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SEPTEMBER 1978 STUDY REPORT

VOLUME I - SUMMARY

SALES GAS CONDITIONING FACILITIES PRUDHOE BAY, ALASKA

AMERADA HESS CORPORATION
ATLANTIC RICHFIELD COMPANY - STUDY COORDINATOR
EXXON COMPANY, U.S.A.
GETTY OIL COMPANY
MOBIL OIL CORPORATION
NATURAL GAS CORPORATION OF CALIFORNIA
NORTHERN NATURAL GAS COMPANY
NORTHWEST PIPELINE CORPORATION
PACIFIC INTERSTATE TRANSMISSION COMPANY (ARCTIC)
PANHANDLE EASTERN PIPELINE COMPANY
PHILLIPS PETROLEUM COMPANY
SOHIO PETROLEUM COMPANY
SOUTHERN NATURAL GAS COMPANY
TENNESSEE GAS PIPELINE COMPANY
TEXAS EASTERN TRANSMISSION CORPORATION
TEXAS GAS TRANSMISSION CORPORATION
TRANSCONTINENTAL PIPELINE CORPORATION
UNITED GAS PIPELINE COMPANY

Job No. 5795-1

The Ralph M. Parsons Company
Engineers / Constructors

SALES GAS CONDITIONING FACILITIES STUDY
PRUDHOE BAY, ALASKA

VOLUME I - SUMMARY

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1. INTRODUCTION

In July, 1977, a group of companies, ultimately eighteen in number, implemented a study by The Ralph M. Parsons Company. The objectives of this Study were to:

- Develop a preliminary design for the Facilities necessary to condition gas produced from the Prudhoe Bay Unit for transmission to the "Lower 48."
- Develop milestone schedules and implementation plans for both a full-capacity startup by July 1, 1983 and a partial-capacity startup by January 1, 1983 with full capacity operation by July 1, 1983.*
- Prepare cost estimate with a plus or minus 20 percent accuracy for these Facilities.
- After appropriate compensation make the Study results available to those parties who ultimately build the Facilities.

In order to accomplish these overall objectives the following major tasks were undertaken:

- Screen the processes available and select a process for carbon dioxide removal and a process for hydrocarbon dewpoint control.
- Develop a process design for conditioning the gas to meet an assumed set of pipeline delivery conditions and quantity and quality specifications.
- Define the auxiliaries and support facilities required for the process plant.
- Prepare equipment data sheets for major equipment.
- Develop a plant layout.
- Identify the permit requirements in the areas of safety, environment and construction.
- Prepare a final report.

The use and availability of the existing first-stage gas injection compressors in the Prudhoe Bay Unit Central Compression Plant to boost the

*At the time this study was implemented the announced startup date for the gas transmission pipeline was January, 1983.

conditioned gas to pipeline pressure of 1440 psig* were assumed as the Study base case. The use of these compression facilities and certain other facilities, owned by the Prudhoe Bay Unit or others, has been assumed in this Study. There is no assurance that these facilities will be available nor have the present owners approved their use. A stand-alone compression plant was considered as an alternate case. The combination of these two compressor options and the two timing options resulted in four different study cases as follows:

- A full-capacity startup July, 1983, assuming use of existing Central Compression Plant first-stage injection compressors to boost gas to pipeline delivery pressure. (Base Case Full Startup.)
- A phased startup, partial capacity in January 1983, full capacity in July, 1983, assuming use of existing Central Compression Plant first-stage injection compressors to boost gas to pipeline delivery pressure. (Base Case Phased Startup.)
- A full-capacity startup in July 1983, using new stand-alone booster compressors to boost to pipeline delivery pressure. (Alternate Case Full Startup.)
- A phased startup, partial capacity in January, 1983, full capacity in July, 1983, using new stand-alone booster compressors to boost to pipeline delivery pressure. (Alternate Case Phased Startup.)

Several important factors which were considered in all aspects of the study included:

- The safety aspects of the Facilities from the point of view of both compliance to applicable regulations and conformance with good engineering practice.
- The continuity of the crude oil production from the Prudhoe Bay Unit.
- The continuity of gas pipeline deliveries.

The results of this study are contained in the following volumes of this report: Volumes II, V, and VI contain confidential SELEXOL® process information. This information has been extracted from distributions to those parties who have not executed a secrecy agreement with Allied Chemical Company.

*At the time the scope of study was defined, this delivery pressure appeared to be a most probable compromise between the economies of higher pressure transmission and the adherence to "proven pipeline technology." It should be noted that the process selected is independent of pipeline pressure within proposed pressure extremes.

VOLUME I	-	SUMMARY contains an overview and a summary of the Study.
VOLUME II	-	FACILITY DESIGN details the process selection work and design.
VOLUME III	-	IMPLEMENTATION PLAN AND COST ESTIMATE SUMMARY presents the plan for design, procurement and construction, and a summary of the cost estimate.
VOLUME IV	-	APPENDIX 1 provides details on Study scope, schedule, and instructions, and a description of related facilities now located at Prudhoe Bay.
VOLUME V	-	APPENDIX 2 contains equipment design specifications and data sheets.
VOLUMES VI/VII	-	FACILITIES COST ESTIMATE, PARTS 1 AND 2 gives the detailed development of the cost estimate.

A detailed Table of Contents for the report is located at the beginning of volumes II through VII. All figures referenced herein are located at the rear of this volume.

2. OVERVIEW OF STUDY RESULTS

The Sales Gas Conditioning Facilities cost estimate was developed with an accuracy of plus or minus 20 percent before contingency was included. This estimate was based on a process design to condition Prudhoe Bay Unit gas to meet an assumed intermediate set of pipeline gas quality and delivery specifications based on the several filings for the proposed pipelines before the United States and Canadian Governments. This gas is sometimes referred to throughout this Report as "sales gas," without regard to the point of sale. The volume of gas available for delivery to the pipeline was as set forth in the Prudhoe Bay Unit Agreement of April, 1977, and in the Field Rules established by the State of Alaska, Division of Oil and Gas Conservation in May, 1977. Assumed study gas delivery and quality specifications were:

Delivery Volume (nominal)	2,000 MMSCFD
Delivery Pressure	1,440 psig
Delivery Temperature (max.)	25°F
Carbon Dioxide Content (max.)	1.0 volume %
Hydrogen Sulfide Content (max.)	1.0 grain/100 SCF
Hydrocarbon Dewpoint (max.)	-10°F @ 1,100 psia
Water Dewpoint (max.)	-35°F @ 1,100 psia

The specified sales gas volume requires a field production rate of 2,600 to 2,800 MMSCFD to account for projected total field and local fuel gas demands (170 to 345 MMSCFD) and shrinkage from gas conditioning (430 to 455 MMSCFD).

The investment requirements for the previously described four cases for the Facilities, including a contingency as determined by risk analysis, are estimated as follows:

	1978 Dollars	Escalated Dollars ('83 dollars)
Base Case Full Start	1,571,500,000	1,962,300,000
Base Case Phased Start	1,581,200,000	1,964,500,000
Alternate Case Full Start	1,688,400,000	2,109,700,000
Alternate Case Phased Start	1,701,700,000	2,107,200,000

A risk analysis was performed to assess the contingency required. In accordance with normal practice in the Engineering Contractor industry, the contingency included in the estimate provides a 90% probability that the project can be completed, as presently designed and scheduled, for the above estimated investment.

The bases and method of preparation for the cost estimate are summarized in Section 11 of this volume.

The process selected for carbon dioxide removal was Allied Chemical's SELEXOL physical solvent process. A mechanical refrigeration process was selected for hydrocarbon dewpoint control. Water dewpoint control is accomplished in the dehydration equipment located in the existing Prudhoe Bay Unit gas/crude oil separation sites called Gathering Centers and Flow Stations. The hydrogen sulfide content of the feed gas is less than maximum specification limit. The hydrogen sulfide removal by the SELEXOL process provides added assurance of meeting this specification. Therefore, it was assumed that no process equipment was required in the Sales Gas Conditioning Facilities for either water dewpoint control or hydrogen sulfide removal.

A Block Flow Diagram, Figure 1, illustrates the basic process flow of the Sales Gas Conditioning Facilities. Feed gas, originating from the gas/crude separators is compressed in the Gathering Centers and Flow Stations and flows to the inlet separation unit. The inlet gas streams are metered, and any solids or free liquids in the gas are removed at this point. The feed gas flows first to the Natural Gas Liquids Extraction section for hydrocarbon dewpoint control. The gas then flows to the SELEXOL section where the carbon dioxide is removed. The conditioned gas then goes to the gas compressors where it is boosted to pipeline pressure, then refrigerated for transmission.

A small portion of the feed gas is fed to the existing Prudhoe Bay Unit Field Fuel Gas Unit. Here a special low water dewpoint and hydrocarbon dewpoint fuel is produced for delivery to the Trans Alaskan Pipeline System to supply certain of its fuel requirements.

SELEXOL solvent characteristically absorbs, along with the carbon dioxide,a significant quantity of hydrocarbons, particularly the heavier hydrocarbons. During the regeneration of the SELEXOL solvent both the carbon dioxide (about 340 MMSCFD) and these hydrocarbons are flashed from the solvent. Thus, the flash gases from the SELEXOL solvent regeneration have a potentially usable fuel content.

To conserve this fuel content of the flash gases, a plan was developed whereby equipment to burn low heating value fuel was utilized in the Sales Gas Conditioning Facilities. Any excess flash gases over local needs would then be utilized in the Prudhoe Bay Unit Facilities after special heating value blending and with field handling precautions. The intermediate pressure (IP) flash gas is compressed and used directly as fuel. The low-pressure flash gas is compressed and fractionated in the Local Fuel Fractionator (LFF) for control of heating value, and the effluent gas is used as fuel. The carbon dioxide in these fuels, which are utilized in the fuel burning equipment, vents in the exhaust at dispersed locations throughout the field.

