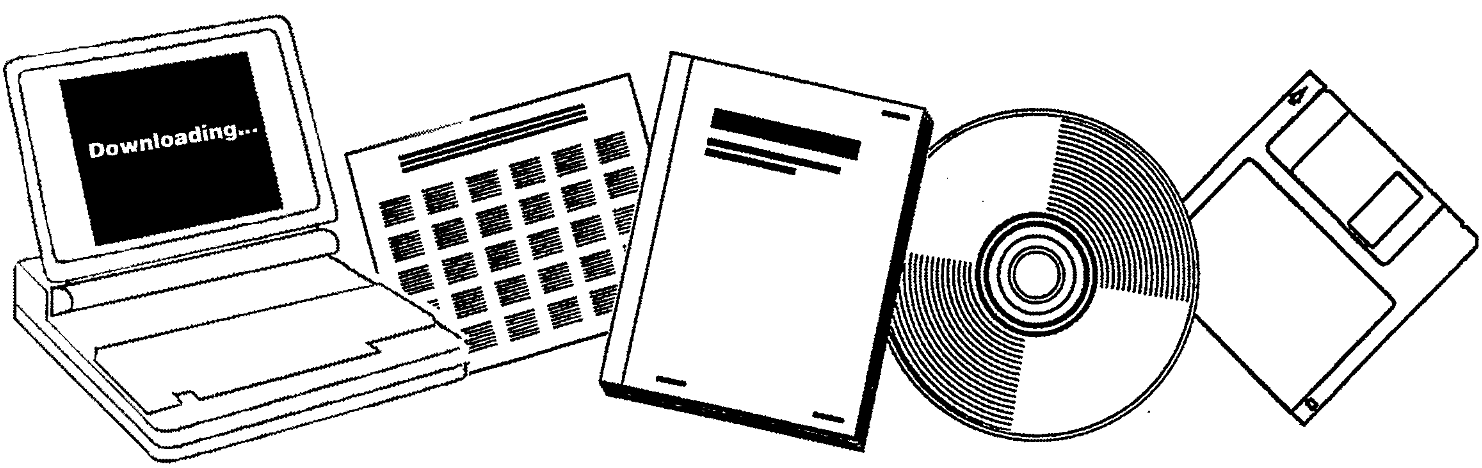
FE248020

LARGE PILOT PLANT ALTERNATIVES FOR SCALEUP OF THE CATALYTIC COAL GASIFICATION PROCESS. FINAL REPORT

EXXON RESEARCH AND ENGINEERING CO.
FLORHAM PARK, NJ

JAN 1979



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LARGE PILOT PLANT ALTERNATIVES FOR SCALEUP OF THE CATALYTIC COAL GASIFICATION PROCESS

Final Report

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Submitted - January, 1979

PREPARED FOR
UNITED STATES DEPARTMENT OF ENERGY
UNDER CONTRACT NUMBER EX-76-C-01-2480

LARGE PILOT PLANT ALTERNATIVES FOR
SCALEUP OF THE CATALYTIC COAL GASIFICATION PROCESS
FINAL REPORT

ABSTRACT

This is the final report for U.S. Department of Energy Contract No. EX-76-C-01-2480, "Scaleup Requirements of the Exxon Catalyzed Coal Gasification Process." The objective of this contract was to develop the information necessary to determine if an existing DOE large pilot plant could be used to obtain the scaleup data necessary to design and construct a Catalytic Coal Gasification (CCG) pioneer plant with acceptable risk. A pioneer plant is a stand-alone facility, whose primary function is to operate as a profitable commercial venture. The pioneer plant would contain all equipment of full commercial size, as defined by the requirements for an optimum-sized commercial plant. However, the pioneer plant could have a single train of equipment in some or all of the plant sections. The results of the three tasks contained in this contract are summarized below.

- Task 1 - Study Design and Cost Estimate for a Grass-Roots Large Pilot Plant

The objective of this task was to define a base case for evaluation of the existing large pilot plants. The capacity of the grass-roots large pilot plant was calculated to be 92 T/SD of Illinois No. 6 coal based on the scaleup requirements of the gasifier. The inside diameter of the gasifier was set at 3.5 feet to assure operation in the bubbling flow regime. The bed height was set at 100 feet. This is close to the height projected for a commercial gasifier. Also, facilities were included to allow operation with both synthesis gas and catalyst recycle. Hardware and system backups were also included in both onsite and offsite areas to promote the achievement of a high service factor. The investment for the grass-roots case at a Gulf Coast location was estimated to be 130 M$. The investment for the Gulf Coast grass-roots facilities on a Pittsburgh location basis is 150 M$. These investments are in escalated dollars, assuming completion of the basic design in mid-1980, with an estimated plant startup date of the 4th quarter of 1982.

- Task 2 - Selection of the Preferred Existing Pilot Plant

This task consisted of an evaluation of three existing large pilot plants to select the one most adaptable to catalytic gasification. The three existing large pilot plants considered were Synthane, Hygas, and Steam Iron. Synthane was selected as the unit most suitable for conversion to CCG operation. The Synthane site had the most usable plot space and the most reusable equipment.

- Task 3 - Study Design and Cost Estimate for Revamp of the Preferred Existing Pilot Plant

This task consisted initially of a study design and cost estimate for a major revamp of the Synthane LPP so that it would have all the features of the grass-roots pilot plant. The objective

was to meet all of the major scaleup needs as in the grass-roots case and to provide a service factor similar to the grass-roots case. The investment for the major Synthane revamp was estimated to be 150 M$ at a Pittsburgh location. The equipment savings for the revamp were offset by reduced labor productivity associated with the revamp.

At the completion of the major revamp case, the DOE contract was extended to develop a minimum modification revamp case. The objective was to develop a cost and schedule estimate for a case where Synthane was converted to CCG operation, in a technically meaningful fashion, but with minimum modification of the existing facilities. For this case, the maximum coal feed rate was calculated to be 55 T/SD based on the capacity of the existing steam system. The existing Synthane gasifier bed diameter is 3.5 feet, the same diameter as specified for the grass-roots case. The bed height was increased to 100 feet, also the same height as in the grass-roots case. Also, facilities for catalyst recovery and recycle were included. Since the preferred catalyst recovery approach will not be fully defined until after the Process Development Unit (PDU) operates, it was assumed that addition of catalyst recycle to the LPP would be done during a second construction stage.

On the other hand, facilities for the separation and recycle of CO/H_2 from the product were not included since this would require major equipment additions and site changes. Although this compromise would directionally increase the risk of scaleup to a pioneer plant, it was judged that if the LPP operates with an adequate service factor, the scaleup risk still would be acceptable. Perhaps of greater significance is the fact that for the minimum revamp much of the equipment redundancy and operating flexibility build into the grass-roots and major revamp cases was eliminated. Thus the plant service factor for the minimum revamp would be less than the service factor for the grass-roots or major revamp cases. As a result there is some risk that the large pilot plant would not be able to operate at steady-state for sufficiently long periods of time to obtain adequate scaleup data, and that, consequently, significant additional facilities modifications would be required with additional cost.

The estimated minimum revamp investment is 46 M$ for the gasification section and 12 M$ for the later addition of catalyst recovery, for a total investment of 58 M$. This represents a substantial investment saving relative to the grass-roots and major revamp cases. Again, these investments are in escalated dollars, assuming completion of the basic design for the gasification section by mid-1979, with an estimated startup in the first quarter of 1981. Delays in this schedule would result in somewhat higher costs due to further escalation.

Since the objective of this contract was to develop cost and schedule information for scaleup options and not to set the design basis for an actual large pilot plant project, a number of simplifying assumptions were made. These assumptions would have to be verified before proceeding with the design basis for an actual project. The major assumptions made were that equipment reused at Synthane would have adequate life remaining for CCG operations, that all existing Synthane utility systems can operate at their nameplate capacity, and that the utilities usage of existing Synthane facilities is accurately represented by information on the original design drawings. In addition, since these studies were not based on data from an integrated Process Development Unit (PDU), it was assumed that data from the Process Development Phase of research would not change the scaleup needs determined from the Predevelopment Phase data base used for the study.

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APPENDIX 2 Details of Study Design for Major Synthane Revamp Case

APPENDIX 3 Details of Study Design for Minimum Synthane Revamp Case

SECTION 1

BACKGROUND AND SUMMARY

1.1 INTRODUCTION

Exxon Research and Engineering Company (ER&E) is engaged in research and development on Catalytic Coal Gasification (CCG) for the production of substitute natural gas (SNG). The Predevelopment phase of catalytic gasification research was sponsored by DOE (Contract No. E(49-18)-2369) and covered the time period from July, 1976 to December, 1977. The Process Development phase of CCG research began in July, 1978, again under DOE sponsorship. In addition to bench-scale research and engineering studies, this program includes the operation of an integrated 1 T/D Process Development Unit.

Early in 1976, DOE (then ERDA) identified the possibility of using an existing large pilot plant (LPP) for the scaleup of CCG. Contract No. EX-76-C-01-2480, "Scaleup Requirements of the Exxon Catalyzed Coal Gasification Process" was undertaken by ER&E to study alternative scaleup routes for CCG. The program was conducted over the period November 1, 1976 to September 30, 1978.

1.2 CONTRACT OBJECTIVES

The objective of this contract was to develop the information necessary to determine if an existing DOE large pilot plant could be used to obtain the scaleup data necessary to design and construct a Catalytic Coal Gasification (CCG) pioneer plant with acceptable risk. A pioneer plant is a stand-alone facility, whose primary function is to operate as a profitable commercial venture. The pioneer plant would contain all equipment of full commercial size, as defined by the requirements for an optimum-sized commercial plant. However, the pioneer plant could have a single train of equipment in some or all of the plant sections. The contract consisted originally of three tasks:

- Task 1 - Study design and cost estimate for a conceptual grass-roots large pilot plant (LPP).
- Task 2 - Evaluation of three existing pilot plants (Synthane, Hygas, and Steam-Iron) to select the one most adaptable to catalytic gasification.
- Task 3 - Study design and cost estimate for converting the pilot plant selected in Task 2 to catalytic gasification.

The objective of Task 1 was to define a base case for evaluation of the existing large pilot plants. The objective of Task 2 was to select the existing LPP most suitable for conversion to CCG. In Task 3, cost and schedule estimates were developed for a major revamp of the selected existing LPP so that it would have all the features of the grass-roots pilot plant. Task 3 was subsequently expanded under a contract extension to include a study of the possibility of making minimum modifications to the selected LPP which would still allow scaleup to a pioneer plant with reasonable risk.

1.3 CCG PROCESS CONCEPT

The catalysts being studied for catalytic coal gasification are the weak acid salts of potassium. The principal benefits from using potassium catalyst in a gasification reactor system are as follows: first, it increases the rate of gasification; second, it prevents swelling and agglomeration when handling caking coals; third, and most important, it promotes gas phase methanation equilibrium. These key features of the catalyst are combined in a novel processing sequence which maximizes their benefit. A schematic flow plan for this processing sequence is shown in Figure 1-1. Catalyst is added to the feed coal and the mixture is gasified at about 1300°F and 500 psia. At these conditions, the gasification rates are high enough to allow reasonable size commercial reactor vessels while at the relatively low temperature, equilibrium favors the formation of methane. Thus, the production of CO and H_2 is decreased, and high direct methane yields can be achieved. The components in the gasifier overhead are separated into CO_2 which is vented, product methane, and carbon monoxide and hydrogen which are recycled to the gasification stage. Since the amount of CO and H_2 fed balances the amount of CO and H_2 leaving the gasifier, the net products of gasification are only methane and CO_2, along with smaller amounts of H_2S and NH_3. The chemistry of this reaction can be represented as follows:

$$\text{Coal} + H_2O = CH_4 + CO_2 \qquad \Delta H \approx 0$$

As indicated, this reaction is thermally neutral; and, in fact, only a small amount of heat is required in the gasifier to preheat the feed coal and provide for heat losses. Also shown on the flow diagram is a catalyst recovery step. This is required because the catalyst leaving the gasifier with the ash/char residue is too costly to discard.

The unique features of the Catalytic Coal Gasification (CCG) process can be summarized as follows: (1) all the methane is formed in one reactor, the gasifier; no separate shift and methanation reactors are required, (2) no significant heat input is required to the reactor; the oxygen plant and potential slagging problems from oxygen use are eliminated, (3) no pretreatment is required for caking coals due to the action of the catalyst, and (4) significant future improvements are possible through the development of improved catalysts.

1.4 FACILITIES DESCRIPTION FOR A COMMERCIAL-SCALE CCG PLANT

As indicated above, the objective of a CCG large pilot plant program would be to provide scaleup data for the design of a CCG pioneer commercial plant. The current concept for a full-size commercial CCG plant is described below.

1.4.1 Catalyst Addition, Recovery, and Recycle

A simplified flow diagram for the commercial system envisioned for catalyst addition, recovery, and recycle is shown in Figure

1-2. Coal is crushed to minus 8 mesh and is dried with circulating flue gas in an entrained system. Catalyst is then added to the dried coal in a gentle mixing step. The catalyst is a solution of potassium hydroxide in water. A small makeup of purchased KOH is required to supplement that which has been recovered and recycled. The mixture is then dried before being fed via a lock hopper system to the fluidized bed gasifier operating at 1275°F and 500 psig.

In the catalyst recovery system, char withdrawn from the bottom of the gasifier and part of the fines entrained overhead are slurried, mixed with $Ca(OH)_2$, and soaked at 300°F for two hours. This "digestion" step frees additional water soluble catalyst such that about 90% of the catalyst is recovered in a downstream staged counter-current washing operation. The balance of the catalyst leaves the plant in the form of water insoluble compounds. For this study, the solid-liquid separation design was based on the use of hydroclones. A major objective of the next stage of research will be to obtain more data on the catalyst recovery system and identify the preferred recovery hardware.

1.4.2 Gas Cleanup and Synthesis Gas Recycle

Figure 1-3 presents a simplified flow diagram for the synthesis gas recycle system envisioned for a commercial CCG plant. The reactor is fluidized with a preheated mixture of steam and recycled hydrogen and carbon monoxide. The coal is fed to the bottom of the fluidized bed, and the residence time is sufficient at 1275°F with catalyst to gasify 90% of the feed carbon. Pyrolysis products are cracked, and essentially no hydrocarbons heavier than methane leave the gasifier. Since the gasifier exit temperature is only 1275°F and heavy hydrocarbons are present in only ppm quantities, the high level sensible heat in the overhead gas can be recovered and used for steam/recycle gas preheat and for high pressure steam generation. A venturi scrubber is used for fines removal prior to low pressure steam generation. H_2S and CO_2 are removed using a physical solvent acid gas removal system. At this point, the stream contains only H_2, CO, and CH_4. The methane is separated in a simple cryogenic distillation system and sent to the pipeline. The CO and H_2 are mixed with gasification steam, preheated to about 1550°F, and recycled to the bottom of the gasifier. The sensible heat above 1275°F in the steam/recycle gas mixture provides all the heat required in the gasifier to compensate for heat losses and for coal preheat. As mentioned above, the overall gasification reaction is essentially thermoneutral.

1.4.3 Fluid Bed Gasifier

A sketch of the catalytic gasifier itself is shown in Figure 1-4. The coal is fed to the gasifier via a lock hopper system which pressures the coal to the gasification pressure of 500 psia. Injection gas picks up the feed coal and conveys it in dense phase to the gasifier and injects it into the bottom of the bed. A number of coal injection points are used to assure good mixing and distribution of coal into the bed. The feed coal pyrolyzes rapidly, and the pyrolysis products flow up through the bed where they are cracked to light gaseous products. The bed dimensions for each of four gasifiers are 22 feet inside diameter by 97 feet in height.

The feed system and recycle synthesis gas are injected into the bottom of the bed through a distributor. Thus, the gasification medium also fluidizes the bed. The principal reactions taking place are the highly endothermic steam gasification reaction, the slightly exothermic water gas shift reaction, and the highly exothermic methanation reaction. The fluidized bed is characterized by the existence of a continuous emulsion phase with intimate gas solids contact and with gas bubbles rising up through the emulsion phase. Since steam enters the bed in bubbles, it must be transferred into the emulsion to react with the carbon. CO and H_2 from the recycle gas are also transferred across the bubble-emulsion interface to react via the catalytic action of the catalyst-char combination to form methane. The reaction rate in the gasifier is primarily kinetically limited, although mass transfer effects are not insignificant.

The top section of the vessel contains a deentrainment zone and external cyclones. The use of internal cyclones is an option that could be investigated. At the bottom of the bed, a solids stream is withdrawn to control bed level and prevent the buildup of ash. This solids stream flows into a small fluidized bed where it is cooled with recycle synthesis gas and then into a vessel where it is slurried with water for feed to catalyst recovery.

1.5 SUMMARY OF RESULTS

A summary of the key contract results is presented in Table 1-1. The capacity of the grass-roots large pilot plant (LPP) was calculated to be 92 T/SD of Illinois No. 6 coal (as-received basis) based on the scaleup requirements of the gasifier. The inside diameter of the gasifier was set at 3.5 feet to assure operation in the bubbling flow regime. The bed height was set at 100 feet. This is close to the height projected for a commercial gasifier. Also, facilities were included to allow operation with both synthesis gas and catalyst recycle. Hardware and system backups were also included in both onsite and offsite areas to promote the achievement of a high service factor. As shown in Table 1-1, the investment for the grass-roots case at a Gulf Coast location was estimated to be 130 M$. The investment for the Gulf Coast grass-roots facilities on a Pittsburgh location basis is 150 M$. These investments are in escalated dollars, assuming completion of the basic design in mid-1980, with an estimated plant startup date of the 4th quarter of 1982.

With regard to the comparison of existing LPP's, Synthane was selected over Hygas and Steam-Iron as the unit most suitable for conversion to CCG operation. The Synthane site had the most usable plot space and the most reusable equipment. A study design and cost estimate were then completed for a revamp of the Synthane LPP for CCG operation. The objective was to meet all of the major scaleup needs as in the grass-roots case and to provide a service factor similar to the grass-roots case. The investment for the major revamp of Synthane was estimated to be 150 M$ at a Pittsburgh location. The equipment savings for the revamp were offset by reduced labor productivity associated with the revamp.

At the completion of the major revamp study, the DOE contract was extended to develop a minimum modification revamp case. The objective was to develop a cost and schedule estimate for a case where Synthane was converted to CCG operation in a technically meaningful fashion but with minimum modification of the existing facilities. For this case, the maximum coal feed rate was calculated to be 55 T/SD based on the capacity of the existing steam system. The existing Synthane gasifier bed diameter is 3.5 feet, the same diameter as specified for the grass-roots case. The bed height was increased to 100 feet, also the same height as in the grass-roots case. Also, facilities for catalyst recovery and recycle were included. Since the preferred catalyst recovery approach will not be fully defined until after the PDU operates, it was assumed that addition of catalyst recycle to the LPP would be done during a second construction stage.

On the other hand, facilities for the separation and recycle of CO/H_2 from the product were not included since this would require major equipment additions and site changes. Although this compromise would directionally increase the risk of scaleup to a pioneer plant, it was judged that if the LPP operates with an adequate service factor, the scaleup risk still would be acceptable. Perhaps of greater significance is the fact that for the minimum revamp much of the equipment redundancy and operating flexibility build into the grass-roots and major revamp cases was eliminated. Thus the plant service factor for the minimum revamp would be less than the service factor for the grass-roots or major revamp cases. As a result there is some risk that the large pilot plant would not be able to operate at steady-state for sufficiently long periods of time to obtain adequate scaleup data, and that, consequently, significant additional facilities modifications would be required with additional cost.

The estimated minimum revamp investment is 46 M$ for the gasification section and 12 M$ for the later addition of catalyst recovery, for a total investment of 58 M$. This represents a substantial investment saving relative to the grass-roots and major revamp cases. Again, these investments are in escalated dollars, assuming completion of the basic design for the gasification section by mid-1979, with an estimated startup in the first quarter of 1981. Delays in this schedule would result in somewhat higher costs due to further escalation.

1.6 QUALIFICATIONS OF RESULTS

The objective of this contract was to develop cost and schedule information for scaleup options, not to set the design basis for an actual large pilot plant project. As such, a number of simplifying assumptions were made which would have to be verified before proceeding with a design basis:

- It was assumed that all Synthane facilities to be reused would have adequate life for operation as a CCG large pilot plant.

- It was assumed that all utility systems could be safely operated at their nameplate capacity.

- It was assumed that the utilities usage of existing Synthane facilities which are reused is accurately represented by information on the original design drawings.

