'))
GO TO 5000
0117   801

C
C C   REPLACE STEAM-C WITH CO2-C REACTION TO AVOID NEGATIVE ROOT
C
0121   JJ=JJ+1
0122   IF(JJ.GT.1) GO TO 5000
0123   XC=CMASS/12.
0124   XCO2=GMOLES*YCO2
0125   XCO=GMOLES*YCO
0126   XH2=GMOLES*YH2
0127   XH2O=GMOLES*YH2O
0128   HHCO2=HCO*2.*28.01-HCO2*44.01-HCARB+HFCO*2.-HFCO2-HFCARB
0129   RMOLES=DHSENS/ABS(HHCO2)
0130   XCO=XCO+2.*RMOLES
0131   XCO2=XCO2-RMOLES
0132   IF(XC.LE.RMOLES) GO TO 600
0133   XC=XC-RMOLES

```

B-4

```

0134      GO TO 20
0135      900 WRITE(6,901)T,XCO,XCO2,XH2,XH2O,XC,KP,TKP,A,B,C
0136      901 FORMAT(/' *****'/2X,'NEGATIVE SQR. ROOT IN SHIFT QUAD.'/2X,
          1'T,XCO,XCO2,XH2,XH2O,XC,KP,TKP,A,B,C ='/11(F10.4,'|'))
0137      GO TO 5000
          C
          C      SUCCESSFUL OUTPUT
          C
0138      1105 WRITE(6,1101)YH2,YCO,YCO2,YH2O,YCH4,YN2
0139      1101 FORMAT(' FINAL GAS COMPOSITION'//
          1' YH2 = ',F10.4/
          2' YCO = ',F10.4/
          3' YCO2 = ',F10.4/
          4' YH2O = ',F10.4/
          5' YCH4 = ',F10.4/
          4' YN2 = ',F10.4///)
0140      5000 IF (K.LT.1) GO TO 3
0141      GO TO 12
0142      5010 STOP
          C
          C      DEBUG SEGMENT
          C      TO USE, REMOVE DEBUG AND FOLLOWING FROM COMMENTS
          C
          C      DEBUG
          C      AT 90
          C      DISPLAY TIN,YCH4,YN2,YCO2,YH2O,YCO,YH2,GMASS,CMASS,TOUT,I
          C      AT 91
          C      DISPLAY GMOLES,XCH4,XN2,XCO2,XH2,XH2O,XC,XCO
          C      AT 92
          C      DISPLAY TIN,HCH4,HN2,HCO2,HCO,HH2,HH2O
          C      AT 93
          C      DISPLAY RMOLES,XCH4,XN2,XCO2,XH2,XH2O,XC,XCO
          C      AT 94
          C      DISPLAY DHSENS,H1,H2
          C      AT 95
          C      DISPLAY HRSC,HCARB
          C      AT 96
          C      DISPLAY A,B,C,KP,TKP
          C      AT 97
          C      DISPLAY HWGS
          C      AT 98
          C      DISPLAY HCCO2,RMOLES
          C      AT 100
          C      DISPLAY RMOLES,XCH4,XN2,XCO2,XH2,XH2O,XC,XX ,XCO
0143      END

```

B-5