The hydrocarbon liquids from the NGL Extraction and SELEXOL low-pressure flash gas are separated in the fractionation unit into propane, butanes, and pentanes plus products to facilitate disposal. Some propane is used for heating value control of certain fuel streams. The remaining propane is injected into the pipeline gas. The butanes are either injected into the pipeline gas up to hydrocarbon dewpoint limits or into the crude oil delivered to the Trans Alaskan Pipeline System. The pentanes-plus are injected into the same crude stream.

It has been assumed that any excess liquid streams and flash gases will be injected into the oil-producing reservoir during the following circumstances:

- Imbalance of supply/demand
- Startup or shutdown
- Equipment outage
- Unit upset
- Other emergencies

A summary material balance is presented in Table 2-1. Sufficient flexibility has been designed into the Sales Gas Conditioning Facilities so that a nominal 2,000 MMSCFD can be produced under virtually all circumstances. The material balance in this table reflects an operation at the maximum anticipated fuel demand for the Prudhoe Bay Unit. It is assumed, though neither assured nor approved, that the Prudhoe Bay Unit will accommodate, at some cost, the high carbon dioxide, high hydrocarbon dewpoint fuel.

Table 2-1. Summary Material Balance

Stream	Volume (MMSCFD)(1)	BPD
Sales Gas (net before any NGL blending)	2000.6	-
Propane (balance over fuel blending)	37.2	24.270
Butanes	41.3	31,290
Pentanes-plus	31.9	28,700
Fuels [approxiimately 50% CO_2](2) 689.0		-
Total Feed	2800.0	

(1) Standard Conditions 14.65 psia and 60°F.

(2) Virtually all the CO_2 removed from the feed gas is reinjected into the fuel streams.

The Facilities require 682,000 total installed horsepower including motors, power recovery units and gas turbines. The bulk of this horsepower is developed by 17 operating gas turbines with 6 spare gas turbines. The flow diagram (Figure 1) shows blocks reflecting the major process and auxiliary systems required. The major auxiliary systems include:

- Refrigeration
- Offsite and general utilities
- Power Generation

A more detailed process flow is shown in Figure 2.

To support the preparation of the estimate at a plus or minus 20 percent accuracy level, the following estimate bases, in addition to the process flow diagrams, were prepared:

- Data sheets for major processes and equipment.
- Detailed plant and support facilities module layout drawings.
- Major piping layout drawings.
- Project implementation plan.

The site location in respect to the entire Prudhoe Bay complex is shown in Figure 3. Figure 4 is a base case design layout of the plant, and also illustrates location of modules for alternate cases and future expansion.

An Implementation Plan provides details for the Facilities funding, design, procurement, fabrication, logistics, and installation.

<u>A full startup of the Sales Gas Conditioning Facilities, by July, 1983, requires strict adherence to the Plan.</u>

A critical timing factor in any construction program at Prudhoe Bay is the "ice window" in the Beaufort Sea. Major plant components can only be delivered via ocean-going barges during the short (4-6 weeks) period each year when the "ice window" is open. Failure to deliver any critical major component during the scheduled period would effectively delay full-capacity startup by one full year.

A milestone schedule for this implementation plan is shown in Figure 5. The critical elements of this schedule include:

- Completion of the process design by January, 1979.
- Timely preparation of all background material, particularly environmental information, necessary to obtain the required permits.

- Long-lead-time equipment to be ordered during the first quarter of 1979.
- Resolution of the impact on and the use of existing facilities at Prudhoe Bay.
- Selection of "Lower 48" fabrication sites.

3. DESIGN BASIS AND CONSIDERATIONS

The Sales Gas Conditioning Facilities preliminary design developed for this Study is based on conditioning gas from the Prudhoe Bay Unit to meet the assumed set of specifications for the pipeline gas as set forth in the Overview (Section 2 of this volume).

The feed gas to the Sales Gas Conditioning Facilities originates from the gas/crude oil separators located in the three Gathering Centers on the West side of the field and the three Flow Stations on the East side of the field. This gas is presently dehydrated in the Gathering Centers and Flow Stations. It was assumed that no further dehydration over current design capability is required to meet the sales gas water dewpoint specification of -35°F @ 1,100 psia. The hydrogen sulfide content of the feed gas is sufficiently low (0.50 grain/100 SCF) so that no hydrogen sulfide removal is required to meet the sales gas specification of 1.0 grain/100 SCF.

For design purposes the feed gas composition was assumed to be that associated with a crude production of 1,500,000 BPD with equal amounts of both crude and feed gas being produced from the East and West sides of the field. The gas composition derived from this crude oil production rate results in a maximum quantity carbon dioxide and NGL feed to the facilities. Lower crude rates would result in increased NGL blending in the pipeline gas and a slightly lower volume of NGL delivered to the Trans Alaskan Pipeline System.

Module and facility layout and design considered the following factors:

- Safety.
- Constraints because of Prudhoe Bay climatic conditions.
- Reliability of operation.
- Ease and economy of construction.

The remoteness and the severe environmental conditions at Prudhoe Bay impose limitations on both the process and mechanical design of the Facilities. The low ambient temperatures (-60°F) and high winds dictate that all equipment be housed in totally enclosed modules for protection of both personnel and equipment. The study assumed the use of safe and efficient methods for Arctic construction used by ARCo., EXXON and SOHIO during development of the Prudhoe Bay Field. Modules with contained equipment are to be fabricated in the "Lower 48." They would be sea-lifted to Prudhoe Bay by ocean-going barges. At Prudhoe Bay they would be offloaded by crawler transportors or rubber-tired vehicles and moved to their pile supports on graveled sites. The pile supports, gravel insulation and an air space below the modules prevent the permafrost from thawing.

Transport by the combination of barge and crawler or rubber-tired vehicles imposes both weight and height limitations on the modules. This in turn establishes the number of carbon dioxide removal trains at four because of the weight and height limitations for the carbon dioxide absorbers. Weight limitations, to some extent, dictated equipment layout. The use of pile supports plus fully housed equipment made it desirable to minimize plot area or "footprint." Thus, vertical vessels were utilized in preference to horizontal vessels. Equipment is stacked wherever possible to minimize footprint.

The process and mechanical design safety factors adopted were consistent with industry practice and recognized the unique conditions prevailing at Prudhoe Bay.

The reliability philosophy that was adopted for the Facilities was aimed at insuring continuity of pipeline gas flow so as not to adversely affect Prudhoe Bay Unit oil production. Several different approaches were utilized to implement this philosophy:

- The use of a spare parallel train such as the fourth train in the carbon dioxide removal section.
- The use of spare equipment such as pumps and compressors.
- The use of commercially proven technology.
- The use of alternate "handling" such as injection for NGL.

Application of this approach generally led to the selection of conservative processing techniques and proven mechanical equipment.

The safety of personnel and equipment was a major design consideration. The following factors were included:

- Compliance with applicable safety regulations.
- Conformance with good engineering practice.
- Consideration of Prudhoe Bay environment.

Where one design was obviously more economical to install and/or operate than another, the less costly was chosen. In some cases process or equipment selections were made without detailed study because the available economic guidelines did not clearly indicate which was preferable. In some cases additional study of these alternatives has been recommended for the future detailed engineering design phase.

Space was allowed and the major process lines were sized to permit a 50% expansion of the gas conditioning capacity. The allowance included space for both process and auxiliary equipment.

4. PROCESS SELECTION

In July of 1977 a Sponsor Study Team was formed to screen various processes for removal of carbon dioxide (CO_2) from gas produced from the Prudhoe Bay Unit. With the support of The Ralph M. Parsons Company, potentially competitive commercialized processes were evaluated including:

Allied Chemical	SELEXOL®
Fluor	Propylene Carbonate
Shell	SUFINOL®
Union Oil	SORBCO-2®
Lotepro	RECTISOL®
Lurgi	PURISOL®
Open-art	DEA

As the result of this study it was determined that the Lotepro, Lurgi and Open-art DEA processes were not competitive in this application and the Union Oil process did not have adequate commercial experience.

Following this initial evaluation the decision was made by the Sponsor's Technical Advisory Committee to evaluate the Allied Chemical, Fluor and Shell processes in more detail. In addition various processes for hydrocarbon dewpoint control (NGL extraction) were to be screened in combination with the CO_2 extraction processes.

The selection of an NGL extraction process was significantly influenced by the pressure level for CO_2 removal. Three primary schemes were considered:

- NGL extraction at 540 psia by mechanical refrigeration, compression to 1400 psia, followed by CO_2 removal at 1400 psia.
- Compression to 1500-1800 psia, expansion to 1000 psia with NGL extraction by expansion refrigeration, CO_2 removal at 1000 psia, followed by booster compression to pipeline pressure.
- NGL extraction by mechanical refrigeration and CO_2 removal at 540 psia, followed by compression to pipeline pressure.

The last scheme, with both NGL extraction and CO_2 removal at 540 psia, was selected.

Preliminary appraisal of the first scheme appeared attractive since, for both physical and chemical CO_2 removal processes, both the solvent circulation rates and the equipment sizes were smaller at the higher pressures. However, it was found that at high pressure, all of the CO_2 removal processes required more absorption columns because the resulting heavier-walled columns exceeded module transport capacity limitations. Also, the additional operating cost of compressing the CO_2 to high pressure before removal proved to be unattractive. While this scheme made the most efficient use of the existing compressors and minimized solvent circulation it was dropped for the reasons cited.