It should also be noted that the study design is not based on data from an integrated process development unit operating at commercial conditions. Such a unit will be operated in the CCG Process Development Program. It is possible that this additional research and engineering could result in significant process changes affecting the design of a large pilot plant. Furthermore, the grass-roots and major Synthane revamp cases are based on the information available prior to the DOE Predevelopment Contract and thus do not have the benefit of that work. On the other hand, the information from the Predevelopment Contract was incorporated into the minimum Synthane revamp case.

TABLE 1-1

SUMMARY OF KEY RESULTS

	Grass Roots		Synthane Revamp: Major Revamp	Synthane Revamp: Minimum Revamp
• Project Type	Grass Roots		Major Revamp	Minimum Revamp
• Location	------Baytown, Texas-------		-------------Pittsburgh, Pa.------------	
• Labor Cost/Productivity Basis	Gulf Coast	Pittsburgh, Pa.	-------------Pittsburgh, Pa.------------	
• **Design Basis**				
+ Coal Feed Rate, T/D Illinois No. 6 (as rec.)	------------92------------		92	55
+ Gasifier Bed Dimensions	------3 1/2' ID x 100'-----		3 1/2' ID X 100'	3 1/2' ID X 100'
+ Catalyst Recycle Loop Closed?	-----------Yes------------		Yes	Yes
+ Synthesis Gas Recycle Loop Closed?	-----------Yes------------		Yes	No
• **Project Scope**				
+ Relocation and Expansion of Existing Facilities	------Not Applicable-------		Major	Minimal
+ Backup for Utility Systems	-----------Major-----------		Major	Minimal
+ Increase Required in Plot Area	------Not Applicable-------		+50%	Minimal
• **Approximate Costs**				
+ Investment, M$	130	150	150	46 Gasification 12 Catalyst Recovery
+ Operating Cost, M$	75	75	80	52
+ Total Program Cost, M$	205	225	230	110
• **Timing**				
+ Design Basis Memorandum Preparation	----------3Q79------------		3Q79	4Q78 Gasification 2Q80 Catalyst Recovery
+ Startup	----------4Q82------------		1Q83	1Q81 Gasification 2Q82 Catalyst Recovery

FIGURE 1-1

EXXON CATALYTIC GASIFICATION PROCESS

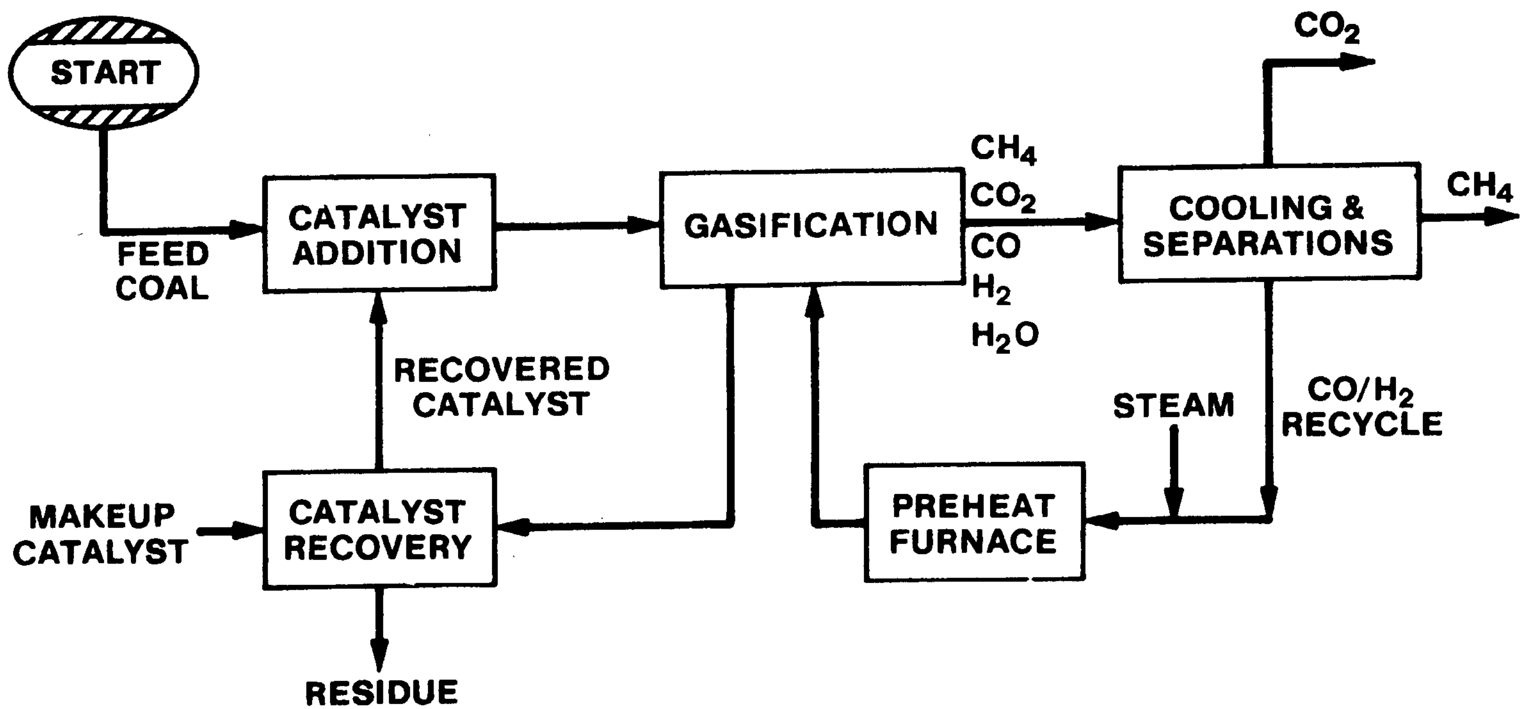


FIGURE 1-2

FLOWPLAN FOR CATALYST ADDITION, RECOVERY, AND RECYCLE

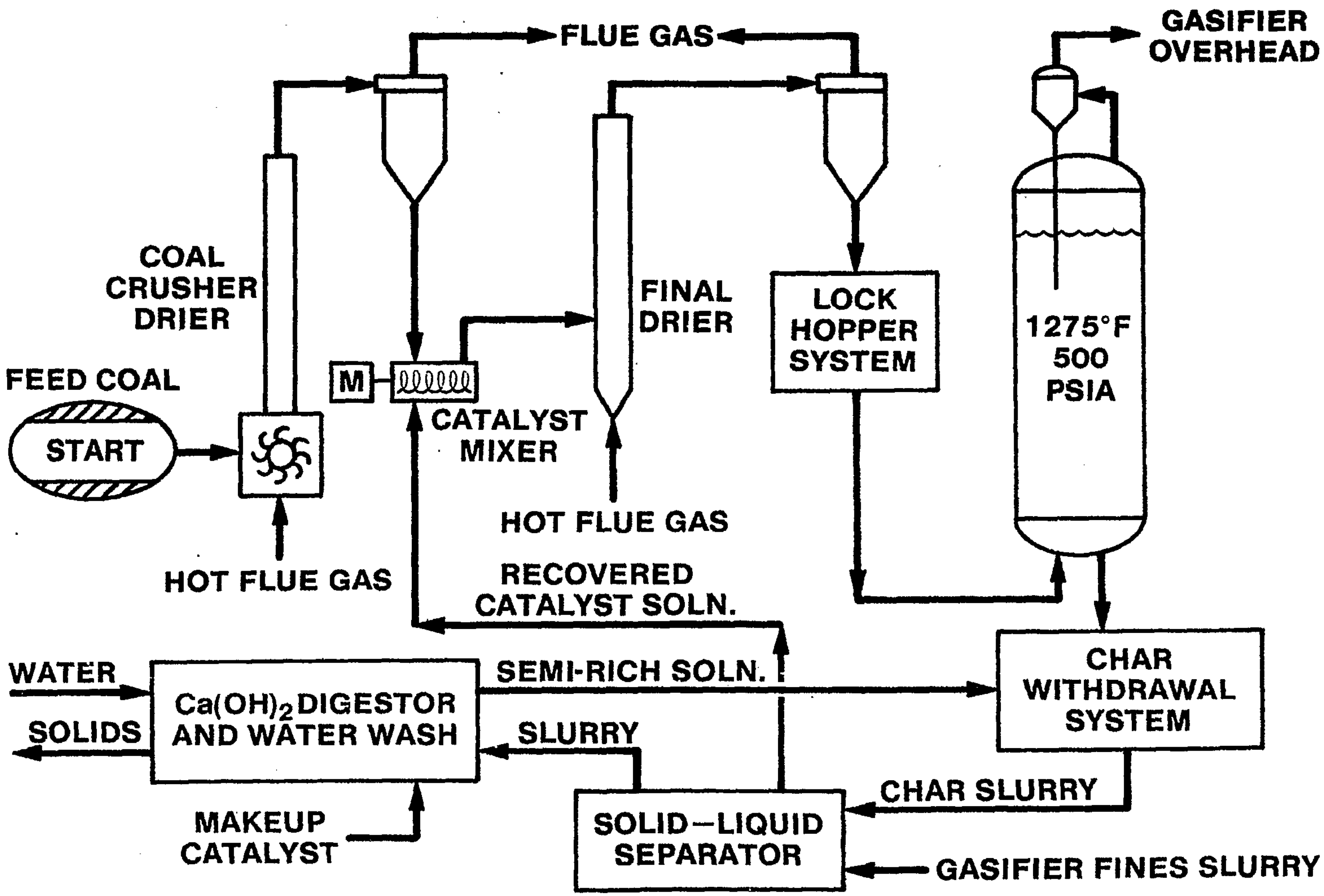


FIGURE 1-3

FLOWPLAN FOR RECYCLE GAS LOOP

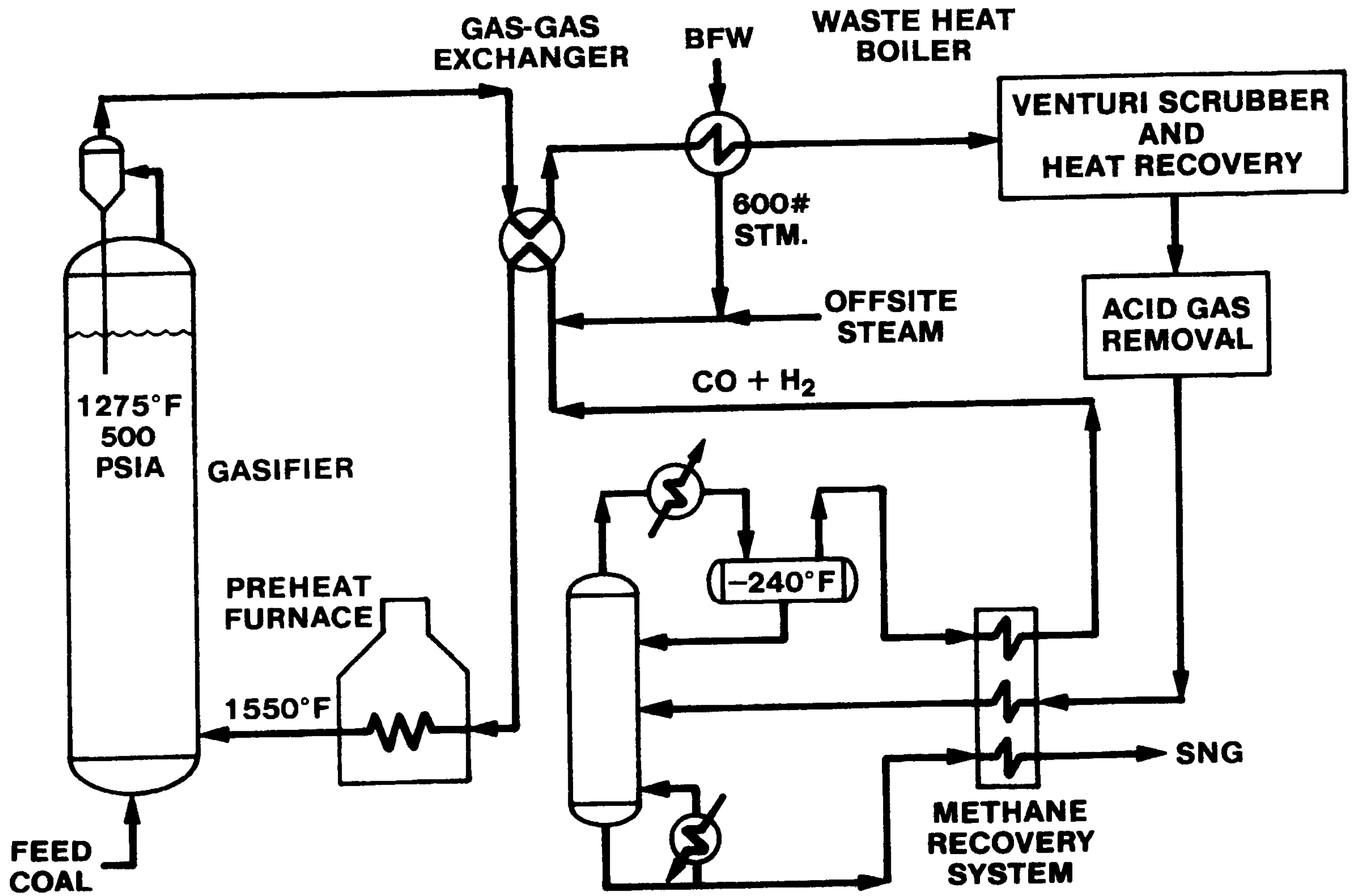
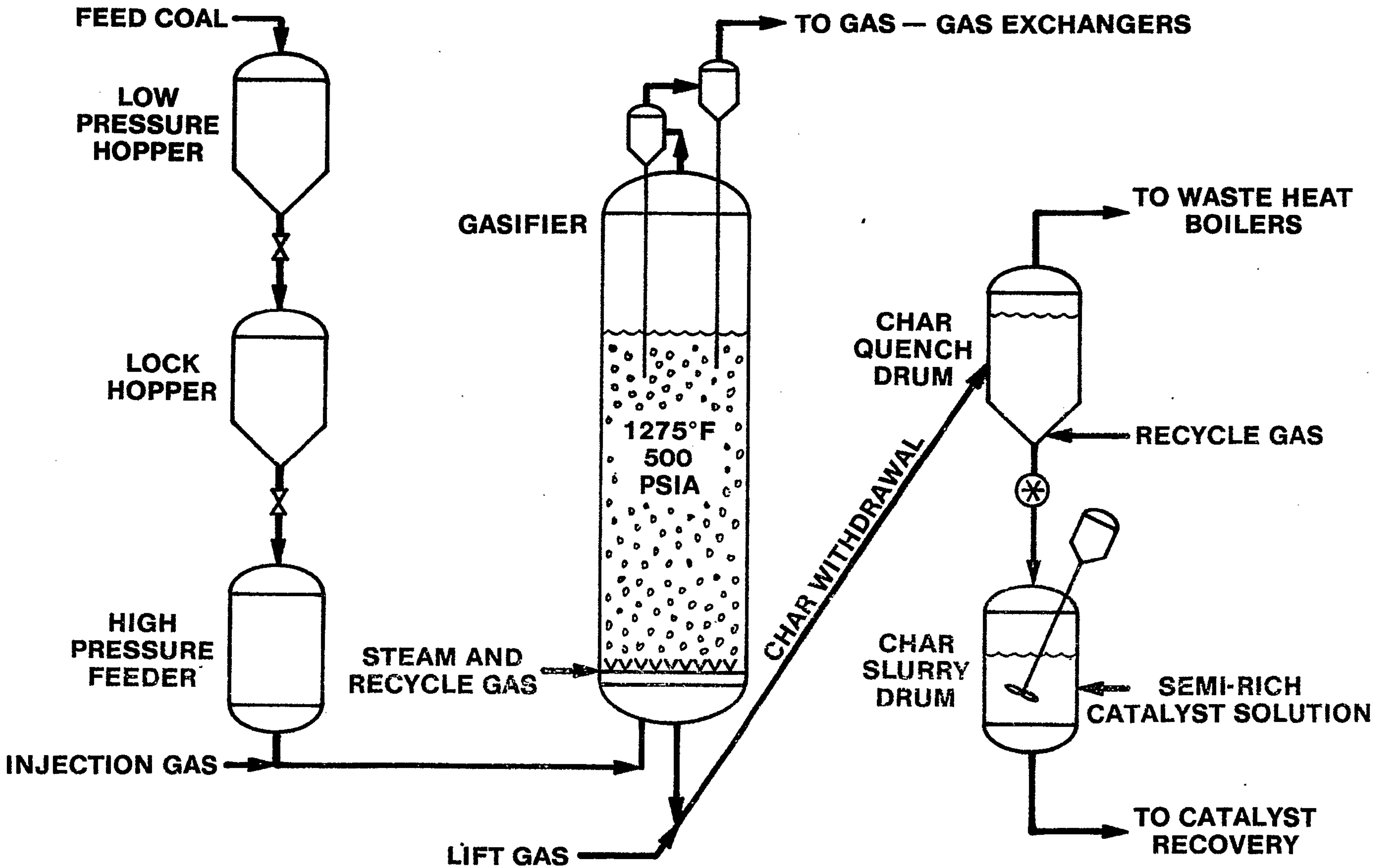


FIGURE 1-4

REACTOR SYSTEM SKETCH



SECTION 2

STUDY DESIGN AND COST ESTIMATE FOR A GRASS-ROOTS LARGE PILOT PLANT

2.1 PROJECT BASIS

The Grass-Roots case is based on construction at a site adjacent to a large Gulf Coast refinery. For purposes of this study, the site was assumed to be adjacent to the Exxon, U.S.A. refinery at Baytown, Texas. As such, the refinery is assumed to supply certain utilities and to accept certain waste streams from the large pilot plant (LPP).

The design feed coal was Illinois No. 6. The analysis used for this study is presented in Table 2-1.

2.2 PROCESS BASIS

At the start of this contract, the CCG process was just entering the predevelopment stage. The process basis for the grass-roots case was thus based on data obtained prior to the DOE Predevelopment Contract. This consisted mainly of bench-scale data with limited operation of a fluid bed gasifier at 10 lb/hr coal feed rate and 100 psia operating pressure. Catalyst had not yet been recycled to the gasifier. Normally, the design of a large pilot plant is based on the data from a Process Development Unit. Thus, considerable judgement was required in establishing the process basis for the LPP. In general, the philosophy used was to provide sufficient flexibility to handle uncertainties in the limited data base.

2.3 DESIGN PHILOSOPHY

The purpose of the catalytic gasification large pilot plant as defined in the contract was to obtain scaleup data to permit design of the commercial plant with acceptable risk. This requirement was used to set the design basis with regard to the type, size, and operating conditions of facilities included in the large pilot plant. In general, the areas of new technology are gasification, catalyst addition, and catalyst recovery.

To perform its function, the LPP must have a reasonable service factor. To achieve this, flexibility and equipment redundancy were built into the pilot plant. Following are some of the features included in the Grass-Roots case to help provide for reliable operation:

- The LPP is designed to operate both with and without synthesis gas recycle. When the recycle gas loop is not in operation, simulated recycle gas is manufactured in a steam reformer feeding purchased methane.

- The LPP is designed to operate with 100% makeup catalyst during startup and periods when there are operating problems in the catalyst recovery loop.

- The coal feed lock hopper is provided with a 100% spare.

- The LPP is designed to operate with or without the high temperature Gas-Gas Exchanger.

- The LPP is designed for the gasifier raw gas to bypass the Gas-Gas Exchanger, High Pressure Waste Heat Boiler, and dry fines removal system. For this operation, the raw gas goes directly to the wet scrubbers.

- The steam reformers are decoupled from the gasifier by cooling and condensing steam from the reformer effluent. This allows independent control of the reformers and gasifier.

- Substantial equipment redundancies are provided in the offsites and utilities to minimize the impact of offsite and utilities equipment outages on the operation of the onsite process sections of the plant.

In addition to the above process features, spares were provided for all critical pumping services. Materials of construction are based on a five-year plant life.

2.4 LPP SIZE - KEY SCALEUP ISSUES

As indicated above, the catalytic gasification large pilot plant must be large enough to allow scaleup to a commercial size plant with acceptable risk. The gasifier was determined to be limiting from a scaleup viewpoint. The diameter of the gasifier was set at 3.5 ft. ID so that it would operate well within the bubbling flow regime which is projected for the commercial gasifier. The bed height was set at 100 ft., the expected bed height for a commercial gasifier. With this bed height, the gas velocity will be close to that expected for commercial operation. This will provide representative scaleup data in the areas of entrainment, solids segregation, temperature distribution, and particle size distribution. Based on bed dimensions of 3.5 ft. ID x 100 ft. bed, the feed rate of as-received Illinois No. 6 coal was calculated to be 9[illegible] T/D.