```
      C
0001  SUBROUTINE ENTH(TIN,HCH4,HN2,HCO2,HCO,HH2,HH2O)
0002  DIMENSION A(6),B(6),C(6),D(6),E(6),F(6),COMP(6)
0003  DATA A/ 58.4016, -.93401, 4.77805, -.97557, 12.32674, -2.46342/
0004  DATA B/ .5717, .255204, .114433, .256524, 3.199617, .457392/
0005  DATA C/-2.943122,-.177935,1.011325,-.229112, 3.927862, -.525117/
0006  DATA D/ 4.231568, .158913,-.264936, .222803,-2.934520, .645939/
0007  DATA E/ -15.2674,-.322032, .347063,-.563256, 10.90069,-2.027592/
0008  DATA F/ 19.45261, .158927,-.131400, .455878,-13.87867, 2.363096/
      C
0009      CH4      N2      CO2      CO      H2      H2O
0010      T=TIN+459.
0011      TREF=537.
0012      TB=(T-TREF)
0013      TC=(T**2-TREF**2)*1.0E-4
0014      TD=(T**3-TREF**3)*1.0E-7
0015      TE=(T**4-TREF**4)*1.0E-11
0016      TF=(T**5-TREF**5)*1.0E-15
      C
0016      DO 10 I=1,6
0017      COMP(I)=A(I)+B(I)*TB+C(I)*TC+D(I)*TD+E(I)*TE+F(I)*TF
0018      10 CONTINUE
0019      HCH4=COMP(1)
0020      HN2=COMP(2)
0021      HCO2=COMP(3)
0022      HCO=COMP(4)
0023      HH2=COMP(5)
0024      HH2O=COMP(6)
0025      RETURN
0026      END
```

B-6

Variable List - in order of appearance

CASE	=	alphanumeric for case title
TIN	=	initial temperature (°F)
YCH4	=	mole fraction CH ₄ in gas
YN2	=	mole fraction N ₂ in gas
YCØ2	=	mole fraction CO ₂ in gas
YH2Ø	=	mole fraction H ₂ O in gas
YCØ	=	mole fraction CO in gas
YH2	=	mole fraction H ₂ in gas
GMASS	=	gas mass flow (lb/unit time)
CMASS	=	char mass flow (lb/unit time)
TØUT	=	final temperature (°F)
I	=	number of temperature increments
K	=	counter for the number of temperature increments
HFCØ	=	heat of formation for CO (Btu/lb mole)
HFCØ2	=	heat of formation for CO ₂ (Btu/lb mole)
HFH2Ø	=	heat of formation for H ₂ O (Btu/lb mole gas)
HFH2	=	heat of formation for H ₂ (Btu/lb mole)
HFCARB	=	heat of formation for char (Btu/lb mole)
GMW	=	gas moles
XCH4	=	moles CH ₄
XN2	=	moles N ₂
XCØ2	=	moles CO ₂



XCØ = moles CO
 XH2 = moles H₂
 XH2Ø = moles H₂O
 XC = moles char
 M = counter for iterations for closure at any temperature T
 JJ = counter for negative root loops at any temperature T
 DELT = temperature increment (°F)
 T = intermediate temperature (between inlet and outlet) being used (°F)
 HCH4 = enthalpy of CH₄ at temperature = T (Btu/lb)
 HN2 = enthalpy of N₂ at temperature = T (Btu/lb)
 HCØ2 = enthalpy of CO₂ at temperature = T (Btu/lb)
 HCØ = enthalpy of CO at temperature = T (Btu/lb)
 HH2 = enthalpy of H₂ at temperature = T (Btu/lb)
 HH2Ø = enthalpy of H₂O at temperature = T (Btu/lb)
 H1 = enthalpy of gas at TIN (Btu)
 H2 = enthalpy of gas at T (Btu)
 DHSENS = sensible heat change of gas from TIN to T (Btu)
 HCARB = sensible heat of solid carbon at TØUT (Btu/mole)
 HRSC = steam-carbon reaction heat of reaction at T (Btu/mole)
 RMØLES = extent of reaction in moles
 TK = T in °K
 KP = shift equilibrium constant at T
 A,B,&C = quadratic equation coefficients

TKP = test value of shift equilibrium constant
 XX = moles of carbon monoxide shifted
 XXP = moles of CO shifted using $+\sqrt{B^2 - 4AC}$
 XXM = moles of CO shifted using $-\sqrt{B^2 - 4AC}$
 HWGS = heat of reaction of water gas shift occurring at T
 HCCØ2 = heat of reaction for carbon - CO₂ occurring at T

In Subroutine ENTH (and not in main program):

A,B,C,D,E&F - coefficients for polynomials expansions of components' enthalpies
 T = temperature in degrees Rankine
 TREF = reference temperature (°R) for enthalpy polynomials
 TB,TC,etc = temperature "differences" for enthalpy polynomials
 CØMP(I) = enthalpy polynomial expansions for component I