The second scheme was a compromise to allow efficient use of existing compressors while reducing the CO_2 removal pressure. The number of CO_2 absorbers was reduced, but was still greater than the number required at 540 psia. Additional problems were encountered in the NGL extraction section with this scheme because the hydrocarbon extraction at 1000 psia is much less selective than at lower pressures. This requires the use of two low-temperature deethanizers, resulting in higher operating and capital costs. This scheme also had many of the disadvantages of the first scheme and was dropped.

The final selection of the CO_2 removal process was based on comparative equipment requirements, fuel usage and an approximate module count.

The Sponsor Study Team concluded that, in this application, a physical solvent process is preferred to a chemical solvent process based on the following advantages and disadvantages:

Physical Solvent Advantages

- Less major equipment - potentially lower investment.
- Less process fuel - more Btu's available for delivery to the sales gas pipeline.
- Inherently better extraction of NGL - better selectivity for disposal and pipeline gas blending.
- Dry process and no water required - no additional dehydration.
- No known corrosion problems.
- No special air coolers - no freezing problems.

Physical Solvent Disadvantages

- More rotating equipment (compressors) vs. static equipment (heaters, exchangers, coolers).
- Less flexibility in achieving a fuel balance.

The major disadvantage of the physical solvents involves the desirability of recovering the hydrocarbons in the CO_2 flash gas streams which requires a much greater number of compressors. The compressors, however, are relatively high-volume, low-pressure units which are well within the proven state-of-the-art. The physical solvent regeneration flash gases have a low heating value and must be blended with higher heating value streams to meet the minimum heating value requirements of various fuel streams. The increased operating problems of fuel blending and balance and compressor maintenance were deemed preferable to operating a potentially corrosive and wet SULFINOL system requiring additional dehydration and reboiling and having potential freezing problems. The higher operating costs for the compressors in the physical solvent system were judged to be more than offset by a greater heating value to the gas pipeline.

The SELEXOL process has the following advantages over the other physical solvent processes:

- Fewer process trains - less major equipment - potentially lower investment.
- No vacuum flash in the regeneration system.
- Better NGL extraction and disposition selectivity for gas or crude blending.
- More potential for optimization.

During the process screening effort it was concluded that the most appropriate method of CO_2 disposal was in the fuel systems. It was assumed that the Prudhoe Bay Unit would accommodate this CO_2 in its fuel.

As a result of these conclusions the Sponsor Study Team recommended to the Technical Advisory Committee:

- That SELEXOL be selected as the carbon dioxide removal process.
- That mechanical refrigeration be used for hydrocarbon dewpoint control.
- That the operating pressure for both hydrocarbon dewpoint control and carbon dioxide removal be about 540 psia.

This above recommendations were accepted by the Technical Advisory Committee.

5. PROCESS DESCRIPTION

The Facilities process systems shown in block diagram form in Figure 1, and in more detail in Process Flow Diagram, Figure 2, include inlet separation and field fuel gas facilities, NGL extraction, CO_2 removal, sales gas compression and chilling, CO_2 solvent regeneration, and NGL fractionation. The process flow is similar for the four cases which were developed.

5.1 INLET SEPARATION AND FIELD FUEL GAS FACILITIES

Feed gases originating from the Gathering Centers and Flow Stations enter the Sales Gas Conditioning Facilities through the existing Central Compressor Plant inlet separators. These separators serve as liquid slug catchers and, in conjunction with downstream filter separators, serve to remove and recover any entrained liquids or particulates from the feed gases. Feed gas for the existing Field Fuel Gas Unit is withdrawn downstream of the filter separators. The Field Fuel Gas Unit feed is compressed in one of the existing first-stage injection compressors to between 1700 and 1800 psig. In the Field Fuel Gas Unit, the gas is cooled to -40°F at 850 psig by heat exchange and Joule-Thompson expansion. Cold vapor and condensed liquid are separated and the net Field Fuel Gas Unit conditioned gas is warmed by heat exchange with feed gas and goes to the Trans Alaskan Pipeline System fuel line. Cold separator liquid is also warmed by heat exchange with feed gas and is partially vaporized at about 635 psig. The separator vapor returns to the main Sales Gas Conditioning Facilities feed, and the net separator liquid joins the deethanizer feed stream.

5.2 NGL EXTRACTION

The feed stream from the inlet separators flows to the four parallel gas conditioning trains of the NGL Extraction and CO_2 Removal processes. Each train can condition 33% of the total flow, thus effectively providing one spare train. Within each of these trains the feed gas is combined with the SELEXOL stripper overhead gas and cooled to -30°F by heat exchange and propane refrigeration. Condensed liquids are separated from the cooled feed stream in the low-temperature separator and pumped through a feed gas heat exchanger where they are heated to 15°F. A partial demethanization flash occurs in the deethanizer feed flash drum, and the remaining liquid is heated to about 88°F by further exchange with feed gas and fed to the deethanizer. This 15°F flash separation serves to simplify both the design and operation of the deethanizer.

5.3 CO_2 REMOVAL

The vapor from the low-temperature separator is heated to about 20°F by exchange with feed gas and fed to the SELEXOL absorber along with deethanizer feed flash drum vapor, deethanizer overhead product gas, and SELEXOL recycle flash gas. In the absorber the feed gas is contacted countercurrently with lean SELEXOL solvent which absorbs the CO_2, a substantial portion of methane and ethane, most of the propane and essentially all of the heavier hydrocarbons from the gas. Cooling is required in the circulating solvent system to maintain the design operating temperatures. A propane refrigerated intercooler is provided on the absorber to effect this cooling. Absorber overhead gas is warmed by heat exchange with feed gas and then routed to the pipeline gas compressors.

5.4 PIPELINE GAS COMPRESSION AND CHILLING

The conditioned gas streams from the four NGL extraction/CO_2 removal trains are combined. The net propane product and most of the butanes product from fractionation can be vaporized into the combined gas stream at this point. After compression and aftercooling in the existing Central Compressor Plant equipment, the conditioned gas stream is chilled to 25°F for delivery to the gas pipeline.

5.5 CO_2 SOLVENT REGENERATION

The SELEXOL solvent system is a simple recirculating loop. Solvent rich in CO_2 first flows from the absorber through a hydraulic power recovery turbine to a recycle flash drum. In the recycle flash drum a large percentage of the methane coabsorbed with the CO_2 is vaporized and compressed back to the absorber feed. Rich SELEXOL from the recycle flash drum flows through another hydraulic turbine to an intermediate pressure (IP) flash drum. A large part of the coabsorbed ethane, as well as CO_2 vapors, are released in the intermediate pressure flash. Solvent from the intermediate pressure flash drum is routed to the atmospheric, or low-pressure flash drum where the bulk of the absorbed CO_2 and coabsorbed propane and heavier hydrocarbons are released. The low-pressure flash gases are compressed to a nominal 325 psig level and routed to the local fuel fractionator. A stripper is required to reduce the CO_2 content of the lean solvent to the level required to condition gas to the 1 percent CO_2 level. Solvent from the low-pressure flash drum is pumped to the SELEXOL stripper where it contacts a slipstream of treated gas from the absorber. The stripping gas from the absorber is depressured through two expander stages for power generation and refrigeration recovery. Stripper overhead vapor is compressed back to feed gas pressure and recycled to the feed gas NGL extraction system for recovery of stripped hydrocarbons. Stripped lean solvent is pumped from the stripper back to the absorber, thus completing the circuit.

5.6 NGL FRACTIONATION

The single-train fractionation facilities consist of the local fuel fractionator, deethanizer, depropanizer, and debutanizer. All of these columns are reboiled by direct fired heaters. Compressed SELEXOL low-pressure flash gas is fed to the local fuel fractionator to recover the bulk of the propane and heavier hydrocarbons from the gas. The column has a refrigerated overhead condenser and is similar to a deethanizer. Separate feed-overhead heat exchangers are used for the local turbine and heater fuel portions of the overhead product. Propane is added to the turbine fuel portion of the overhead product for enrichment. This propane is vaporized in the feed-overhead exchanger. Local fuel fractionator bottoms product is fed to the depropanizer.

The deethanizer feed is made up of deethanizer feed flash liquids and NGL from the Field Fuel Gas Unit. The deethanizer operates at a nominal 450 psig with a propane-refrigerated condenser. Deethanizer overhead vapor product is compressed and can go either to field fuel or to SELEXOL absorber feed. Deethanizer bottoms product is fed to the depropanizer along with local fuel fractionator bottoms. The depropanizer produces a liquid propane overhead product stream. The low propane content depropanizer bottoms product can be blended directly into crude oil or fed to the debutanizer. The debutanizer produces a liquid butanes overhead product and a pentanes-plus bottoms product. The debutanizer overhead product can be almost totally injected into the pipeline gas without exceeding the pipeline gas hydrocarbon dewpoint specification or can be blended into the crude oil up to true vapor pressure limitations. The process design included the cooling facilities required for NGL and crude blending. These facilities are excluded from the summary cost estimate but are included in the cost estimate details in Section 11 of this volume.

A system is provided to inject the liquid feed for any column in the fractionation facilities alternatively into the producing formation. Therefore, an upset or equipment failure in the unspared fractionation facilities will not impair either crude oil production or pipeline gas deliveries.

A small sidestream rectifier is utilized on the depropanizer to provide refrigerant grade propane as makeup for the refrigeration system. This column draws a small ethane-free vapor feed from below the depropanizer feed tray and produces a very pure propane overhead product. The bottoms are pumped back to the depropanizer.