The catalyst recovery and recycle loop were included in the LPP to allow integrated operation with the gasifier. This will generate data on catalyst forms and reactions, buildup of impurities, particle properties and size distributions, and the performance of solid-liquid separation equipment at commercial conditions. An integrated operation is necessary because the performance of the catalyst recovery system is strongly affected by the nature of the char feed stream which comes to it from the gasifier.

The synthesis gas recycle loop was included in the LPP to allow integrated operation with the gasifier. This will provide data on control of the gasifier under closed-loop operation. It also assures synthesis gas recycle to the gasifier with a commercially-representative composition. The cryogenic distillation system does not require data from the large pilot plant for scaleup, other than analysis of trace components in the feed.

2.5 GRASS-ROOTS CASE FLOW PLAN BASIS

A coordination flow plan for the grass-roots case is presented in Figure 2-1. This shows the onsite flow scheme and major equipment pieces. The coal is pressurized and fed to the gasifier via a lock hopper system. The gasifier is a fluid bed 3.5 ft. ID x 100 ft. with primary and secondary cyclones to limit fines loss. The gas cooling and treating system includes a gas-gas exchanger, high pressure waste heat boiler, and tertiary cyclone to withdraw fines for feed to catalyst recovery. A venturi scrubber is used for final solids removal. After final cooling, NH_3 is scrubbed out, followed by H_2S and CO_2 removal. H_2 and CO are then separated from the product methane in a cryogenic distillation system and are recycled to the gasifier. Prior to reinjection to the gasifier, the synthesis gas is mixed with steam and heated in the gas-gas exchanger and preheat furnace.

The LPP also contains a steam reformer to allow operation without syngas recycle. In this case, the design coal feed rate is the same as normal - 92 T/D - so as not to limit data-gathering capability of the rest of the plant when synthesis gas is not being recycled. Even under normal syngas recycle operation, a small amount of methane is reformed and the syngas fed to the gasifier. Methanation of this syngas in the gasifier provides "chemical heat" to make up for the high heat losses in the LPP relative to a commercial unit.

Char withdrawn from the bottom of the gasifier, along with fines from the tertiary cyclones, is fed to catalyst recovery. The solid-liquid separations are assumed to be made with hydroclones. The preferred system for catalyst recovery will be determined during the Process Development research phase.

The condensate from the gasifier overhead circuit and the sour slurry from the venturi scrubber are stripped of H_2S and NH_3 in a slurry stripper.

Two coal preparation trains are provided. One incorporates a steam-heated screw drier and is patterned after the system used in the pre-development research program. The second train contains an entrained drying system as envisioned for commercial CCG plants. Each train will be sized for 100% capacity; thus, each train will be a "spare" for the other.

Other offsite facilities include site preparation, coal and catalyst receipt and storage, coal preparation, utilities supply, wastewater treating, char disposal, fire protection, buildings, control house, and chemicals handling. No specific site has been chosen or offered for this grass roots pilot plant. However, for study purposes only, the characteristics of the planned Exxon Coal Liquefaction Pilot Plant (ECLP) Site at the Exxon Baytown Refinery were assumed. To make the results of this study of greater general validity, it was assumed that no opportunities exist for sharing facilities with the ECLP. The results of this study should then apply broadly at any large Gulf Coast refinery location.

Coal receipt and storage facilities include a rail spur to the plant, a below-the-track, 100 ton hopper for rail car unloading, and a 2500 ton inert gas blanketed silo. Wastewater treating facilities take no credit for facilities at the adjacent Exxon Baytown Refinery (i.e., the large pilot plant effluent will meet nominal Gulf Coast standards). Again, this approach makes the results of the study applicable more generally. The treating facilities consist of sour water stripping in the onsites area, fines removal and dewatering in a thickener and vacuum filter, and biological oxidation of the thickener overflow water. Cooling water and plant air were provided by in-plant systems, whereas steam is obtained by pipeline from the refinery. Electric power is purchased directly from the local power company.

Additional design basis information, equipment lists, and flow sketches for the grass-roots case are presented in Appendix 1.

2.6 INVESTMENT SUMMARY GRASS-ROOTS CASE

The total erected cost (TEC) for the grass-roots Catalytic Coal Gasification Large Pilot Plant is estimated to be 130 M$. This cost is for a Gulf Coast location and, as described above, assumes that there is an adjacent oil refinery to supply certain utilities and services. The investment includes the effect of cost escalation through the design and construction period. Operating costs are not included.

A breakdown of the plant investment is given in Table 2-2. Direct material, labor, and subcontract costs are 47 M$ (1Q77). Table 2-3 presents a section-by-section breakdown of the direct costs. Material costs were developed from equipment specifications and are based on cost levels for domestic purchase. Local sales tax and delivery charges to the site are included. Material charges also include the cost for shop fabrication of piping and structural steel. Labor rates are based on open shop hiring and reflect requirements of the Davis-Bacon Act. The actual job mix labor rate is $10.40/hr., which does not include payroll burdens (payroll taxes and benefits).

Total indirect project costs are 35.5 M$ (1Q77). This includes field labor overheads (17 M$) which cover temporary construction, consumables, field labor supervision, and construction equipment. Also included are payroll burdens of 2.8 M$ which cover payroll taxes and benefits. Detailed engineering, which adds 11 M$, covers design, drafting, procurement, and vendor plant inspection work. Contractors' fees, which are based on published 1Q77 rates, are 4 M$. Also included in this rate is a nominal royalty fee for the acid gas treating facilities.

The investment estimate includes 23.2 M$ to cover the escalation which is expected to occur between 1Q77 (the time basis used for estimating direct costs) and estimated project completion in 4Q82. Figure 2-2 presents the project schedule that was developed for estimating escalation. The June 1, 1980 starting date for detailed engineering is based on the assumption that a Process Development Unit (PDU) of approximately 1 T/D capacity begins operation in early 1979 to generate data for the LPP design. The LPP schedule is thus based on prudent overlap between the Process Development Program and the basic design phase for the large pilot plant project. Since the start of LPP design is keyed to PDU operation—any change to the PDU operating schedule would affect the LPP schedule. Engineering and construction times are estimated from study design specifications and estimated field labor man-hours. Overall escalation rates are 23% for materials, 26% for labor, and 36% for engineering. Details on how these rates were developed are given in Table 2-4.

Finally, the investment estimate includes a 20% project contingency to cover changes normally resulting from the firming of design and construction details. The project contingency excludes any scope or design basis changes or effects of extraordinary random events. No process development allowance for changes resulting from new laboratory data is included. However, costs for additional modifications during turnarounds are included in the pilot plant operating cost estimate.

The operating costs for the grass-roots case are estimated to be 73 M$. This is based on escalated costs for LPP operation over a two-and one-half-year period from January 1, 1983 to July 1, 1985.

A year-by-year breakdown of the operating cost components is presented in Table 2-5. Variable costs such as raw materials, transportation, and utilities are based on an overall LPP service factor of 50%. Details of the service factor basis are presented in Table 2-6. Table 2-7 presents a summary of the items included in each category and the bases that were used in preparing the estimate.

TABLE 2-1

ILLINOIS NO. 6
COAL ANALYSIS

Coal Company: Monterey Coal Company — Date of Sample: Aug. 6, 1971

Mine: Monterey No. 1 — Type Coal (Washed or ROM): Washed

Mine Location (County): Macoupin — Size: 2 x 0

Description of Sample: ______

PROXIMATE

	As Rec.	Dry
% Moisture	16.5	-
% Ash	8.0	9.58
% Volatile	35.24	42.21
% Fixed Carbon	41.79	50.05
Btu	10,700	12,814
% Sulfur	3.50	4.19
% Alkalies as Na_2O	0.15	0.18

ULTIMATE

	As Rec.	Dry
% Moisture	16.50	-
% Carbon	58.17	69.67
% Hydrogen	4.22	5.05
% Nitrogen	1.54	1.84
% Chlorine	0.18	0.22
% Sulfur	3.50	4.19
% Ash	8.00	9.58
% Oxygen	7.89	9.45

FUSION TEMPERATURE OF ASH

	Red.	Oxid.
Int. Def.	2016	2292
Softening (H=W)	2200	2445
Softening (H-1/2 W)	2227	2469
Fluid Temp.	2352	2588

ANALYSIS OF ASH

	Ign. Basis
Phos pentoxide P_2O_5	0.11
Silica SiO_2	43.82
Ferric oxide Fe_2O_3	24.69
Alumina Al_2O_3	17.19
Titanium TiO_2	0.88
Lime CaO	4.96
Magnesia MgO	1.02
Sulfur trioxide SO_3	4.29
Potassium oxide K_2O	1.61
Sodium oxide Na_2O	1.21
Undetermined	0.22

SULFUR FORMS

	As Rec.	Dry
Pyritic	1.33	1.59
Sulfate	0.18	0.22
Organic	1.99	2.38
Total	3.50	4.19

Hardgrove Grindability Index: 55.8

Size Consist: See Screen Analysis

T_{250}	2308
T_{cv}	2352
R_F	0.65
R_S	2.26
B/A	0.54

TABLE 2-2
INVESTMENT SUMMARY FOR
GRASS ROOTS LARGE PILOT PLANT

Cost Breakdown	k$, 1Q77	
Material	27,000	
Labor	12,300	
Subcontracts	7,700	
Total Direct Costs		47,000
Payroll Burdens	2,800	
Field Labor Overheads	17,300	
Vendor Representatives	300	
Loss on Surplus	300	
Insurance	200	
Engineering	10,600	
Fees: Engineering, Construction & Royalty	4,000	
Total Indirect Costs		35,500
Total Prime Contract		82,500
Project Management Services		3,800
Escalation		23,200
		109,500
Project Contingency (20%)		21,900
Total Erected Cost		131,400
CALL		130 M$

TABLE 2-3
DIRECT COST SUMMARY
GRASS ROOTS LARGE PILOT PLANT

	--------1Q77-Gulf Coast--------		
	Material k$	Labor kMH	Subcontract k$
Onsites			
• Coal Feed & Catalyst Recovery	2,780	120	10
• Gasification	3,150	95	570
• Product Gas Cleanup	1,040	35	--
• Methane Recovery	3,170	70	--
• Steam Reforming	870	40	1,060
• Preheat Furnace	240	15	850
• Acid Gas Removal	1,930	60	--
• Common Facilities	1,190	55	360
Total Onsites	14,370	490	2,850
Offsites			
• Coal Preparation	3,250	120	150
• Coal Receipt & Storage	1,400	65	1,000
• Waste Treating	1,640	110	350
• Electrical	1,150	25	--
• Interconnecting Lines	1,620	160	10
• Fire Protection	240	10	10
• Safety	330	15	--
• Site Preparation	--	10	960
• Layout	600	65	700
• Buildings	150	--	1,460
• Utilities	1,830	85	90
• Chemical Handling	130	10	70
• Catalyst Handling	290	15	50
Total Offsites	12,630	690	4,850
Total Onsites & Offsites	27,000	1,180	7,700

TABLE 2-4
BASIS FOR COST ESCALATION ESTIMATE
GRASS ROOTS LARGE PILOT PLANT

Escalation Rates	Yearly Percentage		
Base Point--1Q77	Material	Labor	Engineering
1st Year	1	8	9
2nd Year	8	8	9
3rd Year	8	7	7
4th Year	5	7	7
Centroid	April 1981	*	April 1981
Time From Base Point (yrs.)	4	3	4
Escalation Effect	23	26	36

* Note: Davis-Bacon minimum wage rate to be set at contract award 2Q80.

TABLE 2-5

OPERATING COST SUMMARY

GRASS ROOTS LARGE PILOT PLANT

	2nd Half 1980	1981	1982	1983	1984	1st Half 1985	Total
	k$						
• **Raw Materials**							
- Coal (Illinois)	-	-	-	368	617	419	
- Catalyst (K_2CO_3 Sol'n)	-	-	-	550	426	164	
Total Raw Materials	0	0	0	918	1043	583	2544
• **Transportation**							
- Coal (Illinois to Baytown unit train)	-	-	-	532	890	604	
- Catalyst (spot-shipments)	-	-	-	67	52	20	
Total Transportation	0	0	0	599	942	624	2165
• **Salaries, Wages, Benefits, and Support Services**	259	709	3441	7611	7941	4217	
Total S,W,B, and S.S.	259	709	3441	7611	7941	4217	24178
• **Administrative**							
- Taxes	0	2	293	845	2145	2145	
- Land Leasing Charges	20	40	40	40	40	20	
- Miscellaneous -	10	130	200	210	220	110	
Total Administrative	30	172	533	1095	2405	2275	6510
• **Technical**							
- Miscellaneous Services and Supplies	-	-	100	225	200	100	
Total Technical	0	0	100	225	200	100	625
• **Process Operations**							
- Catalyst and Chemicals	0	0	0	157	264	179	
- Utilities	0	0	0	3638	4851	2528	
- Process Services	0	0	10	34	35	34	
- Miscellaneous Supplies	0	0	50	20	20	20	
Total Process Operations	0	0	60	3849	5170	2761	11840
• **Mechanical**							
- Contract (Labor and Supervision)	0	0	0	4346	4177	1859	
- Material	0	0	1000	5306	5110	2261	
- Miscellaneous	0	0	450	425	375	180	
Total Mechanical	0	0	1450	10077	9662	4300	25489
GRAND TOTAL	289	881	5584	24374	27363	14860	73351

TABLE 2-6
SERVICE FACTOR BASIS

Year	Operating Days: Coal Feed	Operating Days: Catalyst Recovery
1983	128	0
1984	200	100
1985 (first half)	128	128

TABLE 2-7

OPERATING COST ESTIMATING BASIS FOR GRASS ROOTS LARGE PILOT PLANT

Raw Materials

- Coal
  - Illinois No. 6 Bituminous
  - 1980 price of $22.50/ST
  - 15% contingency on annual requirements
  - escalated at 6.6% per year

- Catalyst
  - 47 wt% K_2CO_3 solution
  - 1976 price of $152/ST
  - 15% contingency on annual requirements
  - escalated at 6.6% per year

Transportation

- Coal
  - spot-shipment by rail from St. Louis to Gulf Coast
  - 1977 cost of $26.80/ST
  - escalated at 6.6% per year

- Catalyst
  - spot-shipment by rail to Gulf Coast
  - 1976 cost of $18.50/ST
  - escalated at 6.6% per year

Salaries, Wages, Benefits, and Support Services

- Salaries, Wages, and Benefits
  - staff composed of 29 professionals and 55 technicians/operators during the 2-1/2 year operating period
  - increasing portion of staff deployed onsite during the 2-1/2 year construction period
  - salaries, wages, and benefits based on projected rates through 1982 and escalated at 5% per year through 1985

- Process and Technical Consultation
  - staff composed of 6 professionals during the 2 1/2 year operating period
  - costs based on projected engineering billing rates

- Relocation Costs
  - relocate 13 professionals to pilot plant and return to home office

Administrative

- Taxes
  - tax paid on land and plant value
  - tax rate: 0.65% during construction
1.65% after completion of construction
  - no escalation

- Land Leasing Charges
  - 9% of land market value per year
  - land market value of $450,000
  - no escalation

- Miscellaneous
  - office furniture leasing
  - office supplies
  - telephone service
  - plant security
  - travel

Technical

- Miscellaneous
  - laboratory equipment
  - laboratory technician salaries
  - supplies and services

Process Operations

- Catalysts and Chemicals
  - current costs obtained from Chemical Marketing Reporter
  - 15% contingency on annual requirements
  - escalation at 6.6% per year

- Utilities
  - electric power purchased at a 1982 unit cost of 3.03 ¢/KwH
  - natural gas purchased from an industrial gas supplier at a 1982 unit cost of 3.51 $/MBtu
  - potable water purchased at a 1982 unit cost of 0.44 $/k gal.

- steam
- industrial water
- nitrogen

purchased from Refinery

- 15% contingency on annual requirements
- escalation at 6.6% per year

- Process Services
  - char disposal
  - fire fighting service

- Miscellaneous
  - radio system installation and maintenance
  - safety equipment
  - tools
  - supplies

<u>Mechanical</u>

- Contract Labor and Supervision
  - direct labor was based on an average of 150 contract mechanical men during operating period
  - one supervisor required for every ten direct labor men
  - wage rates based on current data from maintenance contractors in the Gulf Coast
  - escalation at 6.6% per year

- Maintenance Material
  - based on 150% of direct maintenance labor costs

- Miscellaneous
  - equipment rentals
  - vehicles
  - tools
  - supplies

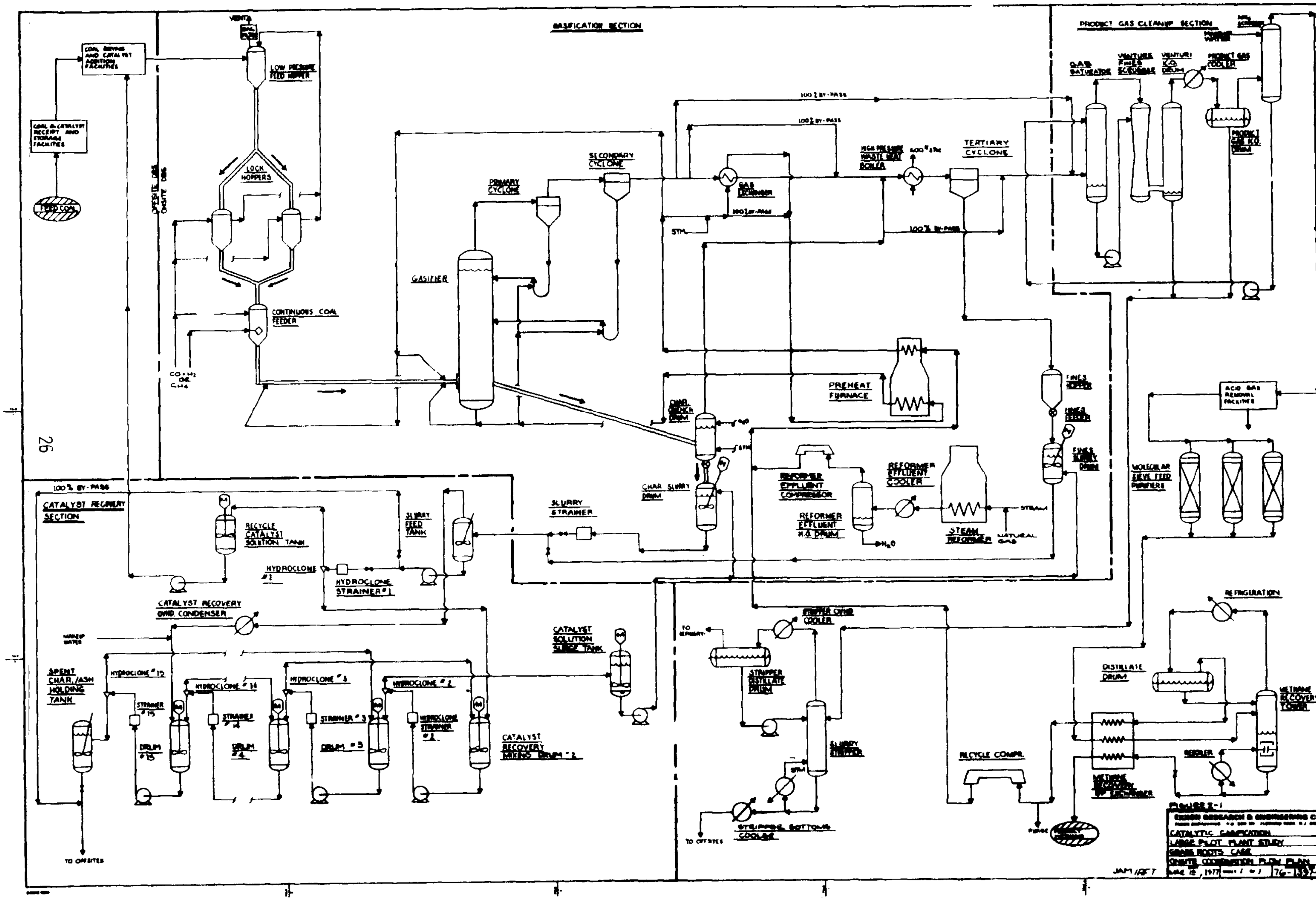
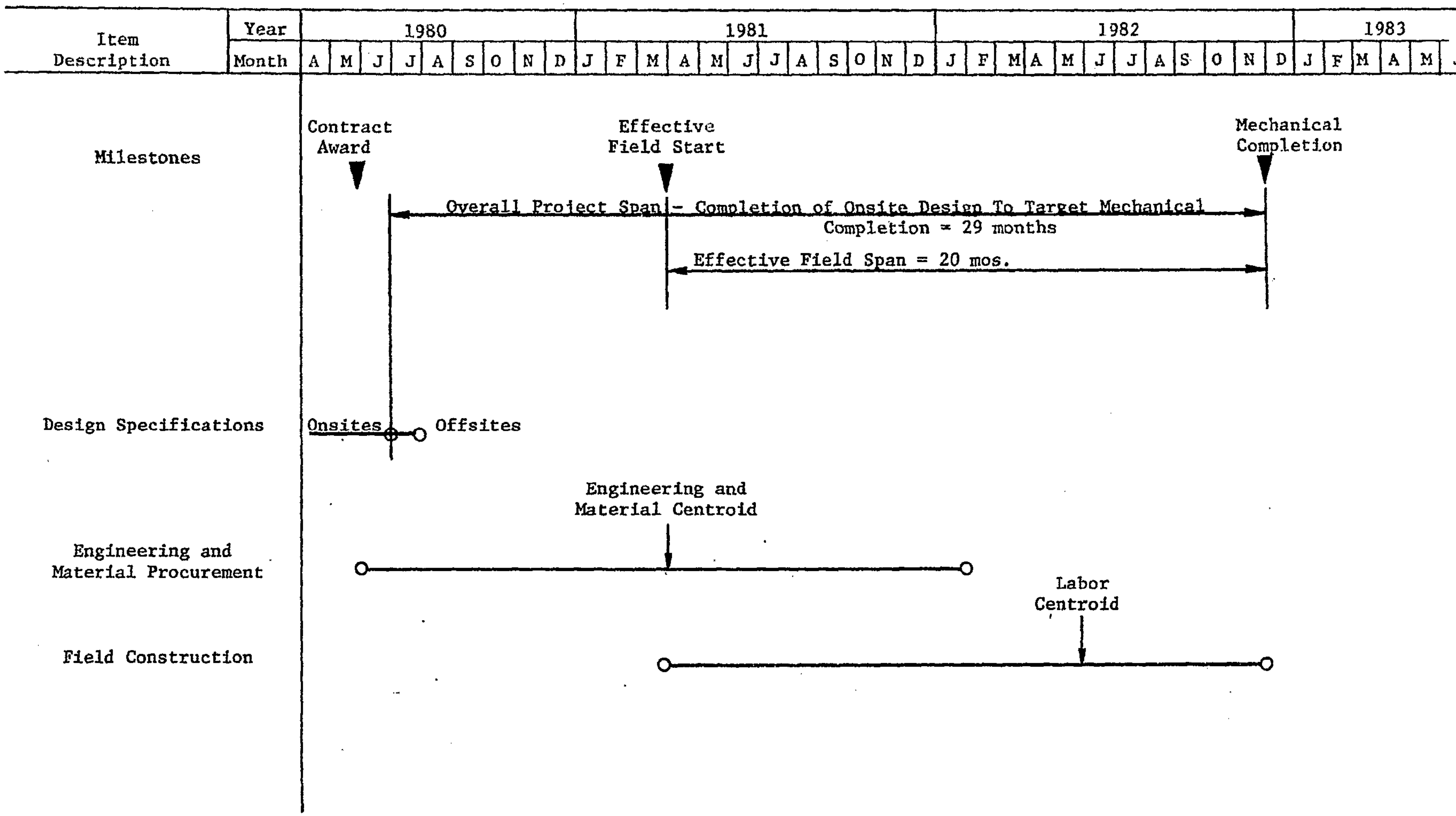