STONE & WEBSTER ENGINEERING CORPORATION
CALCULATION SHEET

▲ 5010.85

CALCULATION IDENTIFICATION NUMBER				PAGE <u>1</u>
J.O. OR W.O. NO.	DIVISION & GROUP	CALCULATION NO.	OPTIONAL TASK CODE	
14665				

DETAILS OF CORRECTING THE EXTENT
OF REACTION FOR SHIFT EQUILIBRIUM

Let a = current number of moles of CO_2
 b = current number of moles of H_2
 c = current number of moles of CO
 d = current number of moles of H_2O
 x = change in moles to reach shift equilibrium
 γ = shift equilibrium value = $f(\text{temp})$

Then

$$\frac{(a+x)(b+x)}{(c-x)(d-x)} = \gamma = KP (\text{in program})$$

$$ab + bx + ax + x^2 = \gamma (cd - xd - xc + x^2)$$

$$ab + bx + ax + x^2 - \gamma cd + \gamma xd + \gamma xc - \gamma x^2 = 0$$

$$x^2(1-\gamma) + x(a+b+\gamma c+\gamma d) + ab - \gamma cd = 0$$

Quadratic constants are

$$A = 1-\gamma$$

$$B = a+b+\gamma c+\gamma d$$

$$C = ab - \gamma cd$$

and

$$x = \frac{-B \pm \sqrt{B^2 - 4AC}}{2A}$$

STONE & WEBSTER ENGINEERING CORPORATION
 CALCULATION SHEET

▲ 5010.65

CALCULATION IDENTIFICATION NUMBER				PAGE <u>2</u>
J.O. OR W.O. NO. 14665	DIVISION & GROUP	CALCULATION NO.	OPTIONAL TASK CODE	

Program checks to avoid imaginary roots, etc:

1) Check for $x=0$. Let TKP (in program) = $\frac{(a)(b)}{(c)(d)}$

If $\gamma = KP = TKP$ then $x = 0$

2) Test for negative root, i.e. $x < 0$ and the shift correction must be endothermic. This is physically impossible and the steam-carbon reaction used before should be replaced by the CO_2 -carbon reaction. A negative root will occur if $TKP > KP$, i.e. shift equilibrium occurs when the moles of CO_2 and H_2 are reduced. If $TKP > KP$, go to statement 800.

3) Test for imaginary roots. If

$B^2 - 4AC < 0$ Go to statement 900

4) Check denominator = 0, i.e. $\gamma = 1, A = 0$

If $KP = 1$ $x = \frac{cd - ab}{a+b+c+d}$ (Statement 60)

STONE & WEBSTER ENGINEERING CORPORATION
CALCULATION SHEET

▲ 5010.65

CALCULATION IDENTIFICATION NUMBER				PAGE <u>3</u>
J.O. OR W.O. NO. 14665	DIVISION & GROUP	CALCULATION NO.	OPTIONAL TASK CODE	

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5) 2nd test for negative root:

A) By definition $\gamma > 0$ and $B > 0$. If
 $\gamma < 1$; $A > 0$. $C < 0$ for a positive root, i.e.

$$B < \sqrt{B^2 - 4AC}$$

\therefore If $A > 0$ and $C > 0$, $x < 0$

Go to statement 800

B) If $\gamma > 1$, $A < 0$

If $C < 0$, $\sqrt{B^2 - 4AC} < B$ and $x > 0$

If $C > 0$, $\sqrt{B^2 - 4AC} > B$ and use

smaller root:

$$x = (-B - \sqrt{B^2 - 4AC}) / 2A$$

This is statement 40

6) Select smaller positive root:

A) If $A > 0$, $C < 0$, use $+\sqrt{B^2 - 4AC}$

B) If $A < 0$, smaller root will depend on C

\therefore Simply calculate roots, compare and
 use the smaller root.

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16. Abstract (Limit: 200 words) This report presents an economic feasibility study for a wood-gasification-based methanol plant. The objectives were to evaluate the current commercial potential of a small-scale, wood-fed methanol plant using the SERI oxygen-blown, pressurized, down-draft gasifier technology and to identify areas requiring further R&D. The gasifier gas composition and material balance were based on a computer model of the SERI gasifier since acceptable test data were not available. The estimated capital cost was based on the Nth plant constructed. Given the small size and commercial nature of most of the equipment, N was assumed to be between 5 and 10. Only large discrepancies in gasifier output would result in significant charges in capital costs.			
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