5.7 FUEL SYSTEMS

The intermediate-pressure flash gas is collected from the CO_2 removal trains and compressed to a nominal 500 psig for use in the field fuel. Compressor discharge heat is used to vaporize propane. The propane is injected into this stream for heating value control. Field fuel requirements greater than those available from this flash gas stream are met by adding Field Fuel Gas Unit conditioned gas (in excess of Trans Alaskan Pipeline System requirements), deethanizer overhead vapor, and low temperature separator vapor in that order. The combined field fuel gas has a relatively high hydrocarbon dewpoint. This gas is heated to 140°F by exchange with the exhaust gas from the field fuel gas compressor turbine driver to prevent condensation in the insulated field fuel distribution system.

In situations where the field fuel requirement is relatively low, there may be an excess of SELEXOL intermediate-pressure flash gas. At these times, excess field fuel compressor discharge will be bled into the local turbine fuel system. This, in turn, will create an excess of local fuel fractionator overhead vapor. During this operation, the excess CO_2-rich local fuel fractionator overhead can be compressed and reinjected into the producing formation. If the local fuel fractionator is shut down, the feed to this column can be used for local fuel, and the excess feed can be injected using both CO_2 compressors. Also, during periods of high local fuel demand, field fuel compressor discharge can be used to supplement local fuel fractionator overhead.

6. PLANT YIELDS

In addition to the nominal 2,000 MMSCFD of pipeline gas conditioned by the Facilities there are a number of other streams that are separated incidental to the pipeline gas conditioning. These include the high CO_2 content flash gases from the regeneration of the SELEXOL® solvent and the NGL. The flash gases are utilized as fuel in the Facilities and to supply the fuel requirements of the Prudhoe Bay Unit. The NGL, which includes separate propane, butane, and pentanes-plus streams, may be blended into either the fuel streams (propane) to control heating value, the pipeline gas to the hydrocarbon dewpoint limitation (propane or butane) or into the crude (butane or pentanes-plus) as limited by the vapor pressure specification.

Figure 8 is a block flow diagram of plant stream sources and disposition and shows one possible blending scheme for the various products. The design anticipates that there will be a significant variation in the fuel requirements of the Facilities between the extremes of summer and winter operation. The demand for fuel by the Prudhoe Bay Unit will vary both as a function of season and time as well as oil production rates. The blending of butanes into either pipeline gas or crude is controlled by the pipeline hydrocarbon dewpoint limitation or by economics. These variations have been allowed for in the design of the Facilities. Thus an almost infinite number of blending schemes is possible permitting effective and economic utilization of all streams from the Facilities. The scheme illustrated in Figure 8 represents the maximum anticipated demand for fuel by the Prudhoe Bay Unit (Field Fuel) and assumes no blending of butanes to the pipeline gas; in actuality about 97.8% of the butanes can be blended into the pipeline gas without exceeding the assumed hydrocarbon dewpoint specification. (See Volume II for butanes blended into 1260 pipeline case.)

The gas requirements of the Trans Alaskan Pipeline System will normally be provided by gas conditioning in the existing Field Fuel Gas Unit. The flash liquid and vapor streams resulting from this operation will be processed in the Sales Gas Conditioning Facilities.

Tables 6-1, 6-2, 6-3, and 6-4 show the quantities, properties and compositions for the various feed and product streams for the same case as Figure 8. In addition, Tables 6-1 and 6-3 show the same information for the maximum butanes blended into pipeline gas. Depending upon the particular case there will be some variation in the quantities, properties and compositions of all streams. For design purposes material balances have been developed only for the minimum and maximum Prudhoe Bay Unit fuel gas demands.

Table 6-1. Feed From Gathering Centers and Flow Stations

6-1A. SOURCES AND COMPOSITIONS	From Gathering Centers	From Flow Stations	Total
Feed Gas (MMSCFD)(1)	1400.0	1400.0	2800.0
Compositions - Volume % as Gas			
H_2S (ppmv)	8 ppmv	8 ppmv	8 ppmv
CO_2	12.72	12.54	12.63
N_2	0.45	0.49	0.47
Methane	74.35	74.00	74.17
Ethane	6.33	6.61	6.47
Propane	3.43	3.52	3.48
Butanes	1.60	1.72	1.66
Pentanes-plus	1.12	1.12	1.12
	100.00	100.00	100.00

6-1B. OVERALL MATERIAL BALANCE

Stream	Volume (MMSCFD)(1)
Feed from Gathering Centers	1400.0
Feed from Flow Stations	1400.0
Total Feed	2800.0
Sales Gas-Net	2000.6
Propane-Balance over fuels	37.2
Butanes	41.3
Pentanes-plus	31.9
Fuels	689.0
Total Product	2800.0

(1) Standard Conditions - 14.65 psia and 60°F.

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Table 6-2. Fipeline Gas

	As Conditioned	Propane Blended (6)	Propane Plus Butane Blended (7)
Volume Produced - MMSCFD (1)	2000.6	2037.8	2078.2
MMSCFD (2)	1989.7	2026.7	2067.0
CO_2 content - vol.% (3)	0.50	0.49	0.49
Hydrocarbon Dewpoint @ 1100 psia-°F	-60 approximate	-50 approximate	-10.
Heating value			
Higher heating value - Btu/SCF (4)	1034	1061	1106
- Btu/SCF (5)	1022	1048	1092
Lower heating value - Btu/SCF (8)	930	955	994
H_2S content Grains/100 SCF	<1.0	<1.0	<1.0
Water Dewpoint @ 1100 psia -°F	-35°	-35°	-35°
Composition Volume % as Gas			
H_2S	1 ppmv	2 ppmv	2 ppmv
CO_2 (3)	0.50	0.49	0.49
N_2	0.62	0.61	0.60
Methane	94.29	92.57	90.77
Ethane	4.47	4.50	4.41
Propane	0.08	1.75	1.72
Butanes	0.03	0.07	1.95
Pentanes	0.01	0.01	0.06
	100.00	100.00	100.00

(1) Standard conditions - 14.65 psia and 60°F (State of Alaska Base).
(2) Standard condition - 14.73 psia and 60°F (FERC Base).
(3) SGCF was designed at 0.5 vol. % CO_2 to guarantee 1.0 vol. % max. in Pipeline Gas.
(4) Gross, dry, actual at 14.65 psia and 60°F.
(5) Gross, wet, actual at 14.73 psia and 60°F.
(6) This is the as-conditioned gas when blended with propane in excess of fuel requirements.
(7) This is the as-conditioned gas when blended with propane in excess of fuel requirements plus butanes to the hydrocarbon dewpoint limitation.
(8) Net, dry, ideal at 14.65 psia and 60°F.

Table 6-3. Fractionator Products

	Propane	Butanes		Pentanes-plus
Net Extracted				
BPD	52,350	31,290		28,700
MMSCFD (1)	80.3	41.3		--
Disposition		Min.	Max.	
To Pipeline Gas				
BPD	24,270	-0-	30,602	-0-
MMSCFD (1)	37.2	-0-	40.4	-0-
To Fuel Gas Systems				
BPD	28,080	-0-	-0-	-0-
MMSCFD (1)	43.1	-0-	-0-	-0-
To Crude				
BPD	-0-	688	31,290	28,700
Composition - Volume % as gas				
H_2S	38 ppmv	-		-
CO_2	0.01	-		-
Ethane	6.40	-		-
Propane	91.59	0.17		-
Butanes	2.00	97.31		6.01
Pentanes-plus	-	2.52		93.99
	100.00	100.00		100.00

(1) Standard Conditions - 14.65 psia and 60°F.

Table 6-4. Fuel Streams

	For TAPS	For Prudhoe Bay Unit (Field Fuel)	For Local Turbine	For Local Heater
MMSCFD (1)	40.9	399.4	192.0	56.7
Lower Heating Value-Btu/SCF	851	825	475	212
MM Btu/hr.	1,450	13,729	3,800	500
Hydrocarbon Dew Point -°F (2)	-40°F @ 840 psig	27°F @ 500 psig	17°F @ 300 psig	60°F @ 60 psig
Composition - Volume % as gas				
H_2S	4 ppmv	22 ppmv	38 ppmv	40 ppmv
CO_2	11.44	36.40	75.08	87.21
N_2	0.58	0.13	-	-
Methane	81.54	37.73	2.71	1.46
Ethane	4.70	16.24	9.23	9.04
Propane	1.37	8.84	12.67	2.29
Butanes	0.31	0.60	0.30	-
Pentanes-plus	0.06	0.06	0.01	-
	100.00	100.00	100.00	100.00

(1) Standard conditions - 14.65 psia and 60°F.
(2) Pressures are system-operating pressure.

7. PROCESS AUXILIARY SYSTEMS

Utility systems for the Sales Gas Conditioning Facilities were developed to the detail necessary for estimating cost. They include:

- Electrical power generation system.
- Glycol cooling medium system.
- Glycol process heating medium.
- Glycol module heating medium system.
- Nitrogen inert gas system.
- Service and instrument air systems.
- Closed hydrocarbon drainage system.
- Potable and utility water system.
- Sanitary sewer system.
- Hydraulic power center systems.
- Communication systems.
- Safety and fire protection systems.

The peak electrical power requirement for the Facilities is about 65 megawatts. Power recovery generators on the stripping gas expanders and solvent streams provide 12 megawatts of this power. The balance is generated by a three-generator unit utility power plant. An on-line spare generator unit is provided during all load conditions. Lower loads during cold ambient temperature conditions permit shutting down a unit for planned maintenance without affecting the minimum design spare generating capacity. Emergency power for critical control, lighting, fire protection and heating systems is provided by smaller turbine generator units.