GASIFICATION SECTION
PRODUCT GAS CLEANUP SECTION
CATALYST RECOVERY SECTION
FEED COAL
COAL & CATALYST RECEIPT AND STORAGE FACILITIES
COAL DRYING AND CATALYST ADDITION FACILITIES
OFFSITE OBS
ONSITE OBS
VENTS
LOW PRESSURE FEED HOPPER
LOCK HOPPERS
CONTINUOUS COAL FEEDER
CO + H2 OR CH4
GASIFIER
PRIMARY CYCLONE
SECONDARY CYCLONE
GAS EXCHANGER
100% BY-PASS
STM
HIGH PRESSURE WASTE HEAT BOILER
600# STM
TERTIARY CYCLONE
PREHEAT FURNACE
CHAR QUENCH DRUM
H2O
CHAR SLURRY DRUM
FINES HOPPER
FINES FEEDER
FINES SLURRY DRUM
REFORMER EFFLUENT COMPRESSOR
REFORMER EFFLUENT COOLER
REFORMER EFFLUENT K.O. DRUM
STEAM REFORMER
STEAM
NATURAL GAS
GAS SATURATOR
VENTURI FINES SCRUBBER
VENTURI K.O. DRUM
PRODUCT GAS COOLER
PRODUCT GAS K.O. DRUM
ACID GAS REMOVAL FACILITIES
MOLECULAR SIEVE FEED DRYERS
SLURRY STRAINER
SLURRY FEED TANK
RECYCLE CATALYST SOLUTION TANK
HYDROCLONE #1
HYDROCLONE STRAINER #1
CATALYST RECOVERY OVHD CONDENSER
MAKEUP WATER
SPENT CHAR/ASH HOLDING TANK
HYDROCLONE #15
STRAINER #15
DRUM #15
HYDROCLONE #14
STRAINER #14
DRUM #4
HYDROCLONE #3
STRAINER #3
DRUM #3
HYDROCLONE #2
HYDROCLONE STRAINER #2
CATALYST RECOVERY MIXING DRUM #2
CATALYST SOLUTION SURGE TANK
TO OFFSITES
TO REFINERY
STRIPPER OVHD COOLER
STRIPPER DISTILLATE DRUM
SLURRY STRIPPER
STM
STRIPPING BOTTOMS COOLER
RECYCLE COMPR.
PURGE
REFRIGERATION
DISTILLATE DRUM
METHANE RECOVERY TOWER
REBOILER
FIGURE 2-1
EXXON RESEARCH & ENGINEERING CO.
CATALYTIC GASIFICATION
LARGE PILOT PLANT STUDY
GRASS ROOTS CASE
ONSITE COORDINATION FLOW PLAN
1977
76-1337-4
JAM/RFT

FIGURE 2-2
ENGINEERING AND CONSTRUCTION SCHEDULE FOR
GRASS ROOTS LARGE PILOT PLANT



SECTION 3

SELECTION OF THE PREFERRED EXISTING PILOT PLANT

3.1 INTRODUCTION

This section covers the work which dealt with the selection of the preferred existing pilot plant for conversion to a CCG pilot plant. In this screening study, the Synthane Pilot Plant, located at Bruceton, Pennsylvania, and the Hygas and Steam-Iron Pilot Plants, located in Chicago, were compared to determine which would be most suitable for conversion to catalytic gasification. These three units were selected by DOE for this analysis. Other large pilot plants were not considered because of obvious mismatches between their equipment and that required for catalytic gasification.

3.2 FACTORS CONSIDERED IN EVALUATION

Several factors were considered in comparing the potential for conversion of each of the three existing gasification pilot plants. The amount of detail required to assess the significance of these factors varied considerably. In cases where all three pilot plants had roughly comparable facilities, the adequacy of each was not investigated in detail. Where differences were large, however, the comparison was done in more detail.

The study design for the grass-roots pilot plant was in an early stage at the time the comparison of existing LPP's was made; therefore, it was necessary to complete rough heat and material balances and screening designs to determine the required equipment sizes or unit capacity of various plant sections. These were important in determining if existing equipment could be reused, and if it were reused, its impact on the large pilot plant capacity after modifications were made. Within this framework, the following elements were considered:

- The number of pilot plant sections that would be provided by existing facilities.
- The extent of facilities that would have to be dismantled and removed from the site.
- The number of plant sections that would have to be relocated to obtain a reasonable plant layout.

- Plant layout considerations, such as providing good neighbor buffer zones, and safe spacing between certain types of equipment or between sections of the pilot plant. In many instances at all three locations, Exxon Research and Engineering Company spacing standards developed for commercial petroleum refineries and chemical plants were not met. Therefore, the seriousness of deviations from these standards were considered in the final comparison.

Originally, it was planned to analyze differences in the level of scaleup data that would be obtained as a result of capacity differences between the three pilot plants. This proved unnecessary because of identical gasifier shell diameters and nearly equal nominal coal feed rates for each pilot plant.

3.3 LAND AVAILABILITY COMPARISON

The Synthane Pilot Plant is located on a 15-acre tract of which only 11 acres are usable because of the hilly terrain. At IGT in Chicago, the total contiguous site for the Hygas, Steam-Iron, and Agglomerating Ash Burner Pilot Plants is only 8 acres. Preliminary estimates were that the land requirement for the grass-roots catalytic gasification pilot plant would be approximately 25 acres. This suggests that Synthane would be marginally acceptable as a large catalytic gasification pilot plant site, whereas the Hygas and Steam-Iron Pilot Plants would not be acceptable unless additional land could be made available. The Hygas Plant is bounded on its four sides by a railroad siding, other IGT pilot plants, a public road, and a large commercial gas distribution compressor station. Steam-Iron has similar limitations in that it is bounded by a railroad siding, the Hygas Pilot Plant, a public road, and open land belonging to a power company. Synthane is located on a triangular tract which is bounded by a public road, another DOE pilot plant, and privately-owned stables. Thus, only in the Synthane case is there a reasonable possibility of securing additional land.

Despite the land limitations, an attempt was made to develop an approximate layout of a catalytic gasification pilot plant for each location. These layouts were made with the knowledge that sacrifices in ER&E minimum spacing standards would have to be made. Since these standards have been developed for refineries and chemical plants, not all are necessarily applicable to a large pilot plant. The results of these plant layout studies are shown in Figures 3-1, 3-2, and 3-3 for Synthane, Hygas, and Steam-Iron, respectively. A comparison of spacing between various components at the three pilot plants is presented in Table 3-1.

At Synthane, there are compromises in the minimum spacing standards, but these compromises appear reasonable. However, if major additions were required to non-process facilities, such as steam generation and electric power, more severe problems might be encountered. The most significant problem would be the location of the steam reforming and preheat furnaces. These would be located for safety in an area that is currently an embankment rising from the main plant elevation. Approximately 5,000 cubic yards of soil and rock would have to be removed to level this area. Only nominal equipment relocation would be required. The most significant item is the removal of existing CO_2 compressors to make room for the cryogenic system. A reasonable alternative to CO_2 compression, which involves installation of small liquid CO_2 pump and vaporization system, has been developed.

With only about 2.5 acres to work with at Hygas, a reasonable layout could not be developed. As shown in Figure 3-3, it would be necessary to relocate major plant sections, including the sulfur plant, the filter building, and the incinerator and flare. Also, it would be necessary to dismantle a large amount of existing equipment. This equipment, which is listed in Table 3-3, is primarily from the feed and coal pretreatment sections. Serious deviations from recommended standards would still remain. The most serious of these is the close proximity of the hydrogen reformer furnace to the compressor house. There is also no buffer zone between the process areas and neighboring land.

A proposed layout for the Steam-Iron Pilot Plant is shown in Figure 3-3. This layout is on a plot area of approximately 4 acres, of which only 3 are presently allocated to the Steam-Iron Pilot Plant. This layout is undesirable because the process furnaces are located centrally among the processing facilities. It would be preferred from the standpoint of safety to locate them at one edge of the process block. However, this would not be possible because the preheat furnace would then be so far from the gasifier that unacceptable heat losses would be incurred. Other drawbacks are the closeness of the process area to property lines and lack of any remaining space for expansion or addition of non-process facilities.

In order to make room for new process equipment at the Steam-Iron Pilot Plant, it would be necessary to move a 35 x 100 foot concrete block warehouse/maintenance building. This building is located in the areas designated on the plot plan for catalyst storage and catalyst recovery. A replacement building would most likely have to be built in the open area next to the control house and maintenance building.

3.4 REUSABLE EQUIPMENT COMPARISON

All of the facilities at the three locations were compared on a section-by-section basis. These comparisons have been made in two categories listed below and are discussed in the following sections:

- Existing facilities that can be reused in the same service.
  - Coal receipt, storage, and preparation - Table 3-2.
  - Gasifier char handling and gas processing - Table 3-3.
  - Utilities systems - Table 3-4.
- Facilities which must be dismantled - Table 3-5.

3.4.1 Coal Handling Facilities

There are similarities in processing capability at all three sites, but process configurations and equipment differ. The major difference is in the coal feed system. The Exxon catalytic gasification process requires dry coal feed with a lock hopper system to minimize the heat load on the gasifier. Synthane has in place a lock hopper feed system capable of feeding up to 120 tons/day of solids and can be used with substantially no modifications. At Hygas and Steam-Iron, there would not only be added cost for installing a new coal feed system but costs for removal of the existing facilities.

3.4.2 Gasifier

The gasifier vessel at each pilot plant has a 5.0 ft. ID shell with varying thicknesses of internal ceramic lining. The Synthane gasifier differs from the other two in that it is an internally-lined cold wall design, whereas the other gasifiers employ jackets. The advantage of the cold wall design is that it is potentially cheaper and that nozzle connections can be made easily.

3.4.3 Char Withdrawal

Both the Synthane and Hygas units have a combination of char quench and filtration. The Steam-Iron unit has only a char quench system. A factor favoring Synthane is that their system also includes a steam, fluidized-bed char cooler which is similar to the system proposed for the grass-roots pilot plant.

3.4.4 Raw Gas Quench

None of the three pilot plants have quench facilities that parallel those planned for the grass-roots design. However, Synthane has more equipment that appears suitable for the new service requirements than either Hygas or Steam-Iron.

3.4.5 Acid Gas Removal

Synthane and Hygas have roughly comparable equipment (towers, heat exchangers, and drums), whereas Steam-Iron has none.

Synthane has an advantage over Hygas in that it is designed for hot carbonate acid gas removal, and its regenerator is capable of the new service requirements. On the other hand, Hygas would most likely require conversion from diglycolamine to hot carbonate to meet the new duty requirements. This would entail extending the regenerator tower from 70 to 110 feet and adding new pumps and solution storage facilities. Both Synthane and Hygas would require new absorption towers.

3.4.6 Sulfur Recovery

Both Hygas and Synthane have sulfur plants; Steam-Iron does not. Synthane's sulfur plant is the Stretford type and has a capacity of 3.1 tons/day, which is adequate for the new acid gas H_2S removal requirements. Hygas has a Claus-type plant with a capacity of 1.8 tons/day. This capacity would not be adequate if it is determined that H_2S must be recovered from sour water streams in the pilot plant.

3.4.7 Steam Reforming

Only Hygas has an existing steam reformer. Its capacity of 2 MSCF/D hydrogen may not be sufficient, since 3 MSCF/D has been found necessary for a 3.5-foot diameter cold wall gasifier. The jacketed wall design of the Hygas gasifier will reduce this requirement. Credits for an existing reformer were not considered great enough to warrant a definitive determination of the minimum reformer capacity required at Hygas.

3.4.8 Non-Process Facilities

In a majority of cases, the three pilot plants have roughly equivalent facilities. However, Synthane ranked behind Hygas and Steam-Iron in electrical load carrying capacity and cooling water distribution. Of greater significance is the fact that Synthane does not have pipeline natural gas available at the site. One solution to this problem is to provide liquified natural gas (LNG) receiving and storage facilities to supply methane feed to the reformer and fire No. 2 fuel oil in the steam reformer and synthesis gas preheat furnaces. The long-term operating cost debit for using liquefied natural gas and No. 2 fuel instead of pipeline natural gas is difficult to determine, since these fuels are required only during operations with once-through synthesis gas. During recycle operations, the reformer feed requirement is reduced, and the methane that is produced in the process (SNG) can be used for reformer feed and furnace fuel. Relative operating times for the two modes of operation will be a function of how well gas cleanup and separation facilities operate.

3.5 CONCLUSIONS

After comparing the three existing pilot plants, it was concluded that the Synthane unit would be the preferred pilot plant for conversion to catalytic gasification. The Synthane Site is the only one that can reasonably be expected to accommodate the new facilities and meet minimum spacing standards for safe operations and equipment maintenance. In addition, with Synthane, more of the existing equipment could be utilized, and less equipment would have to be removed and/or relocated during conversion to a catalytic gasification large pilot plant.

TABLE 3-1

COMPARISON OF FACILITIES SPACINGS
PROPOSED MODIFICATIONS TO EXISTING PILOT PLANTS

Locations	Synthane	Hygas	Steam-Iron
Process Areas to			
Property Lines	150	10	50
Boiler House	60	40	75
Buildings	100	80	40
Char Pond	150	10	200
Control House	65	35	50
Furnaces to			
Control House	125	100	80
Compressors	75	15	100
Main Structure	60	100	60

TABLE 3-2

EXISTING FACILITIES COMPARISON

COAL RECEIPT, STORAGE AND PREPARATION

	Synthane	Hygas	Steam-Iron
• Coal Receipt and Raw Storage	Raw coal is delivered via railcar (approx. 3000 ton shipments) to a local coal yard and stored. It is delivered to the site by 20-ton trucks at Synthane's request. The truck discharges into a bin and from there via a bucket elevator into a 240 ton inerted storage bin.	Raw coal is delivered via railcar (approx. 3000 ton shipment) to the site. Plant uses a rail spur owned by People's Gas Co. Site facilities include: car puller, car shaker, receiving hopper, and staking conveyor. Coal is stored in 1500 ton piles on either side of a concrete wall.	Char is delivered to the site with the HYGAS facilities. A separate conveyor transfers it to their storage pile.
• Coal Pulverizing and Drying System	Raymond gas swept mill system. The mill is a Raymond #53 Impact Mill, the heater is a Raymond #7. The system is designed to produce 10,000 #/hr of dried ground coal. Feed: minus 3/4", 18% total moisture. Product: minus 20 mesh, 4% total moisture. The product is pneumatically conveyed to storage.	William's gas swept mill system. The mill is a William's "Standard" rolling mill. The system is designed to produce 5.6 tons/hr of dried ground coal. Feed: minus 1-1/4", 35% total moisture. Product: 10 x 80 U.S. Standard mesh, 5% total moisture. The product is pneumatically conveyed to storage.	Char is reclaimed by front-end loader, dump through a 3/8" grizzly onto the dryer feed belt. The dryer is direct fired fluid bed type mfg. by the Jeffrey Mfg. Feed: max. 12 tons/hr., minus 3/8", 25% total moisture. Product: 5% total moisture. It is then conveyed to a screen via a bucker elevator, the plus 10 mesh is then fed into a Norberg 36" Gyradisc crusher. The output of the crusher is then cycled back to the screen. The product is minus 10 mesh, plus 100 mesh. From the screens to storage via a bucket elevator.
• Prepared Coal Storage	100 ton inerted storage bin, located above lock hopper system.	60 ton inerted storage bin.	60 ton inerted storage bin.
• Coal Feed System	Lock Hopper (Petrocarb System) - normal rate 3 tons/hr. (5 tons/hr maximum)	Hydrocarbon slurry - normal coal rate is 3 tons/hr.	Water slurry - normal char rate is 1.8 tons/hr.

TABLE 3-3

EXISTING FACILITIES COMPARISON

GASIFIER, CHAR HANDLING AND GAS PROCESSING

	Synthane	Hygas	Steam Iron
• Gasifier	• 5'-0 ID 62'-0 upper section and 5'-0 x 15'-4 and 2'-6 x 17'-8 lower section. Upper section is cold wall design (Monel clad steel plus insulating and refractory internal lining). Lower section is unlined and partially clad (stainless steel). Vessel is supported by structure. Supports located on upper section 8' above flange connecting upper and lower section.	• 5'-6 ID x 62'-2 upper section and 5'-0 ID x 62'-2 lower section. Upper section consists of vessels and piping enclosed in the outer shell which is carbon steel. Lower section is double wall design. Inner wall is C-1/2 Mo steel without cladding but with insulating and refractory liner. Outer wall is carbon steel. The space between the walls is maintained full of water at reactor inside pressure during normal operation. Vessel is skirt supported. Water reservoir for reactor jacket is supported in structure.	• 5'-0 ID x 42'-1 upper section 5'-0 ID x 75'-0 lower section. Both sections are double wall design. Inner wall is C-1/2 Mo steel with insulating and refractory liner. Outer wall is carbon steel. The space between the walls is maintained full of water at reactor inside pressure during normal operation. Vessel is skirt supported and water reservoir for reactor jacket is supported on the reactor.
• Cyclones	• 1 stage-internal, Ducone size 6. 8-1/2" x 3" dia. x 2' - 9" long.	• 1 Stage-external. Solids not returned to gasifier.	• 1 stage-internal. Ducone size 11 m. 12-3/4" x 3" dia. x 4' long.
• Char Withdrawal	• 2'-6 ID x 15'-6 fluidized bed cooler (lower section of gasifier).		
	• Two-3'-6 ID x 19'-0 lock hoppers.		
	• 7'-0 ID x 17'-0 slurry quench tank (5 psig).		
	• Two-160 gpm x 60 psi ΔP x 10 Bhp slurry circulating pumps.		
	• Two-80 gpm x 40 psi ΔP x 5 Bhp filtrate return pumps.		
	• 2'-0 ID x 10'-6 slurry quench tank. (high pressure alternative)	• 3'-0 x 2'-7 and 1'-6 x 14'-0 slurry quench tank.	• 3'-0 ID x 7'-9 and 2'-0 x 15'-6 slurry quench tank.
	• 75 gpm x 1100 psi ΔP x 200 Bhp filtrate return pump.	• Two-95 gpm x 50 psi ΔP x 15 Bhp slurry pumps.	• 90 gpm x 15 psi ΔP x 7.5 Bph quench tank circulation pump.
	• 6'-6 x 10'-0 filter feed tank.	• Twelve-22 sq ft (bare) finned tube, double-pipe char slurry cooler.	• 6'-0 x 4'-0 TT flash tank (50 psig).
	• 100 gpm x 15 psi ΔP x 10 Bhp filter feed pump.	• Flash tank (5 psig).	
	• Two disc-type vacuum filters with 41,000 lb/hr (3050 lb/hr char) max. feed rate, ea.	• Aier vacuum filter.	
	• Two 630 acfm x 22" Hg x 50 Bhp vacuum pumps.		
	• Settling pond	• Settling basin	• Settling basin
		• Eden's separator	

TABLE 3-3 (Cont'd.)