8. FACILITY LOCATION AND ARRANGEMENT

Because this Study assumed use of the existing first-stage injection compressors as a base case, the Sales Gas Conditioning Facilities were located close to the Central Compressor Plant. This location takes advantage of the existing field gas gathering lines to the Central Compression Plant and the existing gas injection wells near the Central Compression Plant. Its proximity to the existing docks reduces time required for low-speed transporter movement of the large number of modules from the docks to the site. A site location plan is shown as Figure 3.

Modular construction is a proven, cost effective method for constructing facilities at Prudhoe Bay. Modular design utilizes a structural steel base supporting a steel frame building housing the equipment. The module size and base structure are designed to allow movement by low-speed transporters. Modules are assembled to the maximum extent possible at the "Lower 48" fabrication site to avoid the significantly higher North Slope costs. The modules are then barged to the North Slope and set on prepared gravel and piling sites. The modules are interconnected after placement, and utilidors (enclosed pipeways/walkways) are used where required to provide the necessary safe module separation.

The layout of the modules included in the facilities is centered around the four gas conditioning trains with space reserved for up to a 50% capacity expansion. Process and support facilities are centrally located near the gas conditioning trains. This was done, without compromising general safety separation criteria, to service the gas conditioning trains as efficiently as possible. The emergency flares are located a safe distance from the process facilities. The crude cooling and NGL crude blending facilities are located near the Trans Alaskan Pipeline System Pump Station No. 1. The Construction Camp and Operations Center are located a convenient but safe distance from the process facilities and the pipeline routes. Overall module layouts of the process facilities and the camps are shown on Figure 4.

Figure 7 illustrates a typical layout drawing for a 33% capacity gas conditioning train, which includes nine modules and one utilidor.

9. MODULE FIRE AND SAFETY CONSIDERATIONS

Totally enclosed modularized facilities are ideal for providing controlled environmental conditions during construction and operation on the North Slope. However, modular construction requires special ventilation and fire protection measures. The modules are maintained within a controlled temperature range using heated makeup air. Exhaust fans insure proper ventilation and prevent an accumulation of hydrocarbons. Other protective measures and safety systems considered in this study include:

- Emergency Shutdown System for process isolation of all or portions of the Facilities.
- Process area isolation to prevent the spread of hydrocarbon vapors or fire.
- Gas Detection and Fire Sensing Systems.
- Halon inerting and fire extinguishing systems.
- Firewater backup extinguishing systems.
- Liquid hydrocarbon and firewater drainage systems.
- Relief, vent and flare systems.
- Mobile fire fighting equipment

10. IMPLEMENTATION PLAN

The implementation plan provides a sequenced checklist of scheduled events for the entire project from initiation of detailed design engineering through startup of the Facilities. The plan provides the basis for estimating the cost of engineering, procurement, fabrication, logistics and erection activities. The plan assumes early and effective use of proven scheduling, quality assurance, cost control, and material control procedures to assure completion of the Project on time and within the authorized funding.

10.1 SCHEDULE

The implementation plan schedule milestones illustrated in Figure 5 show an exceptionally tight sequence of activities from initiation of final process design to startup. With careful planning, timely funding, and absence of regulatory or legal delays it would be possible to complete the Facilities with a 4-1/2 year lead time.

In order to have a 66% capacity startup in January, 1983, it is necessary to initiate final process design in September, 1978. Full capacity startup is not possible before July, 1983. In order to achieve either schedule, it is necessary to conclude detail design engineering contractor selection no later than January, 1979. Final process design must be completed by December, 1978, and detailed mechanical design must commence no later than January 1, 1979. Time sensitive activities during the first three months of detailed design include:

- Preparation of specifications and requisitions for critical delivery equipment and materials such as turbines, compressors, heavy-wall vessels, and low-temperature piping.

- Design of the first phase of the Construction Camp and its associated facilities to support a spring 1980 installation at Prudhoe Bay.

- Initiation of permitting activities including the gathering of baseline air quality data required for construction permits in support of 1980 and 1981 Prudhoe Bay erection activities.

10.2 FUNDING

Required funding approvals are shown on the schedule, phased as required to support critical design and procurement activities. To meet the schedule, funding for these activities should be provided by late 1978 and preferably prior to engineering contract negotiations. Total project funding is required by December, 1979, which coincides with award date for fabrication contracts. Prior to December, 1979, total project funding approval, commitments totaling approximately $500,000,000 will be required for engineering and ordering critical equipment.

10.3 PROCUREMENT

Given present market conditions and recognizing possible impact of concurrent projects such as the Alaska Highway Gas Pipeline System and the Prudhoe Bay Unit production facility work, procurement of long lead items included in the 1981 sealift should begin in March, 1979.

10.4 CONSTRUCTION

The general construction plan assumes three phases of work involving a small sealift in 1980 supplemented by truck hauling followed by two major sealifts in 1981 and 1982. Pre-sealift work must be initiated in 1980 both at "Lower 48" fabrication sites and at Prudhoe Bay.

The implementation plan assumes an optimized North Slope labor force of 1,000. This manpower level dictates the Construction Camp size and associated support costs for Prudhoe Bay. It also indirectly dictates the fabrication manpower level. Scheduled and peak manpower requirements are shown in Table 10-1.

Table 10-1. Manpower Requirements

	"Lower 48" Fabrication		North Slope Erection	
BASE CASE	Total Manhours	Peak Manpower (Year)	Total Manhours	Peak Manpower (Year)
Full Startup	9,366,000	3,000 (81,82)	5,435,000	800 (82)
Phased Startup	9,387,000	4,500 (81)	5,389,000	1,000 (81,82)
ALTERNATE CASE				
Full Startup	10,141,000	3,500 (82)	5,720,000	800 (82)
Phased Startup	10,134,000	4,750 (81)	5,656,000	1,000 (81,82)

The fabrication sites are assumed to be located adjacent to major deep-draft waterways which will accommodate ocean-going barges. West Coast fabrication sites were assumed for planning purposes. The West Coast provides favorable weather conditions, adequate manpower and the shortest shipping distance. Fabrication site size requirements are estimated at 200 acres for 1981 and 80 acres for 1982. The job size and construction manpower requirements indicate that at least two sites must be considered as a planning basis. The estimate basis assumes that sites are completely unimproved and require grading, compacting and construction of module pads, offices, shops, warehouses, utility distribution systems, fencing, and barge loading facilities.

Support facilities at Prudhoe Bay include a temporary Construction Camp and a permanent Operations Center along with required power generation, water supply, and sewage and water treating facilities. The Construction Camp would be erected in two phases. The first phase, erected during early 1980, would provide lodging and support for the approximately 450 staff and craft personnel involved in gravel, piling, and other early construction work from June, 1980, to June, 1981.

Since the peak period of North Slope construction is scheduled for the fall of 1981, the second phase of camp construction would be completed during the spring of 1981. This expansion increases the size of the construction camp to accommodate the projected peak construction manpower of 1,000.

The completed Construction Camp will include lodging, kitchen, dining room, recreational facility, laundry, medical room, and office space. Utility buildings will include garages, warehouses, miscellaneous flammable liquid storage, bulk gasoline and diesel fuel storage and dispensing terminal, welding shop and fabrication shop. Facilities are provided for emergency power generation, mobile fire protection, waste incineration, water treatment, and sewage treatment. Some of these will be permanent facilities associated with the Operations Center.

The Operations Center erected in 1981 will be a 200-man modular facility designed to house and support permanent operations, maintenance, and staff personnel. During construction, the Operations Center would be used to house Plant Owner/Operator representatives, operators, and startup and maintenance personnel assigned to monitor and inspect construction activities. Included in initially installed facilities are utilities which will be used to support both the Construction Camp and the Operations Center. These will be trucked overland in 1980. Operations Center units delivered in the 1981 sealift will include administrative offices, kitchen, medical room, laundry, recreational facility, and support systems similar to those erected earlier for the Construction Camp.

10.5 LOGISTICS

Docking facilities at fabrication sites should be capable of dead loading modules; i.e., loading a barge that is flooded and bottomed. A general cargo storage and staging area should be provided with appropriate lighting and support equipment. Both 1981 and 1982 module loadout programs will require approximately 30 days each to complete preceded by approximately two weeks of module preparation.

Barge requirements to move the completed modules and general cargo are estimated as follows:

	PHASED STARTUP			FULL STARTUP		
	1980	1981	1982	1980	1981	1982
BARGES REQUIRED	2	25	11	2	24	16

The two startup cases require approximately the same square footage of barge space. In the phased startup, larger barges but fewer are required.

Two possibilities were considered for module offloading at Prudhoe Bay as follows:

(1) A new separate causeway and dock for the Facilities could be constructed independent of existing North Slope facilities. Offloading of modules would then be independent of the other projected Prudhoe Bay Unit projects.

(2) The existing dock facilities owned by SOHIO/ARCo./ EXXON could be expanded and the offloadng coordinated with concurrent North Slope projects.

The latter approach is assumed for our planning basis. This approach results in less capital expenditure and tends to minimize environmental impact. The use of these facilities would require close coordination between the Sales Gas Conditioning Facilities project and other concurrent North Slope projects to ensure timely offloading of all modules and general cargo. Module offloading, estimated to take 21 to 34 days, can be accomplished through the use of both crawler transporters and newly developed rubber-tired vehicles.

10.6 STARTUP

Facilities startup activities will begin approximately six months prior to mechanical completion. It is assumed that an Operator/Owner startup team will be formed and used in a technical and supervisory capacity with regular operating personnel performing the physical startup of the facility.