	Synthane	Hygas	Steam-Iron
• Raw Gas Quench	• Ejector Venturi	• Prequench Tower	• Ejector Venturi
	• 5'-0 ID x 14'-0 surge drum	• 2'-0 ID x 35'-0 quench tower	• 4'-0 ID x 32'-0 quench tank
	• Two-188 gpm x 165 psi ΔP x 50 Bhp venturi recycle pumps	• 4'-0 x 18'-8 quench water separator	• Two-120 gpm x 1625 psi ΔP x 150 Bhp process water pumps
	• Two-780 sq ft venturi recycle coolers (55 psi steam generation)	• Two-quench water circulation pumps	
	• 4'-6 x 15'-0 steam drum	• Light oil system - See list of removable equipment (Table)	
	• 2'-6 ID x 39'-0 raw gas scrubber (8 trays + 20' packing)		
	• 185 gpm x 55 psi ΔP x 25 Bhp scrubber water recycle pump		
	• Two-781 sq ft scrubber water recycle cooler (55 psi steam generation)		
	• Waste water cooler		
	• 9'-0 x 10'-0 waste water receiver (5 psig)		
	• Wash oil system and tar handling system - See list of removable equipment (Table)		
• Acid Gas Removal	• Benfield hot carbonate	• Diglycolamine	• None
	• 2'-6 x 57'-0 absorber	• 2'-6 x 70'-0 absorber	
	• Two-250 Bhp semi-lean solution pumps	• 280 gpm x 1600 psi ΔP x 500 Bhp lean solution pump	
	• Two-200 Bhp lean solution pumps	• 300 gpm x 85 psi ΔP x 25 Bhp lean solution filter pump	
	• 4'-0 x 112'-0 regenerator	• 4'-0 x 72'-0 regenerator	
	• 3.1 x 10^6 Btu/hr condenser	• 913 sq ft, 6.0 x 10^6 Btu/hr condenser (gravity reflux)	
	• 2'-6 x 6'-0 reflux drum		
	• Two-10 Bhp reflux pumps		
	• 3.2 x 10^6 Btu/hr reboiler plus live steam to regenerator	• 1532 sq ft, 15.4 x 10^6 Btu/hr reboiler	
		• 886 sq ft, 5.5 x 10^6 Btu/hr lean/rich solution heat exchanger	
	• 3.4 x 10^6 Btu/hr lean solution cooler	• 1280 sq ft, 9.7 x 10^6 Btu/hr lean solution cooler	
		• 1'-6 x 32'-0 caustic + water wash tower	
		• 12 gpm x 50 psig x 10 Bhp caustic circulation pump	
	• 9'-0 x 10'-6 (50000 gal) solution storage tank		
	• 3 Bhp solution makeup pump		

TABLE 3-3 (Cont'd.)

	Synthane	Hygas	Steam-Iron
• Sulfur Recovery	• 3.1 ton/day liquid sulfur • Stretford Unit	• 1.8 ton/day liquid sulfur • Claus Plant	• None
• Reforming	• None	• 2.0×10^6 SCF/D H_2	• None

TABLE 3-4

EXISTING FACILITIES COMPARISONS

UTILITIES AND DISPOSAL SYSTEMS

	Synthane	Hygas	Steam-Iron
• Steam Systems			
+ High Press. (#/hr)	25,000 (1 boiler)	20,000 (1 boiler)	20,000 (1 boiler)
- Press. (psig)	1,150	1,550	1,100
- Temp. (°F)	800	Saturated (a separate superheater will superheat 10,000 #/hr to 1250°F).	573 (a separate superheater will superheat 7,000 #/hr to 1050°F)
+ Low Press. (#/hr)	7,000 (2-100% boilers)	45,000 (1 boiler)	25,000 (1 boiler)
- Press. (psig)	55	150	150
- Temp. (°F)	Saturated	Saturated	Saturated
+ Fuel			
- High Press.	No. 2 Fuel Oil or Synthetic Gas	Natural Gas	Natural Gas
- Low Press.	No. 2 Fuel Oil	Natural Gas	Natural Gas
• Compressed Air + Discharge Press.	2-647 cfm compressors - 105 psig (1 plant, 1 instrument)	2-500 SCFM compressors 100 psig (1 plant, 1 instrument)	2-500 SCFM compressors 100 psig (1 plant, 1 instrument)
• Electric Power	2,700 kW 2,400 volts 1,200 ampers	4,500 kW 4,160 volts 1,200 amperes	3,375 kW 4,160 volts 1,000 amperes 125 kW Emergency Gen. 1,350 kW 480 volts 3,000 amperes
• Cooling Water	2,400 gpm	50% of the capacity of Steam-Iron tower	8,500 gpm
+ Type	tower-fresh water		tower-fresh water
+ Temp. In (°F)	120		115
+ Temp. Out (°F)	90		85
+ Pressure (psig)	45		
• Water System			
+ Fire	City water main, 160 psig.	Canal water, 2-500 gpm pumps 190 psig.	Canal water, system manufactured by Fairbanks Morse Pump Div. of Colt Industries 2-75 HP pumps (1 electric, 1 diesel).
+ Process Water	City	Canal and City Water	City
+ Demin. Water	Cation Unit, Anion Unit, Mixed Bed Unit. Supplies 70 gpm to Deaerator, caustic system, H.P.C. System.	Cation Unit, Decarbonator, Anion Unit. Feeds Deaerator and Caustic System.	Dual Mixed Bed. Feed only Deaerator.
• Fuel Type	Fuel Oil, Propane	Natural Gas	Natural Gas
+ Capacity	Fuel Oil: 1-25,000 gal. tank, 1-100,000 gal. tank. System supplies 8.2 gpm @ 160 psig. Propane: 1-30,000 gal. tank.	2,100 SCFM	

TABLE 3-4

	Synthane	Hygas	Steam Iron
• Gas System			
+ Liquid Storage	O_2 and CO_2 (facilities rented)	O_2 and N_2 (facilities rented)	N_2 (facilities rented
+ Gas	CO_2-2 compressors 720 SCFM, 1,277 psia discharge. Inert gas generation-20,000 SCFH, blower to 7.3 psig. Compressor-55 cfm inlet, 1,200 psig discharge.	H_2 plant - 2.0×10^6 SCFD, 1,580 psia.	Natural Gas compression 10,000 SCFH, 1,300 psig. N_2-1,200 psig from HYGAS 1,825 SCFM.
• Waste Disposal			
+ Incinerator	80' high stack	80' high stack	100' high stack-heat duty 49×10^6 Btu/hr (does not burn liquid).
+ Flare	80' high stack	90' high stack	100' High Stack
+ Solids	Filter, emergency settling basin	Settling pond and filters	Settling basin
• Sulfur Handling	Liquid sulfur storage 1,300 gal.	Liquid sulfur transferred off site	No sulfur plant

<u>TABLE 3-5</u>

<u>FACILITIES TO BE DISMANTLED</u>

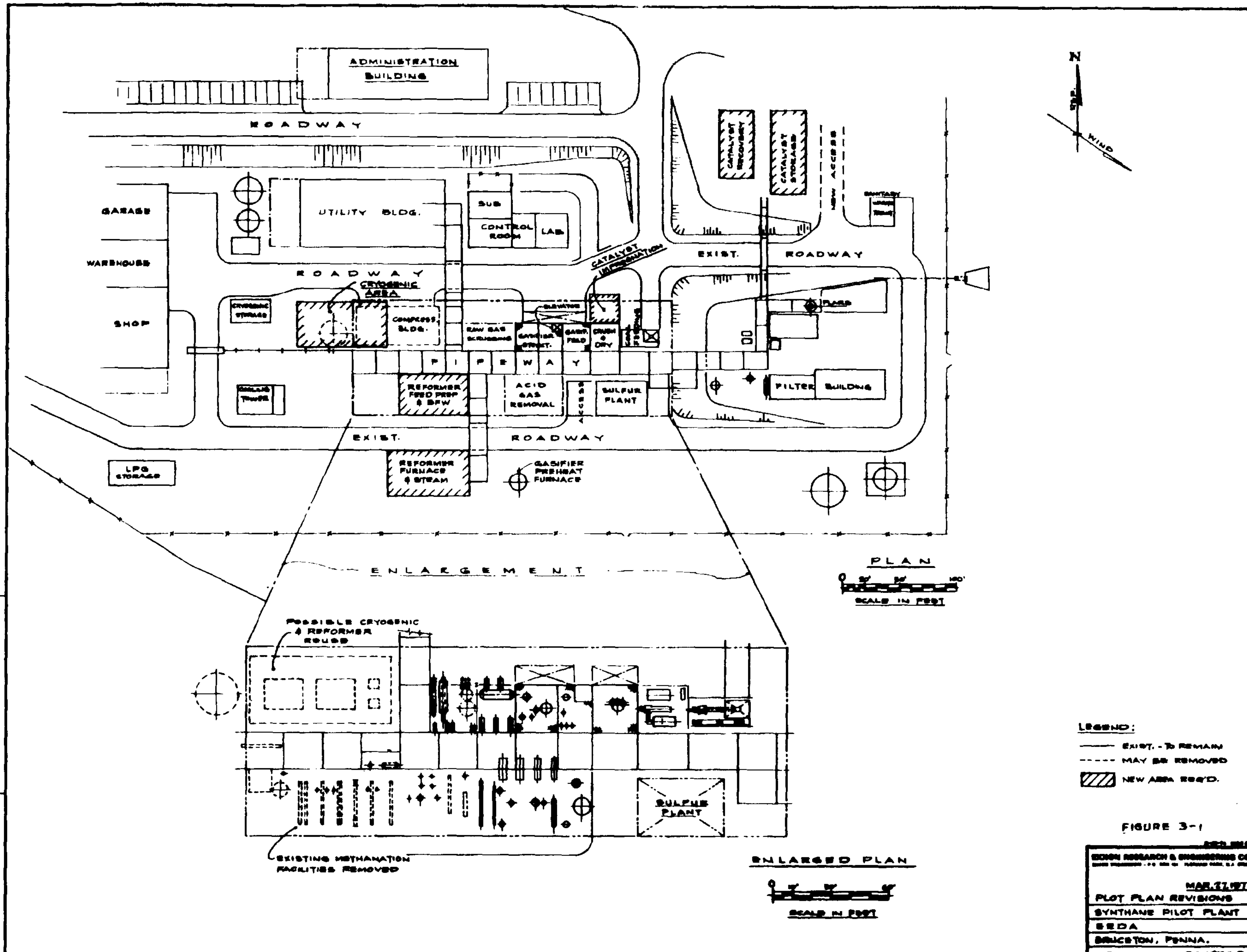
	Synthane	Hygas	Steam-Iron
• Coal Pretreater	• 8 inches ID x 48'-0 reactor (1000 psig) • 1'-8 ID x 13'-0 reactor	• 1590 ACFM x 15.5 psi ΔP x 150 Bhp compressor • 1.75 x 10^6 Btu/hr direct fired air heater • 8'-2 ID x 17'-0 reactor (10 psig) • 3'-0 ID x 16'-0 treated coal cooler • 5 ton/hr x 74' high bucked elevator • 5 ton coal hopper • 18 ton coal hopper • Miscellaneous feeders and conveyors • Dust collection system • 3'-6 ID x 15'-0 treater off-gas quench tower • Ejector Venturi • 40 gpm x 50 psi ΔP x 5 Bhp quench water circulation pump • 3'-6 ID x 10'-0 quench water separator • 0.8 x 10^6 Btu/hr quench water cooler	• None
• Coal Feeding	• (Existing solid coal feed system is reusable)	• Two-8'-0 ID x 8'-0 agitated slurry drums (0 psig) • Two-160 gpm x 73 psi ΔP x 25 Bhp slurry circulating pumps • Two-40 gpm x 1540 psi ΔP x 75 Bhp charge pumps • 1.8 x 10^6 Btu/hr double pipe slurry heater (HP steam)	• 5800 gal agitated slurry tank (0 psig) • Two-60 gpm x 30 psi ΔP slurry circulating pumps • Two-25 gpm x 1870 psi ΔP x 40 Bhp charge pumps • 4'-0 ID x 6'-6 slurry blowdown drum • 10.2 x 10^6 Btu/hr (absorbed) slurry preheat furnace

TABLE 3-5 (Cont'd.)

	Synthane	Hygas	Steam-Iron
• Oil and Tar Handling	• 6'0- ID x 6'-0 decanter (0 psig) • Two vessels 250 gal each • Five pumps 5 Bhp, One-7.5 Bhp pump	• 4'-0 ID x 20'-0 quench separator • 0.6 x 10^6 Btu/hr recycle oil cooler • Oil storage tank • Light oil steam stripper	• None
• Methanation	• Several small equipment items - 0.2 x 10^6 SCF/D methane produced	• Several small equipment items - 0.3 x 10^6 SCF/D methane produced	• None
• Other			• Iron ore feed system • 5'-6 OD x 182' high steam-iron reactor • 2'-0 ID x 28'-9 product quench tank with spray contactor • 4'-0 ID x 42'-9 preheater quench tank with spray contactor • Process air compressors - 4300 acfm, 0 to 300 psig • Ingersol Rand centrifugal model C50MA and stage reciprocating model 13, 9 and 6-1/4 - 12 - 3HHE - 3 - NL - 2. • 3.5 x 10^6 Btu/hr (absorbed) process air preheat furnace

CAUTIONARY NOTICE REGARDING USE OF PLOT PLAN REVISION DRAWINGS

These plot plans were developed to provide a basis for site evaluation only. They are to be considered screening quality and are subject to change depending on more definitive evaluations of the adequacy of existing equipment and the requirements for new plant sections.


ADMINISTRATION BUILDING
ROADWAY
GARAGE
WAREHOUSE
SHOP
UTILITY BLDG.
SUB
CONTROL ROOM
LAB.
CATALYST IMPREGNATION
CATALYST RECOVERY
CATALYST STORAGE
NEW ACCESS
EXIST. ROADWAY
ROADWAY
CRYOGENIC AREA
CRYOGENIC STORAGE
COMPRESS. BLDG.
RAW GAS SCRUBBING
ELEVATOR
CRUSH & DRY
FLARE
FILTER BUILDING
PIPEWAY
REFORMER FEED PREP & BFW
ACID GAS REMOVAL
ACCESS
SULFUR PLANT
EXIST. ROADWAY
REFORMER FURNACE & STEAM
GASIFIER PREHEAT FURNACE
LPG STORAGE
N
WIND
ENLARGEMENT
PLAN
SCALE IN FEET
POSSIBLE CRYOGENIC & REFORMER REUSE
SULFUR PLANT
EXISTING METHANATION FACILITIES REMOVED
ENLARGED PLAN
SCALE IN FEET
LEGEND:
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MAY BE REMOVED
NEW AREA REQ'D.
EXXON RESEARCH & ENGINEERING CO.
MAR. 27, 1977
PLOT PLAN REVISIONS
SYNTHANE PILOT PLANT
ERDA
BRUCETON, PENNA.
PP-1577-D

FIGURE 3-1

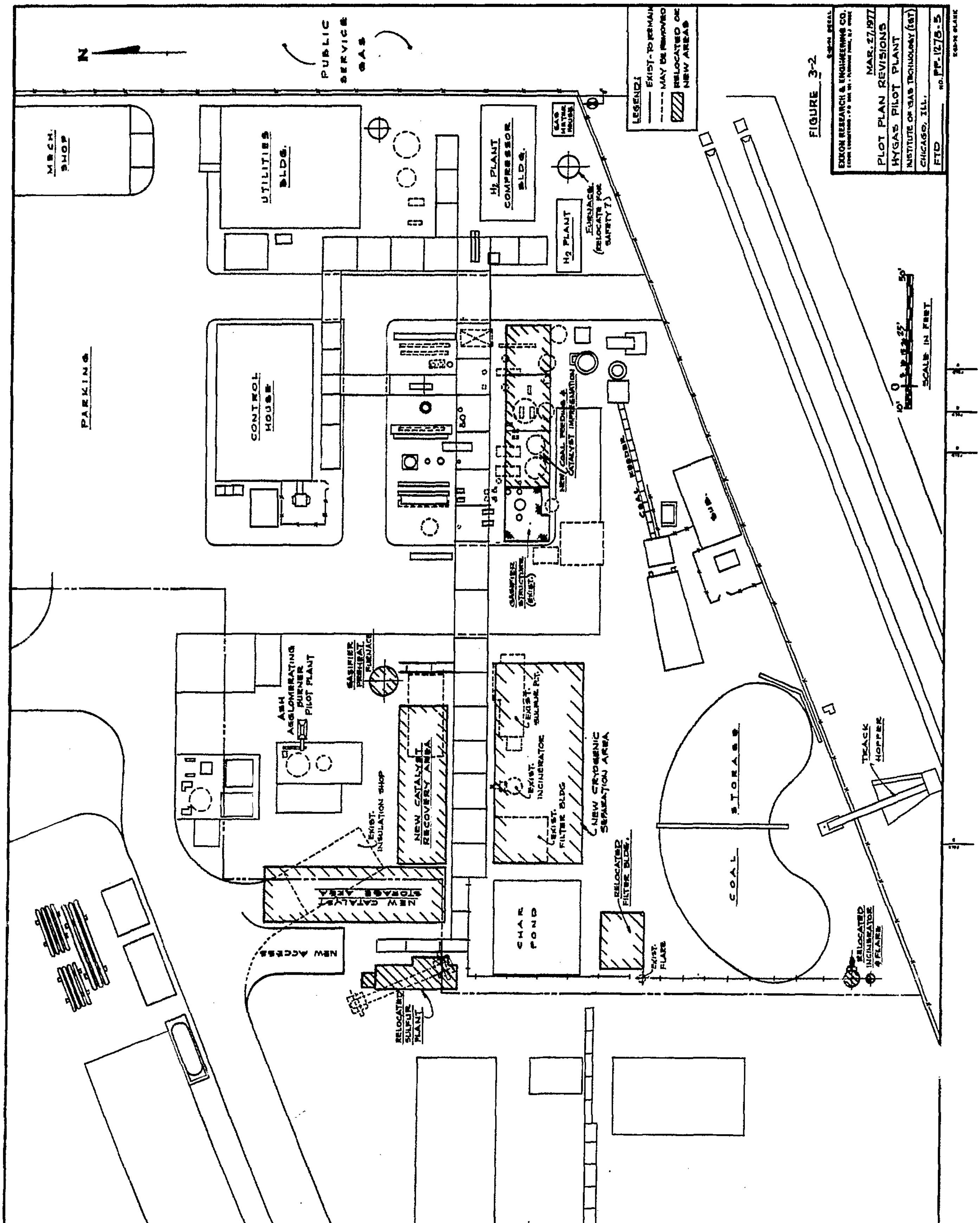
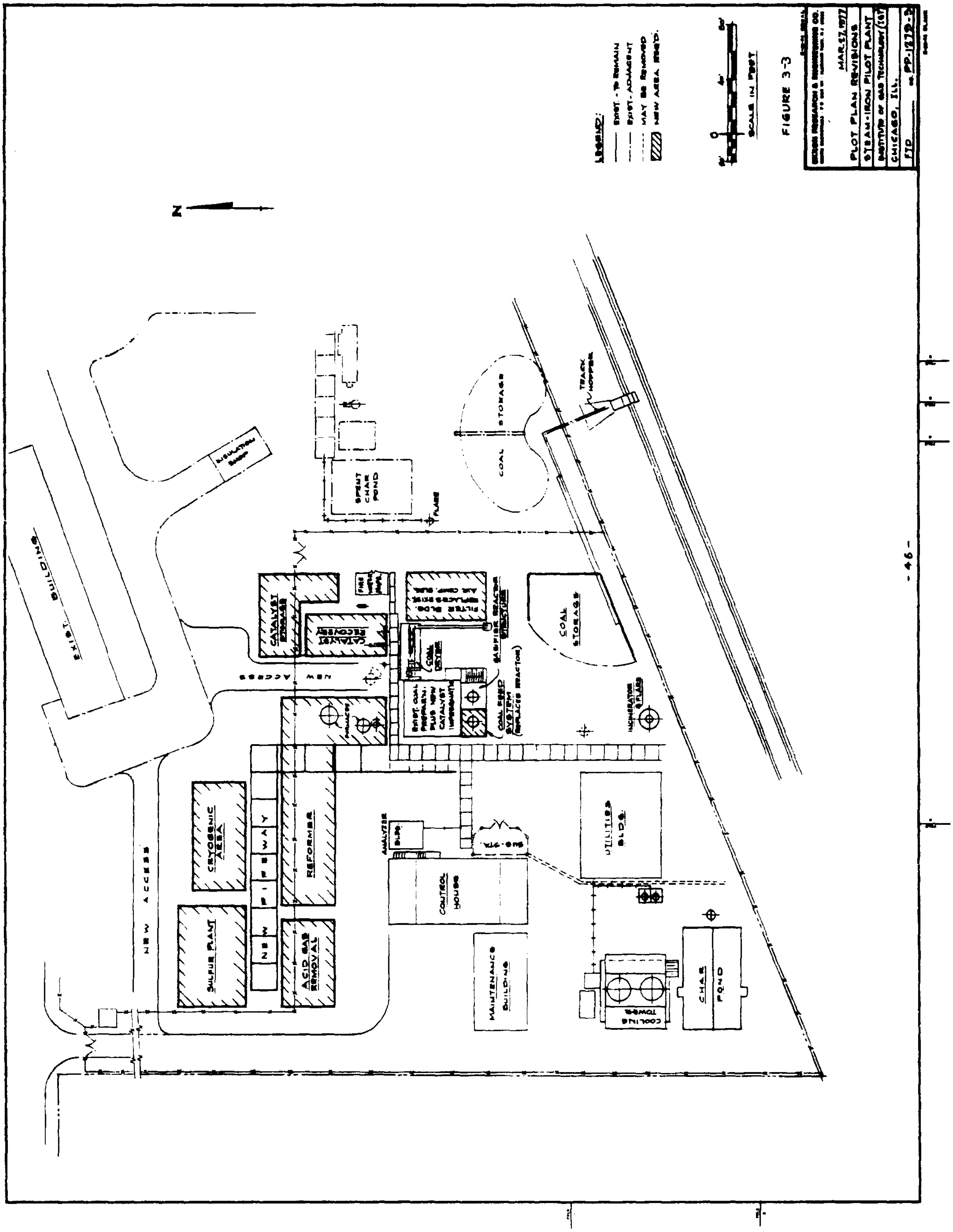