11. ESTIMATED COSTS

The Sales Gas Conditioning Facilities cost estimates were prepared on a mid-1978 cost basis. Escalation was estimated on an item by item basis and is extremely sensitive to implementation schedule.

11.1 ESTIMATING METHOD

The estimating method utilized the modular construction cost experience of Atlantic Richfield Company (Unit Operator - East) and The Ralph M. Parsons Company, from production facilities development on the east side of the Prudhoe Bay Field.

The estimate was developed from:

- Individual Module Layouts to Determine Size and Weight.
- Major Equipment Estimating Quotations.
- Major Piping Layout.
- Single-Line Electrical Drawings.
- Detailed Analysis of "Lower 48" and North Slope Labor Requirements.
- Logistics Analysis.

The estimate basis is presented in detail in Volume III of this Report.

11.2 ESTIMATED COSTS

Estimated costs, including contingency and excluding the Crude Oil Cooling Facilities, for each of the four cases studied are tabulated below:

	Mid-1978 Basis (Dollars)	Escalated Basis (Dollars)
Base Case Full Startup	1,571,500,000	1,962,300,000
Base Case Phased Startup	1,581,200,000	1,964,500,000
Alternate Case Full Startup	1,688,400,000	2,109,700,000
Alternate Case Phased Startup	1,701,700,000	2,107,200,000

A Unit detail of these costs is shown in Table 11-1.

The escalation applied to each of the four cases is based upon a Ralph M. Parsons Company composite analysis of various inflation forecast sources listed in the estimate basis. (Detailed in Volume III Section 12.) The escalation percentages were then applied to the costs in accordance with the implementation schedule. Based on the current plan and schedule, a weighted average annual escalation of 6.73% was applicable.

11.3 CONTINGENCY ANALYSIS

Contingency funds were included in the costs presented in this report; the contingency percentage used was determined by risk analysis using RISKAN.® Each major cost account was evaluated for probable accuracy variation taking into account the engineering design basis, extent and accuracy of material takeoffs and estimating and pricing methods. Where escalation was included in the costs, the escalation was considered in the risk analysis. The details of the risk analysis are shown in Volume III of this report.

The contingency selected gives a 90 percent probability that the estimated cost including contingency will not be exceeded, provided the design and schedule are followed as planned in this study.

11.4 OTHER FACILITY COSTS

Crude oil cooling facilities costs are excluded from the total cost of the Facilities. These costs are shown separately in Table 11-2.

11.5 DELAY COST TREND

Delays in proceeding with this project will increase the cost of the facilities in accordance with escalation forecasts. Delay cost trends calculated for delay of completion year to 1984, 1985, and 1986 are as follows when utilizing the same 4-1/2 year implementation plan:

	Incremental Delay Cost	
Full Capacity Delivery Year	Base Case (Dollars)	Alternate Case (Dollars)
1984	$160,000,000	$170,000,000
1985	$350,000,000	$370,000,000
1986	$540,000,000	$580,000,000

Delay costs are calculated by an application of escalation to equipment and materials, labor and support, engineering design and management and plant operator's costs, then with recalculation of contingency using RISKAN procedure. This procedure indicates a delay cost of approximately 9 percent per year. These costs are not included in the project estimated costs presented in subsection 11.2.

11.6 COSTS INCLUDED IN THE ESTIMATES

The following costs have been included in the estimates. The estimate detail (Volumes VI and VII) provides information on these costs:

- Engineering, Procurement

- Equipment and Materials
- Module Fabrication
- North Slope Erection
- Construction Management and Supervision
- Fabrication Site Lease, Site Development and Temporary Facilities
- North Slope Camp and Other Temporary Facilities
- Barging Costs
- Operations Center
- SELEXOL® License Fee
- Initial Charge of Chemicals and Operating Supplies.
- Spare Parts
- Dual fuel system and combustor retrofit in Central Compression Plant
- Costs Paid Directly by the Plant Operator/Owner (Plant Operator Cost):
  - Preliminary Engineering Design Study *
  - Plant Operator Engineering and Construction Supervision
  - Plant Operator Commissioning and Startup Costs *
  - Alaska ad Valorem and Borough Taxes *
  - Dock Use Charges

11.7 COSTS EXCLUDED

Costs excluded from the estimate are:

- Financing Costs
- Reimbursement for cost of Prudhoe Bay Unit Central Compression Plant

*Normally expensed by Operator/Owner

- Operating Costs
- Operating Supplies (SELEXOL, Glycol, Lube Oils), except Initial Charge Included in Estimate
- Sales Tax on In-Transit Materials
- Cost for Lost or Damaged Materials (or Insurance for)
- All-Risk Insurance
- Land/Mineral Leases at Prudhoe Bay
- Permit and Environmental Impact Statement Costs
- Unexpected Deviation from, Adverse Interperatation of, or Revision to Current Environmental Regulations
- Landing Strip or Airport Facilities at Prudhoe
- Dock Expansion Costs (Dock Use Charge Included)
- Use of Prudhoe Bay Unit Roads and the haul road from Fairbanks to Prudhoe Bay
- Facility Removal Costs at End of Production Life
- Revision to Schedule or Major Delays
- Salvage, Scrap or Transfer Value of Surplus or Cancelled Equipment/Material
- Process Optimization Engineering Cost
- Contingency for significant scope changes

11.8 CASH FLOW

Projected cash flow including contingency and escalation, but exclusive of Crude Oil Cooling Facilities (Unit 62), for the four cases included in this study are tabulated in Table 11-3. The base case, full startup commitment and expenditure curves are shown on Figure 9. The commitment curve reflects material commitments when ordered whereas construction costs are committed as spent.

TABLE 11-1. ESTIMATE SUMMARY
TOTAL DOLLARS IN MILLIONS

UNIT No.	DESCRIPTION	MID-1978 DOLLARS				ESCALATED DOLLARS			
		BASE CASE		ALTERNATE CASE		BASE CASE		ALTERNATE CASE	
		FULL STARTUP	PHASED STARTUP	FULL STARTUP	PHASED STARTUP	FULL STARTUP	PHASED STARTUP	FULL STARTUP	PHASED STARTUP
31-34	NGL Extr. & CO_2 - Train #1-4	177.6	177.6	177.6	177.6	177.6	177.6	177.6	177.6
41	Sales Gas Compression	6.7	6.7	-	-	6.7	6.7	-	-
42	Inlet Separation	-	-	7.8	7.8	-	-	7.8	7.8
43	Sales Gas Compression	-	-	59.2	59.2	-	-	59.2	59.2
45	Refrigeration	81.1	81.1	81.1	81.1	81.1	81.1	81.1	81.1
46	Field Fuel & CO_2 Fac.	28.3	28.3	28.3	28.3	28.3	28.3	28.3	28.3
51	NGL Fractionation	51.3	51.3	51.3	51.3	51.3	51.3	51.3	51.3
61	Utilities/Pipeways	140.5	140.5	141.8	141.8	140.5	140.5	141.8	141.8
63	Power Generation	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2
80	Operations Center	22.2	22.2	22.2	22.2	22.2	22.2	22.2	22.2
	Total Equipment & Material	527.9	527.9	589.5	589.5	527.9	527.9	589.5	589.5
	Lower 48 Labor & Support	256.1	259.4	272.7	276.7	256.1	259.4	272.7	276.7
	North Slope Labor & Support	401.7	403.2	416.5	419.2	401.7	403.2	416.5	419.2
	Engineering, Design & Mgmt.	91.0	91.0	98.0	98.0	91.0	91.0	98.0	98.0
	Plant Operator's Costs:								
	Owner/Operator Engr. Teams	44.9	45.0	47.8	48.0	44.9	45.0	47.8	48.0
	Dock Use Charge	7.6	7.6	8.0	8.0	7.6	7.6	8.0	8.0
	Commissioning & Start-up	49.0	49.0	49.1	49.1	49.0	49.0	49.1	49.1
	Ad Valorem & Borough Taxes	63.5	67.5	67.4	72.7	63.5	67.5	67.4	72.7
	Mid-1978 Subtotal	1441.7	1450.6	1549.0	1561.2	1441.7	1450.6	1549.0	1561.2
	Escalation	-	-	-	-	294.9	287.9	318.0	303.6
	Contingency (@9% '78 Dollars / @13% Esc.Dollars)	129.8	130.6	139.4	140.5	225.7	226.0	242.7	242.4
	Total Facility Estimate	1571.5	1581.2	1688.4	1701.7	1962.3	1964.5	2109.7	2107.2

NOTE: This estimate summary excludes Unit 62, Crude Oil Cooling Facilities.