N
PUBLIC SERVICE GAS
MACH. SHOP
UTILITIES BLDG.
H2 PLANT COMPRESSOR BLDG.
GAS METER HOUSE
H2 PLANT
FURNACE (RELOCATE FOR SAFETY ?)
LEGEND:
EXIST.-TO REMAIN
MAY BE REMOVED
RELOCATED OR NEW AREAS
EXXON RESEARCH & ENGINEERING CO.
MAR. 27, 1977
PLOT PLAN REVISIONS
HYGAS PILOT PLANT
INSTITUTE OF GAS TECHNOLOGY (IGT)
CHICAGO, ILL.
FTD
NO. PP-1275.5
PARKING
CONTROL HOUSE
NEW COAL FEEDING & CATALYST IMPREGNATION
COAL FEEDER
SUB.
GASIFIER STRUCTURE (EXIST.)
ASH AGGLOMERATING BURNER PILOT PLANT
GASIFIER PREHEAT FURNACE
EXIST. SULFUR PIT
EXIST. INCINERATOR
EXIST. FILTER BLDG
NEW CRYOGENIC SEPARATION AREA
NEW CATALYST RECOVERY AREA
EXIST. INSULATION SHOP
NEW CATALYST STORAGE AREA
NEW ACCESS
CHAR POND
RELOCATED FILTER BLDG.
EXIST. FLARE
COAL STORAGE
TRACK HOPPER
RELOCATED INCINERATOR & FLARE
RELOCATED SULFUR PLANT
SCALE IN FEET
10' 0 50'

FIGURE 3-2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EXIST. - TO REMAIN
EXIST. ADJACENT
MAY BE REMOVED
NEW AREA REQ'D.
SCALE IN FEET
PLOT PLAN REVISIONS
MAR. 27, 1977
STEAM-IRON PILOT PLANT
CHICAGO, ILL.
PP-1272-2
EXIST. BUILDING
NEW ACCESS
CATALYST STORAGE
CATALYST RECOVERY
CRYOGENIC AREA
REFORMER
NEW PIPEWAY
SULFUR PLANT
ACID GAS REMOVAL
CONTROL HOUSE
MAINTENANCE BUILDING
UTILITIES BLDG.
COOLING TOWER
CHAR POND
COAL STORAGE
SPENT CHAR POND
TRACK HOPPER
FILTER BLDG.
GASIFIER REACTOR STRUCTURE
COAL FEED SYSTEM
INCINERATOR & FLARE
FLARE
N

FIGURE 3-3

- 46 -

SECTION 4

STUDY DESIGN AND COST ESTIMATE FOR MAJOR SYNTHANE REVAMP

4.1 PROJECT BASIS

The location for the Major Synthane Revamp is at the existing Synthane Site which is in Bruceton, Pennsylvania. Steam is generated at the site, electric power is purchased, and cooling water is provided by a recirculating system. The design feed coal is Illinois No. 6. The coal composition is the same as for the Grass-Roots case, as shown in Section 2.

4.2 PROCESS BASIS

The process basis for the Major Synthane Revamp is the same as that for the Grass-Roots case, as described in Section 2.

4.3 DESIGN PHILOSOPHY

The design philosophy for the Major Synthane Revamp is the same as that employed for the Grass-Roots case, as described in Section 2. The pilot plant is designed to obtain all the data necessary to design a pioneer commercial plant with reasonable risk.

4.4 LPP SIZE

The reactor diameter was set at 3.5 feet, with a bed height of 100 feet, as in the Grass-Roots case. The design feed rate is 92 T/D (Illinois No. 6 coal, as-received). Both catalyst and synthesis gas recycle loops are closed to permit operation in integrated fashion, as in the Grass-Roots case.

4.5 SUMMARY OF MODIFICATIONS TO SYNTHANE PLANT

The scope of the revamp is illustrated in Figure 4-1, which is a schematic diagram of onsite and offsite facilities. The extent of modifications for each section is depicted by the type of line used to outline each block of facilities. Four categories are shown:

- Existing facilities which will be used with essentially no modification.

- Plant sections in which a large portion of the existing facilities will be used but where additions or modifications are required. The additions in some of the offsite areas may be significant.

- Plant sections where significant modifications are required. In some instances, it may be more economic to replace rather than upgrade the existing facilities.

- Grass-roots facilities for which there is no existing counterpart.

The only existing systems used without modification are the coal feed lock hopper system and sulfur recovery facilities. However, major utilization of the following faciliites is possible: coal receipt and storage, acid gas removal, char withdrawal, fuel and utilities, waste treating and disposal, solids separation, buildings, shops, control room and laboratories. The existing coal preparation system, gas cooling and the gasifier can be used but only after significant additions or modifications. Entirely new facilities include catalyst receipt and storage, catalyst recovery, the cryogenic separation and the preheat furnace.

A summary of offsite additions and modifications for the Major Synthane Revamp is as follows:

- Catalyst addition - A ribbon blender is used to mix the coal with catalyst solution and a torus disk, steam heated drier is used to remove the water that enters with the catalyst solution.

- Fuel systems - New LNG storage and handling facilities are provided. A new 20,000 gal fuel oil storage tank is added along with a new distribution system to provide fuel to new furnaces and boilers as well as the relocated boiler.

- Steam - Two new high pressure steam boilers are added to provide for increased process steam loads and steam turbine spare drivers. The latter were added to critical service equipment to provide improved pilot plant operating reliability. Expansion of the demineralization plant is required to meet the new steam raising capacity of the plant.

- Power - Substation and switch gear changes and additions were defined. Because of the major plot plan changes and equipment relocations, existing electrical conduit, that is exposed in the piperacks, is replaced with underground wiring. The latter is considered safer and more reliable, and the expected incremental cost is small.

- Compressed air - A new air compressor, having the same capacity of each of the existing ones, is added. It has a steam turbine drive to ensure a supply of instrument air during a power failure.

- CO_2 and inert gas - Cryogenic pumps are provided for supplying 600 psi liquid CO_2 to the coal feed system. A vaporizer is provided onsite to supply the high pressure CO_2 that is required for the coal feed lock hoppers. The inert gas generation equipment and low pressure blowers are relocated.

- Waste Water Treating - A secondary treatment plant similar to the one provided in the grass roots pilot plant study design is provided. Most of the equipment is smaller, primarily because there is less rainwater run-off to process.

- Layout, Buildings and Site Preparation - In order to accommodate new facilities and comply with minimum required safety standards, it is necessary to relocate some existing equipment and buildings and use land that is outside the present fence line.

- The existing CO_2 compressor building is removed to make room for the cryogenic synthesis gas recovery equipment. (This eliminates the capability of using by-product CO_2 for the coal feed system. Purchased CO_2 will be used for this purpose for CCG operations).

- The O_2 and methanation facilities, which are not required for CCG operations, are eliminated.

- The cooling tower is relocated away from the process block.

- Realignment of the road that runs along the south side of the process block is necessary to meet minimum safe spacing requirements for the process furnaces.

A comparison of facilities differences between the grass-roots case and the Major Synthane Revamp is presented in Table 4-1. The major differences are in the coal preparation and steam reforming areas. Because of a lack of plot space, it was not possible to provide two trains of coal preparation equipment for the Synthane Revamp. This would likely result in a lower service factor for the revamp case.

In the steam reforming area, the reformer was sized to operate the plant at about 50% turndown when the synthesis gas recycle loop is not closed. The objective was to minimize the use of expensive LNG which must be consumed in the reformer. Additional design basis information for the Major Synthane Revamp as well as equipment lists and flow sketches are presented in Appendix 2.

4.6 INVESTMENT SUMMARY FOR MAJOR SYNTHANE REVAMP

The investment required to modify the Synthane Unit for CCG operation is estimated to be 150 M$. An investment breakdown is presented in Table 4-2. The cost for constructing a grass-roots LPP at a Gulf Coast location is 130 M$, as previously reported.

The investment for the Synthane Revamp includes escalation to an April 1, 1983 startup. The escalation basis is presented in Table 4-3. The schedule for the revamp is presented in Figure 4-2. As in the grass-roots case, this is based on obtaining LPP design data from a Process Development Unit (PDU) which begins operating in early 1979. The revamp schedule assumes prudent overlap with the PDU, and any delay in PDU operations would, therefore, delay the LPP schedule. The project execution time for the revamp is approximately four months longer than for a grass-roots pilot plant. This reflects delays for removal and relocation of existing equipment. Construction time also has to be spread out because of the high manning levels and limitations on the number of field labor personnel that can be effectively utilized on the congested site.

The investment for the Synthane Revamp is compared in Table 4-4 with the investment for the grass-roots LPP on a Gulf Coast location (see Section 2). To further illustrate the difference between the two cases, the investment for the Gulf Coast grass-roots facilities was adjusted to a Pittsburgh location basis. As shown in Table 4-4, the grass-roots investment of 130 M$ at a Gulf Coast location becomes 150 M$ at Bruceton. It should be emphasized that the location adjustment reflects only the difference in economic conditions. It does not reflect other differences such as climate, terrain, and specific site factors which could further increase the cost of building the grass-roots LPP in the Pittsburgh area.

The direct cost for the revamp is 41.5 M$ on a 1Q77 basis. A breakdown of the direct cost is presented in Table 4-5. This compares to a direct cost of 47.0 M$ for the grass-roots LPP at a Gulf Coast location and to 48.5 M$ for the grass-roots facilities on a Pittsburgh location basis. Most of the savings are in materials and result from reuse of the coal feed and acid gas removal facilities, pipeways, and the control room. Smaller materials savings were made in the areas of coal receipt, storage, and preparation. Subcontracts - which are principally for refractory lining of vessels and furnaces, installation of solids handling equipment, buildings, and site preparation - were 8 M$ in all three cases. Direct labor charges, on the other hand, were higher for the Synthane Revamp: 15.2 M$ versus 12.3 M$ - 14.7 M$ for the grass-roots cases. This is due to increased labor man-hours resulting from the need to relocate or remove equipment and lower labor productivity at a revamp site. In addition, both Pittsburgh location cases reflect a higher general labor cost than for the U.S. Gulf Coast. A section-by-section breakdown of the combined costs for direct material and labor and subcontracts is presented, along with comparable information for the grass-roots case in Table 4-6.

The savings in direct cost for the revamp is offset by increased indirect costs relative to the grass-roots case. For example, the indirect costs for the revamp are 44.6 M$, compared to 40.4 M$ for the grass-roots facilities on a Pittsburgh estimating basis and 35.5 M$ for a Gulf Coast basis. The increased indirect costs for the revamp are a result of the inefficiencies associated with a revamp job and differences in the productivity and payroll burden between the Gulf Coast and the Pittsburgh area.

As shown in Table 4-4, the escalation for the revamp is 34.2 M$ versus 32.9 M$ for Pittsburgh grass-roots estimate and 23.2 M$ for the Gulf Coast grass-roots case. The reason for the large difference between the Gulf Coast and Pittsburgh locations is a difference in labor escalation. The Gulf Coast grass-roots case is based on the use of open shop hiring and, under the terms of the Davis-Bacon Act, a higher initial wage rate which is subject to less escalation through the construction period. The revamp is based on a closed shop with wage rate escalation over the course of the contract.

Estimated costs for 2-1/2 years of operation of the revamped Synthane unit are 80 M$. A year-by-year breakdown of the operating cost components is presented in Table 4-7, and information on the estimating basis is presented in Table 4-8. Overall, the operating cost for the Synthane Revamp is approximately 10% higher than for the Gulf Coast, grass-roots CCG Pilot Plant. Principal cost increases are for fuel (17.4 vs. 11.0 M$) and maintenance (31.7 vs. 25.4 M$). There is a saving of 5.6 M$ in property taxes and land leasing costs for using the existing DOE property at Synthane.

The high cost for fuel at Synthane is the result of choosing LNG for feed to the synthesis gas generator (steam reformer). This choice was made because pipeline natural gas was not available, and the pipeline quality propane that is available is not suitable without considerable treatment. Adding treating facilities would have increased investment and utilities requirements and increased equipment layout problems. The increased operating cost is attributable to the higher labor cost in the Pittsburgh area.

TABLE 4-1

CATALYTIC GASIFICATION LPP
SUMMARY OF FACILITIES DIFFERENCES
GRASS ROOTS VERSUS MAJOR SYNTHANE REVAMP

Basis

- Each case is designed to get the data necessary to scaleup to a commercial plant with acceptable risk. Design feed rate is 92 ST/SD of as received Illinois No. 6 coal.

- Compromises which affect non-critical scaleup items or service factor had to be made for the Synthane revamp because of site specific limitations. These are:

Item	Grass Roots	Major Synthane Revamp	Comment
Coal Shipping	Covered RR Cars	Open RR Cars	Possible coal degradation. However, no apparent problems with current Synthane operation.
Coal Storage	Inert Silo	Open Pile	
Coal Preparation	Two Trains + One similar to unit in predevelopment program + Entrained system envisioned for commercial plant	One Train + Second train added in later turnaround	Service Factor Debit
Reformer	Two Furnace Cells 1 and 4 MSCFD H_2 Equiv.	One Furnace 3 MSCFD H_2 Equiv.	Coal Feed 42 ST/SD during Once Thru Syngas Operations
Gasifier Cyclones	External	Internal	Potential 2nd Stage Operability Problems
Acid Gas Removal	Heavy Glycol System	Benfield	Either process is acceptable
Equipment Spacing	All ERE Minimum Standards Met	Encroachments Evaluated on Item by Item Basis	Greater But Acceptable Fire Risk

TABLE 4-2

INVESTMENT SUMMARY FOR
MAJOR SYNTHANE REVAMP

Cost Breakdown	k$	
Material	18,300	
Labor	15,200	
Subcontracts	8,000	
Total Direct Costs (1Q77)		41,500
Payroll Burdens	7,200	
Field Labor Overheads	20,600	
Vendor Representatives	300	
Loss on Surplus	200	
Insurance	200	
Engineering	11,700	
Fees: Engineering, Construction & Royalty	4,400	
Total Indirect Costs (1Q77)		44,600
Total Prime Contract (1Q77)		86,100
Project Management Services		4,000
Escalation		24,200
		124,300
Project Contingency (20%)		24,900
Revamp Contingency		2,700
Total Erected Cost		151,900
CALL		150 M$

k = Thousand
M = Million

TABLE 4-3

BASIS FOR COST ESCALATION ESTIMATE - MAJOR SYNTHANE REVAMP

Escalation Rates Base Point--1Q77	Annual Percentage Material	 Labor	 Engineering
1st year	1	8	9
2nd year	8	8	9
3rd year	8	7	7
4th year	5	7	7
5th year	5	7	7
6th year	--	7	--
Centroid	July 1981	Aug. 1982	July 1981
Time from Base Point (yrs.)	4.25	5.33	4.25
Cumulative escalation effect, percent	25	47	39

TABLE 4-4

TOTAL ERECTED COST COMPARISON
MAJOR SYNTHANE REVAMP VERSUS GRASS ROOTS

Project Type	Revamp	Grass Roots	
Location	Bruceton, PA	Gulf Coast	Pittsburgh
Direct Costs, M$			
Material	18.3	27.0	26.0 (1)
Labor	15.2	12.3	14.7
Subcontracts	8.0	7.7	7.8
Direct Cost Total (1Q77)	41.5	47.0	48.5
Indirects	44.6	35.5	40.4
Project Management	4.0	3.8	4.1
Escalation	34.2	23.2	32.9
Total Cost Excluding Contingency	124.3	109.5	125.9
Project Contingency	24.9	21.9	25.2
Revamp Contingency	2.7	-	-
Total Erected Cost	151.9	131.4	151.1
CALL	150	130	150

Note:

(1) Lower material cost shown is to reflect sales tax differences between Pennsylvania and Texas.

TABLE 4-5

DIRECT COST SUMMARY - MAJOR SYNTHANE REVAMP

	----- 1Q77 Bruceton, Pa. -----		
	Material k$	Labor kMH	Subcontract k$
Onsites			
• Catalyst Recovery	1,830	130	--
• Gasification	2,700	180	500
• Product Gas Cleanup	920	60	--
• Acid Gas Removal	270	20	--
• Methane Recovery	3,150	105	--
• Steam Reforming	690	50	940
• Preheat Furnace	250	20	890
• Common Facilities	430	60	390
• Onsite Dismantling	--	65	--
Total Onsites	10,240	690	2,720
Offsites			
• Coal Receipt & Preparation	770	60	50
• Waste Treating	1,260	140	110
• Electrical	340	20	190
• Safety	430	35	--
• Site Preparation	140	55	1,310
• Layout	180	40	50
• Buildings	10	5	1,410
• Potable, Industrial, & Firewater	280	30	160
• LNG, LPG, & Fuel Oil	510	40	20
• Cooling Water	360	35	80
• CO_2 & Inert Gas	370	65	--
• Boilers & Steam Distribution	2,420	170	1,790
• Compressed Air	520	50	--
• Chemical Handling	190	15	60
• Catalyst Handling	280	20	50
Total Offsites	8,060	780	5,280
Total Onsites & Offsites	18,300	1,470	8,000

TABLE 4-6

DIRECT COST BREAKDOWN

GRASS ROOTS VERSUS MAJOR SYNTHANE REVAMP

Facilities	1Q77 Costs, k$ (3) Grass-Roots (Gulf Coast)	Synthane Revamp	Comments on Revamp
Onsites			
Coal Feed	1,595	----	Existing can be used at Synthane
Gasification	4,870	5,690	2 times labor for major revamp
Steam Reformer	2,415	2,320	Synthane units must handle dual fuels
Preheat Furnace	1,270	1,415	
Product Gas Cleanup	1,465	1,750	2 times labor for major revamp
Acid Gas Removal	2,655	545	New absorber
Methane Recovery	4,015	4,600	Duplicate of Grass Roots
Catalyst Recovery	2,650	3,625	Duplicate of Grass Roots
Common Facilities	2,215	1,650	Substantially less material
Unused Equipment Dismantling	--	900	Primarily methanation
Total Onsites	23,150	22,495	
Offsites			
Coal Receipt & Storage	3,185	--	Existing at Synthane
Coal Preparation	4,850	1,650	Only one train
Catalyst Handling	520	605	Duplicate of Grass Roots
Utilities			
Interconnecting Lines (1)	3,580	--	
Steam & BFW (2)	825	6,560	Relocation & major expansion
CO_2 and Inert Gas (2)	10	1,270	Relocation
Compressed Air	1,010	1,210	Relocation & expansion
Fuel Systems (2)	250	1,085	Pipeline NG not available
Cooling Water	735	925	Relocation & major expansion
Fire Protection	490	855	Includes all water systems
Chemicals Handling	320	455	Relocation & expansion
Electrical	1,450	805	Expansion & upgrading
Waste Water Treating	3,320	3,305	Smaller - some existing reused
Safety	510	915	Expansion
Buildings	1,610	1,490	Relocation
Layout	2,085	780	Expansion/compact
Site Preparation	1,080	2,210	Hilly terrain at Bruceton
Total Offsites	25,830	24,120	
Total Direct Costs	48,980	46,615	

Notes:

(1) Steam, nitrogen and fuel gas are supplied to the pilot plant and fuel gas product and acid gas streams are returned to the refinery.

(2) For the grass roots case, costs listed are for distribution lines within the pilot plant boundaries.

(3) Direct costs in this table include payroll burden on direct labor (2.0 M$ for grass-roots case and 5.1 M$ for Synthane case). Thus, the total direct cost for Synthane presented here is 5.1 M$ higher than the direct cost present in Table 1, where payroll burden for direct labor and field supervision is shown as a separate item.

TABLE 4-7

OPERATING COST SUMMARY - MAJOR SYNTHANE REVAMP

	2nd Half 1980	1981	1982	1983	1984	1st Half 1985	Total
	k$						
• Raw Materials							
- Coal (Illinois)	-	-	-	168	441	419	
- Catalyst (K_2CO_3 Sol'n)	-	-	-	251	293	164	
Total Raw Materials	0	0	0	419	734	583	1736
• Transportation							
- Coal	-	-	-	296	775	736	
- Catalyst	-	-	-	22	26	15	
Total Transportation	0	0	0	318	801	751	1870
• Salaries, Wages, Benefits, and Support Services							
	259	709	3500	7859	8219	4354	
Total S,W,B, and SS	259	709	3500	7859	8219	4354	24,900
• Administrative							
- Miscellaneous	10	80	155	185	195	98	
Total Administrative	10	80	155	185	195	98	723
• Technical							
- Miscellaneous	-	-	25	200	200	100	
Total Technical	0	0	25	200	200	100	525
• Process Operations							
- Catalyst & Chemicals	-	-	-	186	312	212	
- Utilities	-	-	-	5449	7441	4531	
- Process Services	-	-	10	34	35	35	
- Miscellaneous	-	-	20	20	20	20	
Total Process Operations	0	0	30	5689	7808	4798	18,325
• Mechanical							
- Labor	-	-	-	5639	5400	2374	
- Material	-	-	1000	6864	6572	2885	
- Miscellaneous	-	-	200	300	300	130	
Total Mechanical	0	0	1200	12803	12272	5389	31664
GRAND TOTAL	269	789	4910	27473	30229	16073	79743

TABLE 4-8

OPERATING COST ESTIMATING BASIS - MAJOR SYNTHANE REVAMP

Raw Materials

- Coal
  - Illinois No. 6 Bituminous
  - 1980 price - $22.50/ST
  - 15% contingency on annual requirements
  - escalated at 6.6% per year

- Catalyst
  - 47 wt% K_2CO_3 solution
  - 1976 price - $152/ST
  - 15% contingency on annual requirements
  - escalated at 6.6% per year

Transportation

- Coal
  - spot-shipment by rail from St. Louis to local supplier in Pittsburgh area
  - truck shipment from local supplier to plant site
  - 1977 rail shipping cost of $22.40/ST
  - 1977 truck shipping cost of $10.25/ST
  - escalated at 6.6% per year

- Catalyst
  - truck shipment from Niagara Falls to Pittsburgh
  - 1976 truck shipping cost of $13.60/ST
  - escalated at 6.6% per year

Salaries, Wages, Benefits, and Support Services

- Salaries, Wages, and Benefits
  - staff composed of 29 professionals and 70 technicians/operators during the 2½-year operating period
  - increasing portion of staff deployed onsite during the 2½-year engineering and construction period
  - salaries, wages, and benefits based on projected rates through 1982 and escalated at 6.6% per year through 1985

- Process and Technical Consultation
  - staff composed of 6 professionals during the 2½-year operating period
  - costs based on projected engineering billing rates

- Relocation Costs
  - relocate 13 professionals to and from Pittsburgh

Administrative

- Miscellaneous
  - office supplies
  - telephone service
  - plant security
  - travel

Technical

- Miscellaneous
  - laboratory technician salaries
  - supplies and services

Process Operations

- Catalysts and Chemicals
  - current costs obtained from Chemical Marketing Reporter
  - 15% contingency on annual requirements
  - escalation at 6.6% per year

- Utilities
  - electric power, potable water, industrial water: purchased from West Penn at prevailing rates
  - LNG, LPG, No. 2 fuel oil, liquid CO_2: purchased from local supplier at prevailing rates
  - 15% contingency on annual requirements
  - escalation at 6.6% per year

- Process Services
  - char disposal
  - fire fighting service

- Miscellaneous

  - radio system maintenance
  - safety equipment
  - tools
  - supplies

Mechanical

- Contract Labor and Supervision

  - direct mechanical labor based on an average of 175 men during plant operating period
  - one supervisor required for every ten direct labor men
  - wage rates based on current data from Synthane
  - escalation at 6.6% per year

- Maintenance Material

  - based on 150% of direct labor costs

- Miscellaneous

  - equipment rentals
  - supplies

FIGURE 4-1

BLOCK PROCESS DIAGRAM
MAJOR SYNTHANE MODIFICATION CASE

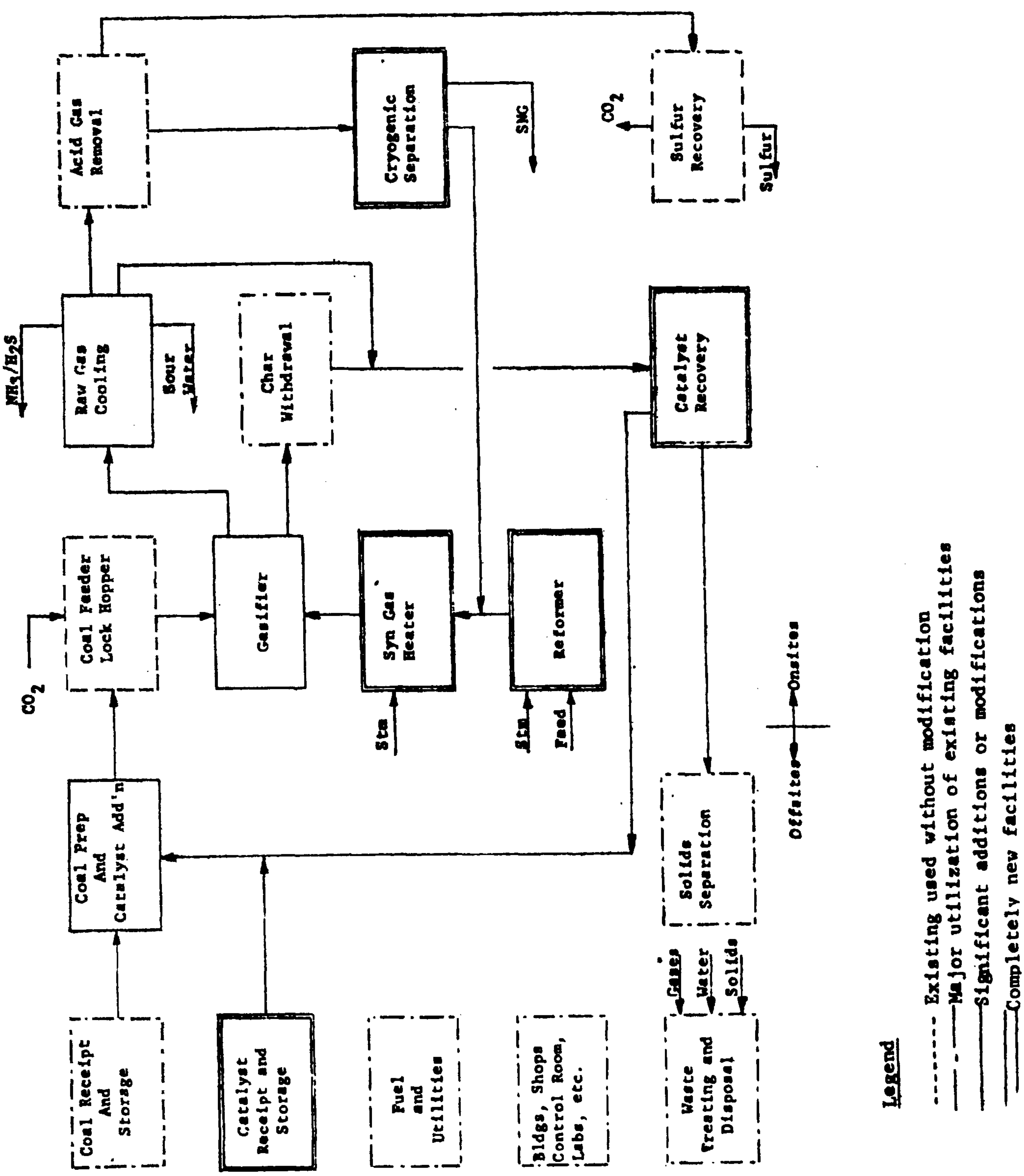
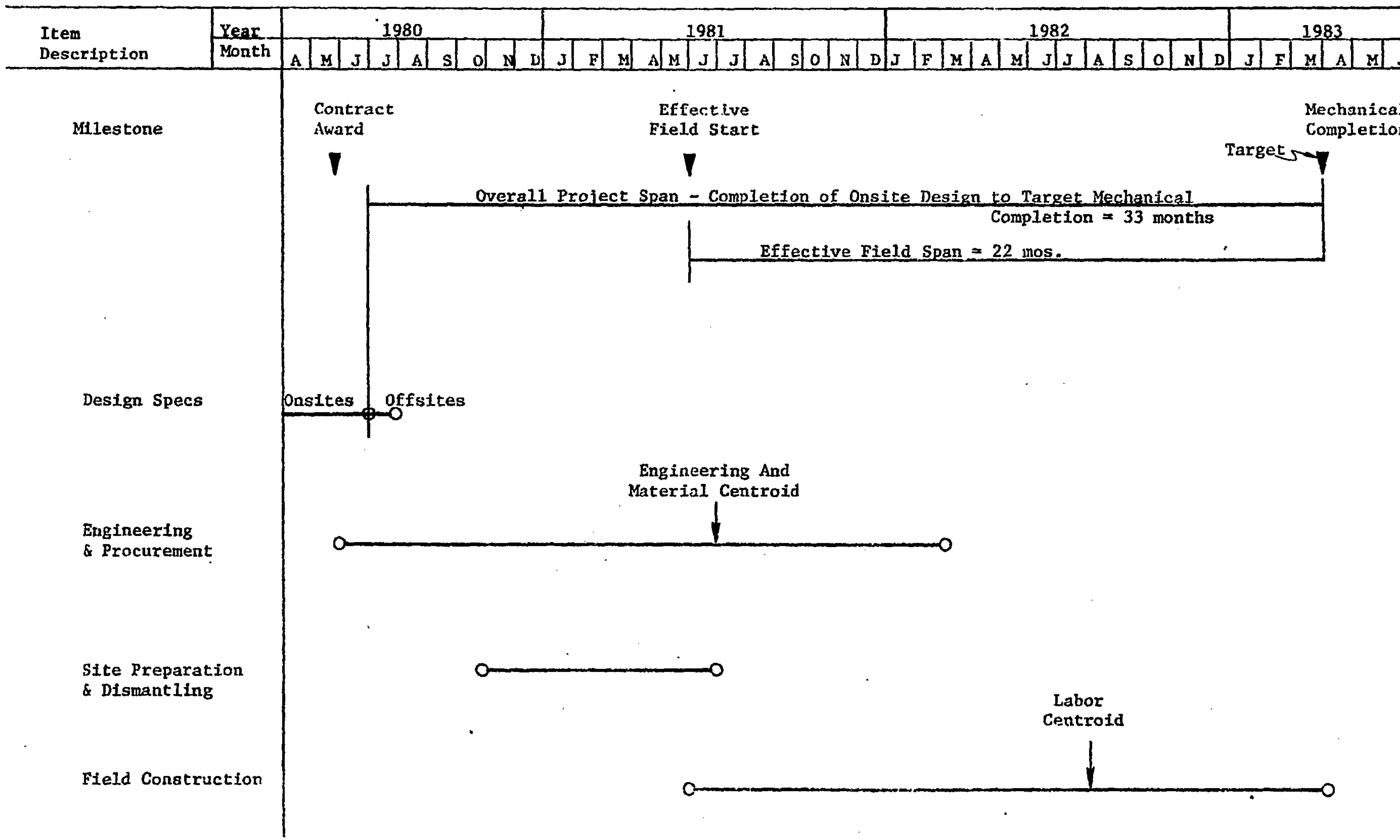


FIGURE 4-2

ENGINEERING AND CONSTRUCTION SCHEDULE - MAJOR SYNTHANE REVAMP



SECTION 5

STUDY DESIGN AND COST ESTIMATE FOR MINIMUM SYNTHANE REVAMP

5.1 PROJECT BASIS

The Project Basis for the Minimum Synthane Revamp is the same as that for the Major Synthane Revamp. The revamp is located at the existing Synthane site at Bruceton, Pennsylvania. Steam is generated at the site, electric power is purchased, and cooling water is provided by a recirculating system. The design feed coal is Illinois No. 6, the same composition as for the Grass-Roots Case, as shown in Section 2.

5.2 PROCESS BASIS

Since the Minimum Synthane Revamp Case was initiated near the end of the DOE Predevelopment Contract for CCG, the process basis was changed to reflect the data base developed under that contract. The major changes include a switch from K_2CO_3/Na_2CO_3 as makeup catalyst to KOH and the addition of calcium hydroxide digestion to the catalyst recovery system.

5.3 DESIGN PHILOSOPHY

The design philosophy for the Minimum Synthane Revamp was considerably different from that for the Grass-Roots and Major Synthane Revamp Cases. For the minimum revamp, the intent was to make the minimum changes to the existing facilities to allow operation in the catalytic mode in a technically meaningful fashion. Much of the equipment redundancy and operating flexibility built into the grass roots and major revamp cases was eliminated. Thus the plant service factor for the minimum revamp would be less than the service factor for the grass-roots or major revamp cases. As a result there is some risk that the large pilot plant would not be able to operate at steady-state for sufficiently long periods of time to obtain adequate scaleup data, and that, consequently, significant additional facilities modifications would be required with additional cost. Certain other compromises were made which would directionally increase the risk of scaleup to a pioneer plant. For instance, facilities for the separation and recycle of CO/H_2 from the product were not included since this would require major equipment additions and site changes. Although these compromises would directionally increase the risk of scaleup to a pioneer plant, it was judged that if the LPP operates with an adequate service factor, the scaleup risk still would be acceptable.

5.4 LPP SIZE

The maximum coal feed rate for the Minimum Synthane Revamp was set at 55 T/D of as-received Illinois No. 6 coal based on the capacity of the existing steam system. However, the existing Benfield acid gas removal system would also require expansion at coal feed rates in excess of

55 T/D. At the same time, the gasifier size was maintained at 3.5 feet diameter with a 100 ft. bed depth, as in the grass-roots and full revamp cases.

5.5 SUMMARY OF MODIFICATIONS FOR SYNTHANE REVAMP

The Minimum Revamp would be performed in two stages. In the first stage, operation would be once-through; that is, without the recycle of catalyst or synthesis gas. For this stage, design of the modifications could begin before startup of the PDU, since sufficient data are available from the Predevelopment Research Phase. The second construction stage would consist of adding catalyst recovery and recycle facilities. This could only be implemented after sufficient data were obtained from the PDU.

An important consideration for the minimum revamp was the source of high temperature synthesis gas for once-through operation. It was determined that the preferred approach would be to use partial oxidation of LPG. Relative to the steam reforming route specified in the Major Synthane Revamp, partial oxidation has the potential advantages of requiring fewer plant layout modifications, requiring less steam and allowing a shorter construction schedule. However, relative to steam reforming, partial oxidation produces a synthesis gas not as close in composition to that which actually would be obtained in recycle operations.

As noted above, utility system capacities were a major factor in setting the maximum coal feed rate. The steam system was judged to be the most difficult to expand due to space limitations; and with the existing steam system, the coal feed rate is limited to approximately 55 T/D of as-received Illinois No. 6 coal. Other utility systems were expanded as necessary to meet the 55 T/D feed rate.

Figure 5-1 is a schematic flowplan for the minimum modification case identifying the major processing blocks that: (1) can be utilized without major modifications, (2) will require substantial modifications, or (3) will be new facilities. The new and modified facilities are described below:

5.5.1 Catalyst Receipt and Storage

Potassium hydroxide will be used as gasification catalyst in the minimum revamp rather than a mixture of potassium and sodium carbonates, which was specified in the previous study designs. This change reflects results from the DOE Predevelopment Contract for Catalytic Coal Gasification (CCG). KOH will be purchased and stored as a 45 wt% solution and then diluted with industrial water before adding it to the coal.

5.5.2 Catalyst Addition and Final Drying

Coal is mixed with catalyst solution in a ribbon blender and then dried in an entrained drier system. The new equipment items include a feeder, the blender, a drying column, a drying air heater, gas solids separating equipment, and a circulating gas fan.

5.5.3 Synthesis Gas Generation

The synthesis gas generation system has been designed to produce sufficient gas to meet the requirements for feed coal injection, simulated synthesis gas recycle, and instrument purge. The synthesis gas generation system is composed of LPG and oxygen preheaters, a partial oxidation burner, and a recirculating, demineralized water cooling system for the burner. Additional equipment items which are included to produce coal feed injection gas are a waste heat boiler, gas cooler, condensate knockout drum, and an electric reheat furnace. All of the equipment specified in these two systems is new.

5.5.4 Raw Gas Quench

Particular attention has been paid to heat integration in the quench system so that utility demands will be minimized. Since little heat was available in the scrubber pumparounds, a heat exchanger that cools the gasifier effluent from 400°F to 325°F was added to make low level steam. The system now has two venturi scrubbers in series which, based on Synthane experience, are required to reduce the solids level in the gasifier effluent to acceptable levels.

5.5.5 Sour Slurry Stripper

The sour slurry stripper is substantially the same as the one provided in the major Synthane Revamp Study Design. However, the operating pressure was reduced to 35 psig so that low pressure steam could be used. This approach makes it possible to keep steam that is generated from process waste heat segregated from offsite boiler generated steam.

5.5.6 Char Quench

The char quench system will be reused with some modifications. The char cooling vessel (lower part of the existing gasifier) will be reused with the addition of an external cyclone and new top and bottom heads. Char quench and filtration facilities will be reused with slight modifications to the pumping equipment.

5.5.7 Gasifier

The existing Synthane gasifier has an inside diameter of 3-1/2 feet, the equivalent of that specified in the grass-roots case. However, 20-30 feet is the maximum bed depth which could be accommodated by the existing vessel. Since commercial gasifiers for catalytic gasification will likely use bed depths of about

100 feet, the existing gasifier would be expanded by adding a new section above the flanged head. This would permit operation closer to that projected for commercial plants. While a bed depth of 50-60 feet would be adequate to accommodate the 55 T/D coal feed rate, a brief study showed that the cost and schedule debits for specifying a full 100 ft. bed height would be small. The extra height will provide greater flexibility and enable the generation of more representative scaleup data.

5.5.8 Acid Gas Removal

An analysis of the Benfield unit showed that although it would be operating close to its existing capacity limit, it would be able to handle the 55 T/D gasifier feed rate without major revamp. To prevent flooding in the absorber at this feed rate, the CO_2 removal specification was relaxed, which would permit using a larger size packing. This would pose no problem since the product gas would be flared.