TABLE 11-2
CRUDE OIL COOLING FACILITIES - UNIT 62
ESTIMATE SUMMARY
TOTAL DOLLARS IN MILLIONS
(ALLOCATED FROM TOTAL ESTIMATE)

UNIT No.	DESCRIPTION	MID-1978 DOLLARS				ESCALATED DOLLARS			
		BASE CASE		ALTERNATE CASE		BASE CASE		ALTERNATE CASE	
		FULL STARTUP	PHASED STARTUP	FULL STARTUP	PHASED STARTUP	FULL STARTUP	PHASED STARTUP	FULL STARTUP	PHASED STARTUP
62	Crude Oil Cooling Facilities	22.4	22.4	22.4	22.4	22.4	22.4	22.4	22.4
	Gravel	.3	.3	.3	.3	.3	.3	.3	.3
	Total Equipment & Material	22.7	22.7	22.7	22.7	22.7	22.7	22.7	22.7
	Lower 48 Labor & Support	13.0	13.2	12.9	13.1	13.0	13.2	12.9	13.1
	North Slope Labor & Support	10.2	10.2	9.9	10.0	10.2	10.2	9.9	10.0
	Engineering, Design & Mgmt.	3.5	3.5	3.5	3.5	3.5	3.5	3.5	3.5
	Plant Operator's Costs								
	Owner/Operator Engr. Teams	1.7	1.7	1.7	1.7	1.7	1.7	1.7	1.7
	Dock Use Charge	.3	.3	.3	.3	.3	.3	.3	.3
	Commissioning & Start-up	1.9	1.9	1.7	1.7	1.9	1.9	1.7	1.7
	Ad Valorem & Borough Taxes	2.5	2.6	2.4	2.6	2.5	2.6	2.4	2.6
	Mid-1978 Subtotal	55.8	56.1	55.1	55.6	55.8	56.1	55.1	55.6
	Escalation	-	-	-	-	13.1	11.2	12.9	11.1
	Contingency (@9% '78 Dollars / @13% Esc.Dollars)	5.0	5.0	5.0	5.0	9.0	8.7	8.8	8.7
	Total Unit 62 Estimate	60.8	61.1	60.1	60.6	77.9	76.0	76.8	75.4

RMP

Table 11-3. Cumulative Cash Flow
(Millions of Dollars)

Year	Quarter	Base Case Phased Startup	Base Case Full Startup	Alternate Case Phased Startup	Alternate Case Full Startup
1978	1-4	1.7	1.7	1.7	1.7
1979	1	8.0	8.0	8.0	8.0
	2	51.0	35.0	57.0	48.0
	3	100.0	60.0	117.0	87.0
	4	200.0	122.0	223.0	163.0
1980	1	288.0	189.0	319.0	231.0
	2	431.0	309.0	484.0	344.0
	3	584.0	457.0	658.0	487.0
	4	733.0	626.0	834.0	654.0
1981	1	887.0	798.0	1,015.0	852.0
	2	1,053.0	984.0	1,211.0	1,066.0
	3	1,172.0	1,127.0	1,321.0	1,211.0
	4	1,291.0	1,271.0	1,432.0	1,355.0
1982	1	1,396.0	1,434.0	1,606.0	1,537.0
	2	1,492.0	1,584.0	1,767.0	1,705.0
	3	1,626.0	1,635.0	1,824.0	1,762.0
	4	1,741.0	1,678.0	1,873.0	1,810.0
1983	1	1,821.0	1,795.0	1,958.0	1,935.0
	2	1,875.0	1,874.0	2,015.0	2,018.0
	3	1,938.0	1,936.0	2,080.0	2,082.0
	4	1,964.5	1,962.3	2.107.2	2,109.7

12. FUTURE DESIGN CONSIDERATIONS

A number of ideas were developed during the progress of the Study which, if determined to be viable and incorporated into the final design, could potentially result in significant capital savings and/or reduction in operating cost. It was not possible to evaluate all of these ideas during the Study because of schedule limitations.

Although some of these ideas apparently have obvious cost benefits and need only to be developed, others require economic evaluation and justification before they could be implemented. More detailed information on Facility ownership, value of conditioned gas, value of fuel gas, value of NGLs, transportation tariff, etc., must be known before these studies can be concluded. Some of these ideas may be evaluated concurrently with the development of the final design. These ideas are put forth in Volume IV.

Other of these ideas will impact the schedule and should be the first item of work if the Facilities are to be completed by July, 1983. The decision as to whether or not these ideas are to be incorporated in the final design should be made before starting the final design because:

- Incorporation will influence the scope of work for the final design.
- Time is required for the development work necessary for their application.

These ideas, which impact schedule, are outlined below and developed in detail in Volume II.

12.1 BASIC SELEXOL PROCESS DESIGN

The basic NGL/CO_2 extraction processes can potentially be modified by increasing the operating temperature of the low-temperature separator, altering the feed gas chilling and altering SELEXOL solvent chilling systems. The above changes would reduce both capital and operating costs.

12.2 REFRIGERATION SYSTEM OPTIMIZATION

Several possible configurations for the refrigeration facilities were examined in the Study. A centralized system with parallel compressors was selected. Further optimization of this system is possible. There are a number of areas where either the refrigeration level or load can be reduced. The impact of these improvements would be to reduce both the capital and operating costs of the refrigeration system. During the selection of the final configuration, it is suggested that a detailed analysis of the refrigeration process control system be conducted.

12.3 WASTE HEAT RECOVERY

Potential exists for further fuel conservation via additional waste heat recovery from the gas turbine exhaust gas. This potential is over and above the heat recovery by using regenerative gas turbines in the design. The potential uses for this heat include:

- Reboiling of the fractionator towers
- Module heating

The impact of using waste heat would be to increase available gas for future disposition. The evaluation should consider the additional requirements for disposal of surplus low heating value gas during early and late years of the Facilities and Field operation.

12.4 REEVALUATION OF THERMODYNAMIC DATA

The proper selection and sizing of process equipment for the Facilities is largely dependent on thermodynamic data. For the purposes of this study the best available data were used. The quality of this data and hence the quality of the final design may be improved by:

- A comprehensive literature search and data evaluation covering the expected composition and operating conditions for this facility.
- Generation of new data for these facilities in the laboratory, where suitable data are not in the literature.

12.5 PIPELINE GAS SPECIFICATIONS

The gas pipeline specifications and delivery pressure adopted for this Study were established on the basis of the best available knowledge at the time the Study was started. A higher allowable CO_2 content of the delivered gas will have a significant impact on the facilities costs. It is recommended that the cost impact of the possible alternative specification for CO_2 content and gas pipeline pressure be evaluated for the overall conditioning and gas transmission system.

12.6 GAS TURBINE COMBUSTORS

It was determined that gas turbines with the proper combustor and control configuration can burn both low-Btu/high-CO_2 content fuel gas and normal-Btu fuel gas. Combustor testing and development will be required for the specific fuels available from these facilities. This work by the selected vendor or vendors should begin in the first quarter of 1979 to insure timely delivery of the gas turbines. In addition, data from these tests will be required to obtain permits.

12.7 SELEXOL ABSORBER/STRIPPER INTERNALS

There are several vendors of tower packings who claim very high hydraulic capacity for some of their packings. If these claims are valid it would be possible to eliminate one SELEXOL train. However, these claims should be validated by independent testing prior to use of these packings.

12.8 ENVIRONMENTAL

Administration by local agencies of the Clean Air Act and Amendments of 1977 could impose requirements not perceived by this Study. Upon project initiation, the requirements should be determined as soon as possible and the impact on Facilities design and implementation should be established.