5.5.9 Offsites

The steam system was found to be adequate for the 55 T/D coal feed rate. Also, capacity expansion would be unnecessary for the boiler feed water system and cooling water facilities. Modifications to the electric power distribution system would be necessary to accommodate new drivers. Expansion of the LPG and oxygen supply facilities would be necessary to meet the requirements of the partial oxidation synthesis gas generation unit. The existing compressed air facilities would be used, but a third air compressor would be added to meet new requirements.

The existing potable and industrial water systems are presently fed from the same city water supply lines. In order to prevent potable water contamination via backflow from the industrial water system, facilities have been specified to segregate the two systems. Firewater facilities would be modified by providing a new pumping system to take suction from the new industrial water tank. Six hours of firewater storage is provided. The existing flare is adequate for CCG operation; all streams will flow to a new seal drum before entering the flare. However, conversion to once-through catalytic operation would result in the wastewater containing some water soluble catalyst salts which could plug the incinerator. In a commercial CCG plant, the wastewater stream containing catalyst would be recycled to catalyst recovery. After considerable investigation of alternatives, it was decided that the preferred pilot plant approach would be to concentrate the solutions to a small volume for disposal by a local waste disposal contractor. A submerged combustion unit is specified for this purpose.

5.5.10 Addition of Catalyst Recovery Facilities

In the second stage of the minimum Synthane Revamp, facilities are added to recover and recycle gasification catalyst. Although research work is still in progress to determine the preferred approach to recover catalyst, it was necessary to estimate roughly the future cost of adding such facilities in the second revamp stage. For this purpose, the Synthane catalyst recovery system was based on the system specified for the CCG Commercial Plant Study Design done under DOE Contract E(49-18)-2369. This system included $Ca(OH)_2$ digestion to free part of the water insoluble potassium followed by staged water wash to recover essentially all of the soluble catalyst. The solids-liquid separation is assumed to be made using hydroclones. Offsite facilities are included for lime receipt and handling. The preferred catalyst recovery system will be defined as part of the future DOE PDU program on CCG.

Additional design basis information, equipment lists, and flow sketches for the minimum revamp case are presented in Appendix 3.

5.6 INVESTMENT SUMMARY - MINIMUM SYNTHANE REVAMP

The cost of the first stage of the minimum revamp is estimated to be 46 M$ based on March 1, 1981 mechanical completion. Details of the cost estimate are presented in Table 5-2. The schedule is presented in Figure 5-2; it assumes that a Design Basis Memorandum summarizing the process basis for the first stage is completed by 1/1/79, and that the basic design would be completed by 7/1/79. The time from the start of detailed engineering to mechanical completion is 20 months. The shutdown period of the existing Synthane pilot plant for tie-in of new facilities is estimated to be six months as presented in the Estimate Basis in Table 5-3. A breakdown of the cost estimate into plant sections is presented in Table 5-4.

The cost of the second stage, catalyst recovery and recycle facilities, is estimated to be 12 M$. The actual cost for this stage will, of course, depend on the final process basis which will be developed as part of the ongoing research performed under DOE sponsorship. This cost estimate is based on completion of the basic design by 1/1/81, with mechanical completion on 5/1/82. The shutdown time for Stage 2 is expected to be minimal.

It would be possible to undertake a third stage modification to Synthane in which synthesis gas recycle facilities would be added. This would result in a pilot plant containing essentially all the process features of the "major revamp" case. However, because of the greatly increased plot area and utility requirements, this would entail major modifications to the existing Synthane Site, similar to those required for the major revamp. The cost for a third stage modification was not estimated but would be expected to be similar to that for the major revamp (150 M$ total).

The operating cost for the minimum Synthane Revamp is estimated to be 53 M$ for a 2-1/2 year operation period. An operating cost breakdown is presented in Table 5-5.

The investments and operating costs presented above are in escalated dollars assuming completion of the basic design and startup as specified. Delays in the schedule from that assumed above would result in somewhat higher costs due to further escalation.

TABLE 5-1

COMPARISON OF REVAMP ALTERNATIVES
CONVERSION OF SYNTHANE TO CCG OPERATION

	Total Revamp	Minimum Revamp
• **Design Basis**		
+ Coal Feed Rate, T/D Illinois No. 6 (as rec.)	92	55
+ Gasifier Bed Dimensions	3 1/2′ IDX 100′	3 1/2′ IDX 100′
+ Catalyst Recycle Loop Closed?	Yes	Yes
+ Synthesis Gas Recycle Loop Closed?	Yes	No
• **Project Scope**		
+ Relocation and Expansion of Existing Facilities	Major	Minimal
+ Backup for Utility Systems	Major	Minimal
+ Increase in Site Utilization	+50%	Minimal
• **Approximate Costs**		
+ Investment, M$	150	46 Gasification 12 Catalyst Recovery
+ Operating Cost, M$	80	52
+ Total Program Cost, M$	230	110
• **Timing**		
+ Design Basis Memorandum Preparation	3Q79	4Q78 Gasification 2Q80 Catalyst Recovery
+ Startup	1Q83	1Q81 Gasification 2Q82 Catalyst Recovery

TABLE 5-2

SYNTHANE MINIMUM MODIFICATION REVAMP
PROJECT COST SUMMARY

	1Q78 Bruceton k$ (1)		
	Stage I	Stage II	Total
Direct Costs			
Material	5500	1900	7400
Labor	5400	1400	6800
Subcontracts	1500	--	1500
Subtotal Direct Costs	12400	3300	15700
Indirect Costs			
Burdens	2200	600	2800
Field Labor Overheads	4900	1400	6300
Vendor Representatives/ Loss on Surplus/Insurance	300	100	400
Basic Design, Project Management	1800	800	2600
Detailed Engineering	5800	1100	6900
Fees: Engineering & Construction	2100	400	2500
Escalation 1Q78 → Phase I M.C.:1Q81, Phase II M.C.:2Q82	5000	2000	7000
Total Ex. Contingency	34500	9700	44200
Contingency (Phase I 33%, Phase II 26%)	11400	2500	13900
Total Erected Cost	45900	12200	58100
CALL	46 M$ (2)	12 M$ (2)	58 M$ (2)

Notes:

(1) k equals thousands
(2) M equals Millions

TABLE 5-3

ESTIMATE BASIS

SYNTHANE MINIMUM MODIFICATION REVAMP

The Estimate and Schedule were developed as for an Exxon project, i.e., Design Specifications and Project Management by ER&E and "normal" Exxon owner/ER&E interface. No attempt has been made to reflect the effect of the U.S. Government's ownership/control of the facility or execution of the project.

- Project Schedule
  - See Table 3.
- Material
  - Based on published 1Q78 cost levels (Domestic Purchase).
  - Includes delivery charges to site.
  - No local sales tax included. (Government Project exempt from tax.)
  - Pipe and structural steel to be shop fabricated.
- Labor
  - Source - Union Shop.
  - Productivity (vs. Standard Location - Gulf Coast, 1957 = 100%) 75%; adjusted for job size, activity, and revamp effects to 66%.
  - Work Week - 40 hrs. plus one hour of spot overtime. Schedule reflects double shifts during six-month Phase I Turnaround (~320 kMH).
  - Total direct manhours (ex. subcontracts) including contingency:
    - Stage I - 640 kMH
    - Stage II - 160 kMH
  - Average direct wage rate (excluding burdens) escalated to labor centroid:
    - Stage I - $13.71/MH (reflects T/A shift premium)
    - Stage II - $15.11/MH
- Field Labor Overheads
  - Include temporary construction and consumables, field supervision, and construction equipment.
  - Reflect published percentages of total direct labor for Pittsburgh location, adjusted for job size and revamp nature of project.
- Burden
  - Employer's payments for retirement plans, workman's compensation, sick pay, vacations, etc. of the direct and indirect labor.
  - Reflects Pittsurgh area union labor benefits.

TABLE 5-3 (Cont'd.)

- **Detailed Engineering**
  - U. S. prime contractor
  - Total reimbursable manhours (including contingency)
    - + Stage I -275 KMH
    - + Stage II - 45 KMH
  - Engineering (including design and drafting) individual rates escalated to the centroid of engineering (average rates, excluding fee, shown below)
    - + Stage I -$33.57/MH
    - + Stage II -$39.59/MH
  - Net productivity (typical U. S. prime contractor) - 90%
  - No allowance has been made for split engineering and field contractors.

- **Fees**
  - Based on published 1Q78 rates.

- **Escalation**
  - The following escalation rates are included in the estimate:

	ESCALATION FROM 1Q78		
	Mat'l	Labor	Engineering
1st year	3%	8%	9%
2nd year	6%	7%	9%
3rd year	7%	7%	9%
4th year	7%	7%	9%
5th year	-	7%	-
Stage I			
Centroid	4Q79	4Q80	1Q80
Time from 1Q78 (yrs)	1-3/4	2-3/4	2
Escalation Effect	8%	22%	19%
Stage II			
Centroid	3Q81	1Q82	3Q81
Time from 1Q78 (yrs)	3-1/2	4	3-1/2
Escalation Effect	13%	32%	35%

- **Contingency**
  - A project contingency of 25% has been included in the estimate.
  - A revamp contingency of 10% (on the revamp portion only) has been included in the estimate
    - + Stage I Revamp contingency - 8%
    - + Stage II Revamp contingency - 1%
  - Total contingencies are 33% and 26% for Phases I and II, respectively.

TABLE 5-3 (Cont'd.)

- Major Estimate Exclusions

  - Maintenance equipment.
  - Catalyst requirements.
  - Research Expense.
  - Process facility rental charges and leasing costs.
  - Owner's non-recurring expenditures such as:

    + Operating personnel wages during construction period.
    + Construction period office operating costs.
    + Startup costs.
    + Inventory costs.
    + Organization and financing costs.
    + Warehouse spare parts.

  - Salvage value of dismantled equipment.
  - Reconditioning of reused equipment.
  - Data logger.
  - Plant modifications after mechanical completion.
  - Special chromatagraphic analysis system.

TABLE 5-4

SYNTHANE MINIMUM MODIFICATION REVAMP
DIRECT COST SUMMARY

	1Q78 BRUCETON, PA.		
	Material k$	Labor k$	Subcontracts k$
STAGE I			
Onsites			
• Gasification/Char Withdrawal	1010	1180	460
• Product Gas Cleanup	980	790	-
• Acid Gas Removal	190	110	-
• Synthesis Gas Generation	640	340	-
• Onsite Dismantling	40	280	-
• Common Facilities/Tie-Ins	340	680	550
Total Onsites	3200	3380	1010
Offsites			
• Coal Receipt and Preparation	660	620	60
• Waste Treating	760	340	50
• Electrical	100	110	-
• Site Prep	10	60	40
• Buildings	-	-	60
• Potable, Industrial & Firewater	200	280	210
• Fuel Systems	260	280	-
• CO_2 & Inert Gas	60	110	-
• Compressed Air	150	110	-
• Catalyst Handling	100	110	70
Total Offsites	2300	2020	490
TOTAL STAGE I	5500	5400	1500
STAGE II			
Onsites			
• Catalyst Recovery	1830	1290	-
Total Onsites	1830	1290	-
Offsites			
• Lime Handling	70	110	-
Total Offsites	70	110	-
TOTAL STAGE II	1900	1400	-
TOTAL STAGES I AND II	7400	6800	1500

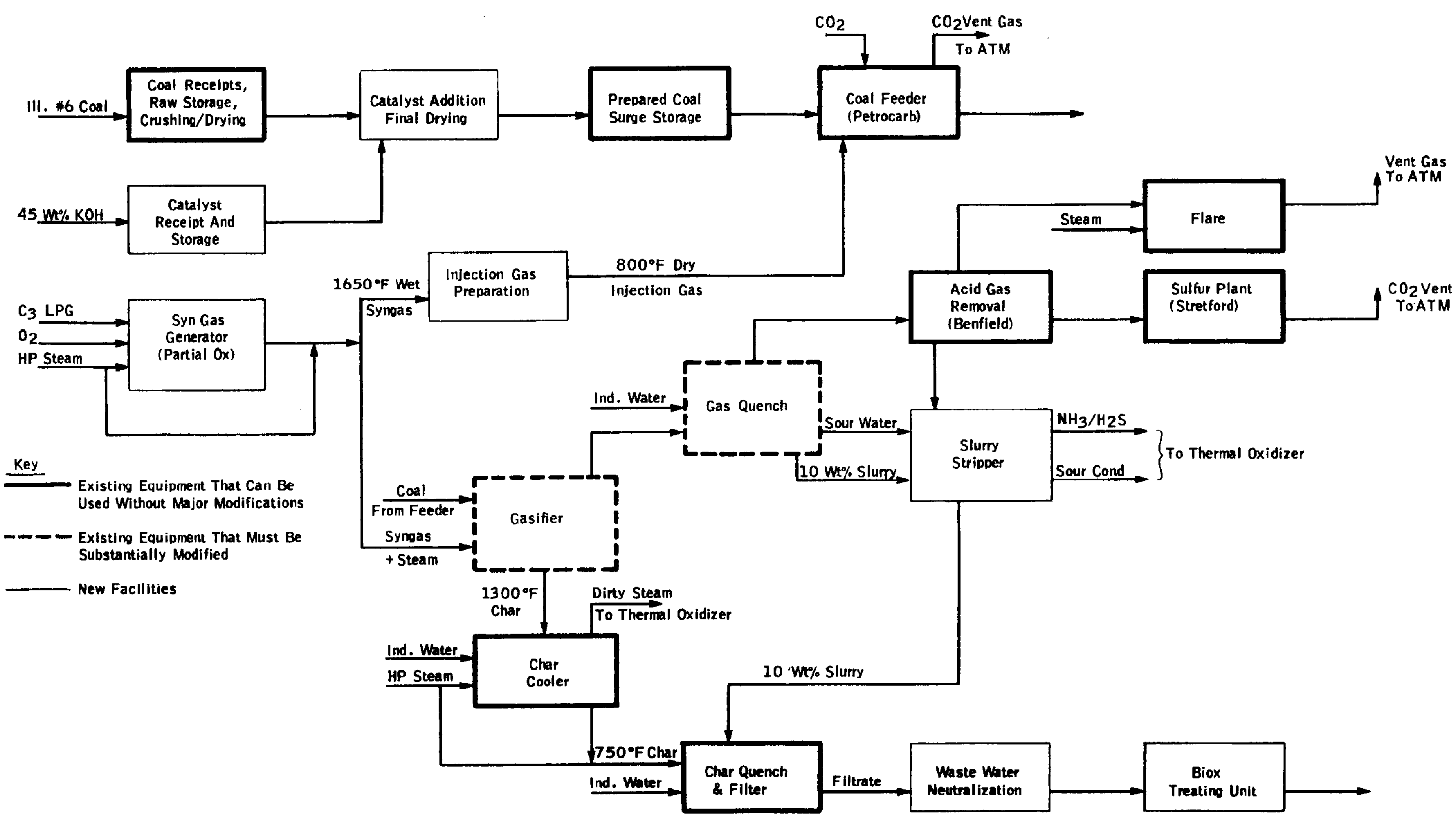
TABLE 5-5

SYNTHANE MINIMUM MODIFICATION REVAMP
OPERATING COST SUMMARY

	1980	1981	1982	First Half 1983	Total
	←		k$		→
• **Raw Materials**					
Coal	–	95	210	215	
Catalyst	–	195	135	40	
Total	–	290	345	255	890
• **Transportation**					
Coal	–	165	365	365	
Catalyst	–	20	15	5	
Total	–	185	380	370	935
• **Salaries, Wages**					
Benefits	860	6,060	7,235	5,745	
Total	860	6,060	7,235	5,745	19,900
• **Administrative**					
Miscellaneous	40	160	170	130	
Total	40	160	170	130	500
• **Technical**					
Miscellaneous	25	175	200	150	
Total	25	175	200	150	550
• **Process Operation**					
Catalyst and Chemicals	–	2,790	5,815	5,480	
Utilities	–	535	1,170	1,200	
Process Services	–	130	35	50	
Miscellaneous	–	15	20	30	
Total	–	3,470	7,040	6,760	17,270
• **Mechanical**					
Labor	–	1,435	1,835	2,720	
Material	600	2,345	2,235	1,670	
Miscellaneous	140	300	210	135	
Total	740	4,080	4,280	4,525	13,625
Grand Total	1,665	14,420	19,650	17,935	53,670

Figure 5-1

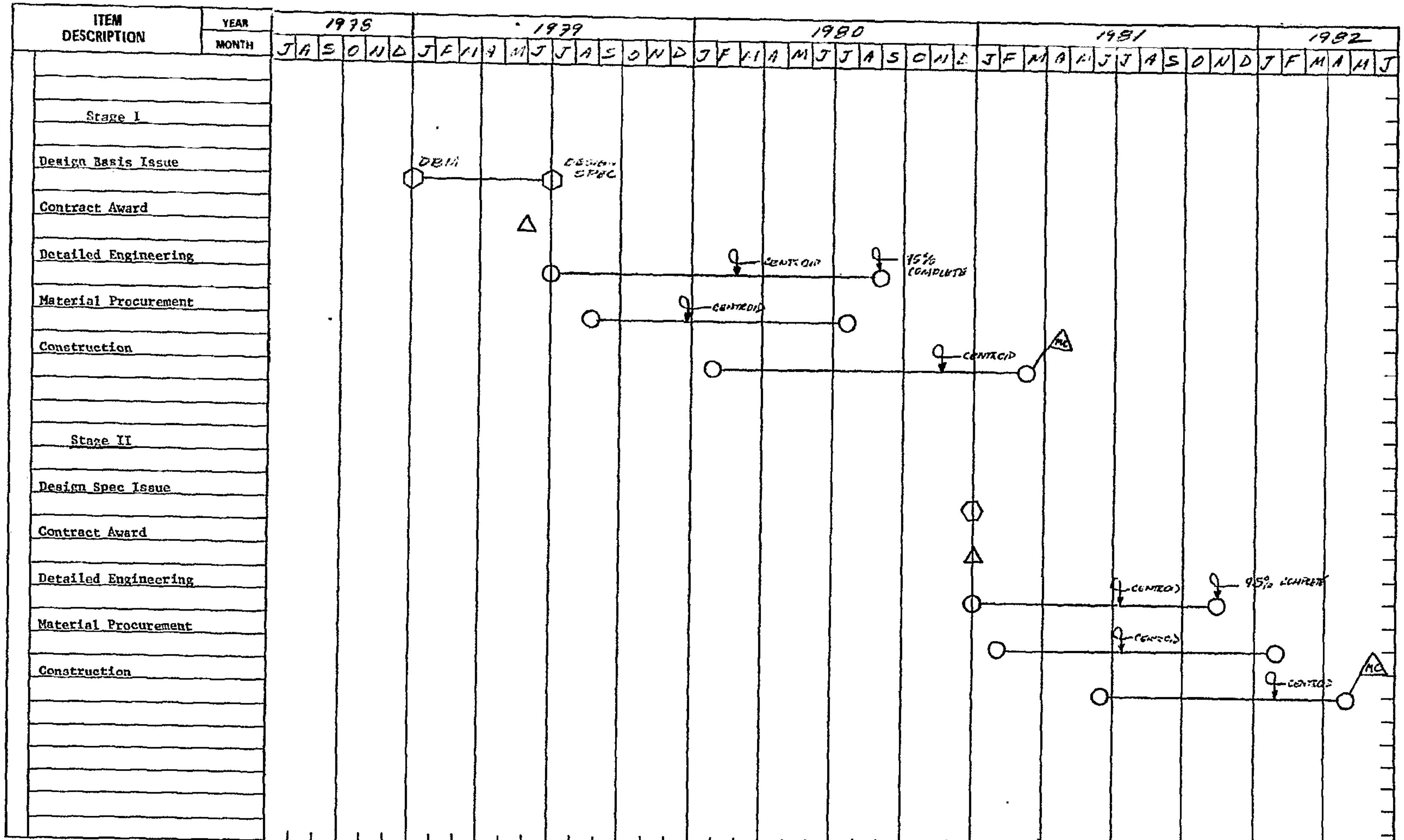
SYNTHANE REVAMP
MINIMUM MODIFICATION CASE



39 months

FIGURE 5-2

SYNTHANE MINIMUM MODIFICATION REVAMP
PROJECT SCHEDULE



Phase I

- Total direct manhours including contingency = 640 kMH.
- Turnaround (~320 kMH) is approx. 6 months.
- Peak labor = 400 men.
- Avg. labor = 250 men.

Phase II

- Total direct manhours including contingency = 160 kMH.
- Peak Labor = 140 men.
- Avg. Labor = 90 men.

EXXON RESEARCH & ENGINEERING CO.
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Class V Project Schedule

Synthane Minimum Modification Revamp

DATE	SHEET OF	DRAWING NO.
12 July 1978		