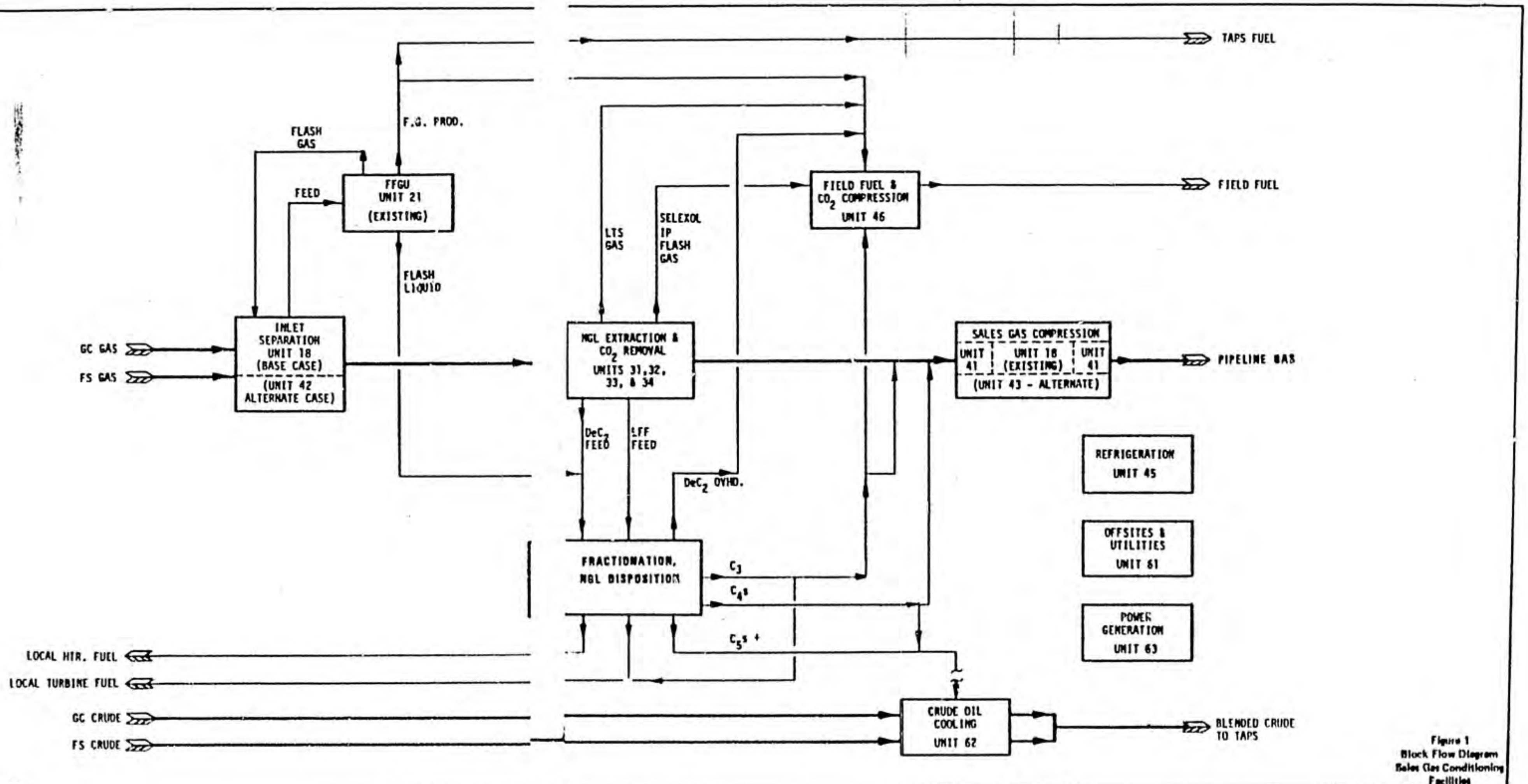
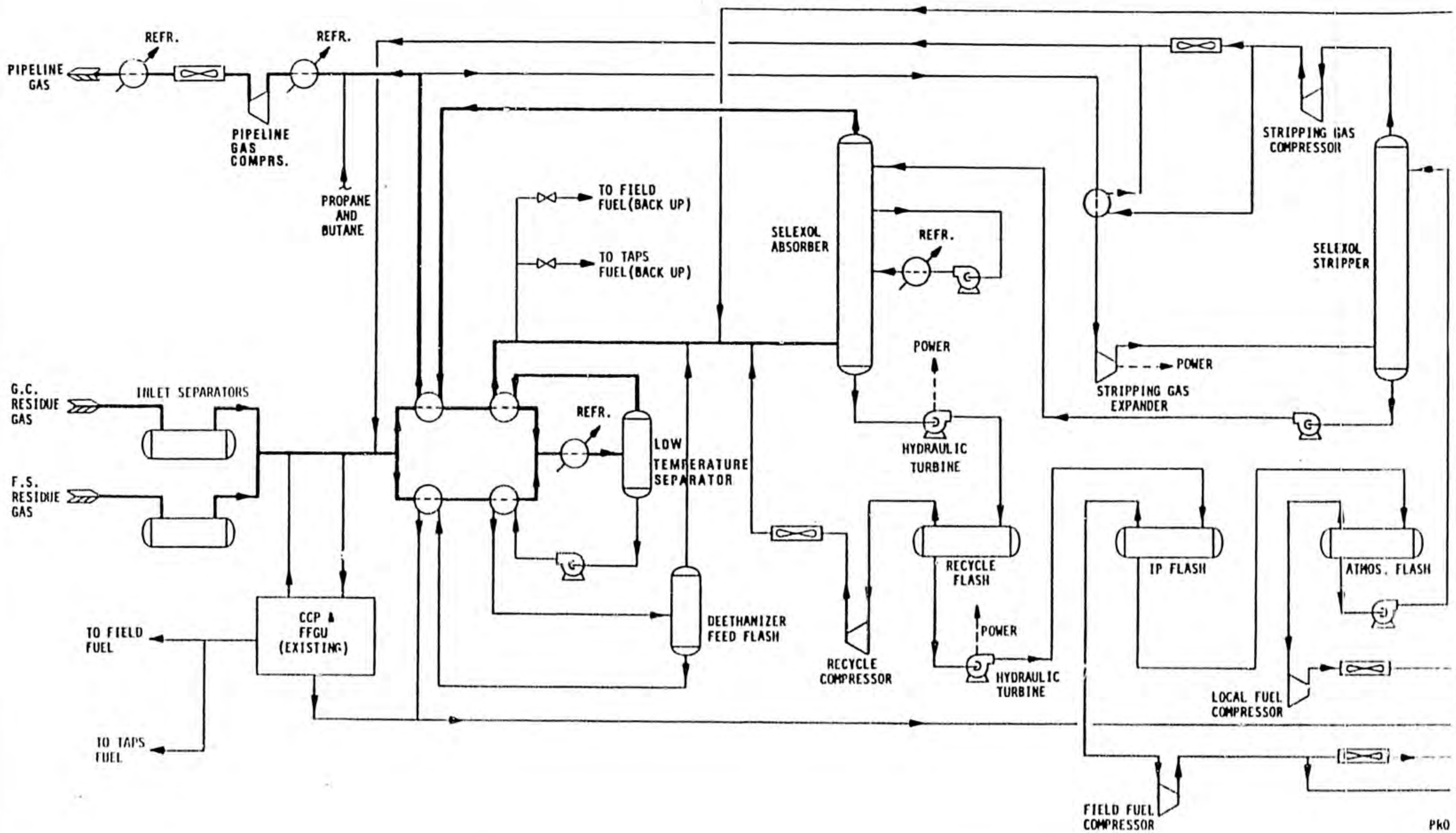


Figure 1
Block Flow Diagram
Sales Gas Conditioning
Facilities



REFR.
REFR.
PIPELINE GAS
PIPELINE GAS COMPRS.
PROPANE AND BUTANE
TO FIELD FUEL (BACK UP)
TO TAPS FUEL (BACK UP)
SELEXOL ABSORBER
REFR.
STRIPPING GAS COMPRESSOR
SELEXOL STRIPPER
POWER
POWER
STRIPPING GAS EXPANDER
G.C. RESIDUE GAS
INLET SEPARATORS
REFR.
LOW TEMPERATURE SEPARATOR
HYDRAULIC TURBINE
F.S. RESIDUE GAS
RECYCLE FLASH
IP FLASH
ATMOS. FLASH
CCP & FFGU (EXISTING)
TO FIELD FUEL
DEETHANIZER FEED FLASH
RECYCLE COMPRESSOR
POWER
HYDRAULIC TURBINE
LOCAL FUEL COMPRESSOR
TO TAPS FUEL
FIELD FUEL COMPRESSOR
PkO

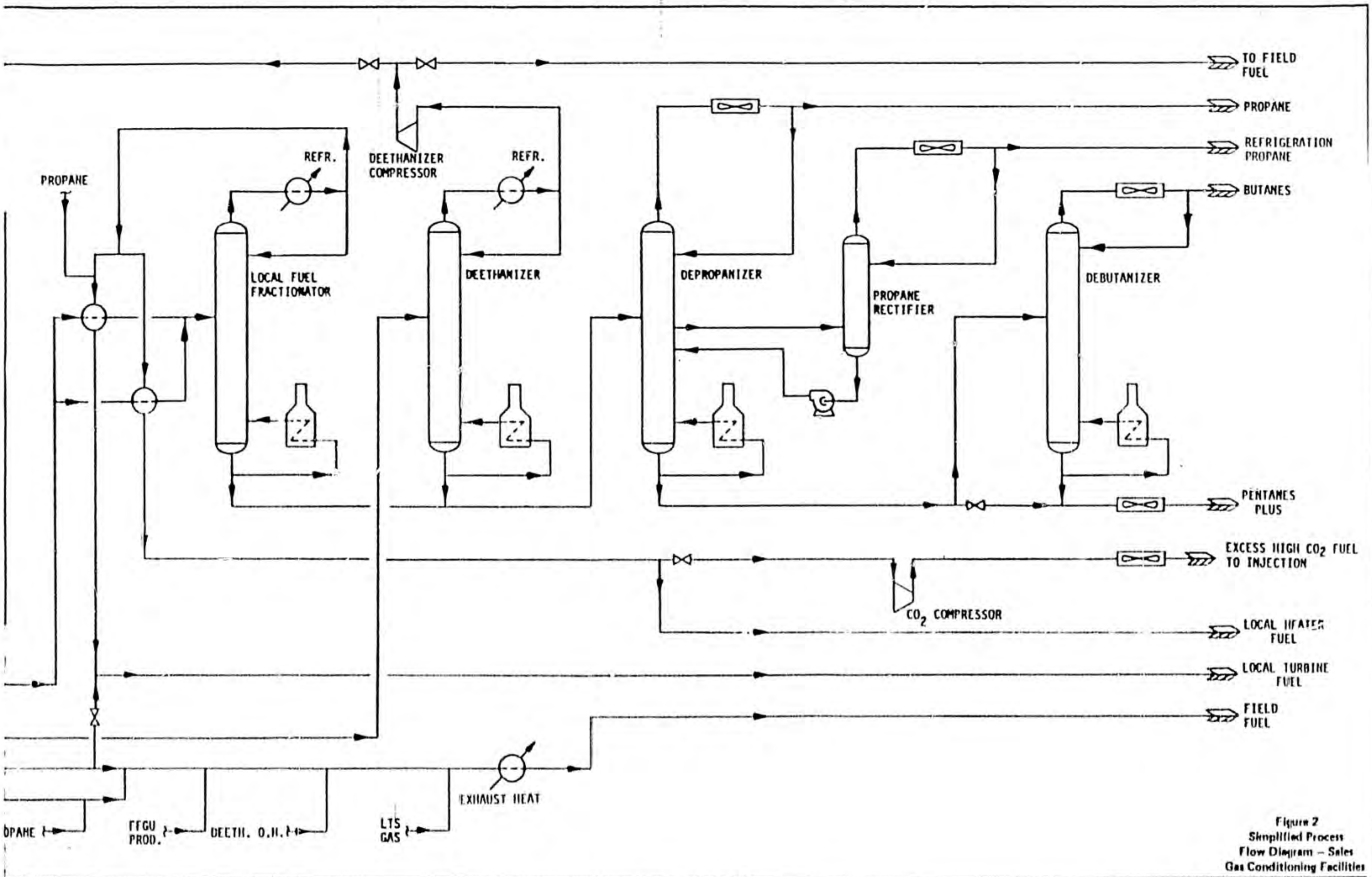


Figure 2
Simplified Process Flow Diagram – Sales Gas Conditioning Facilities

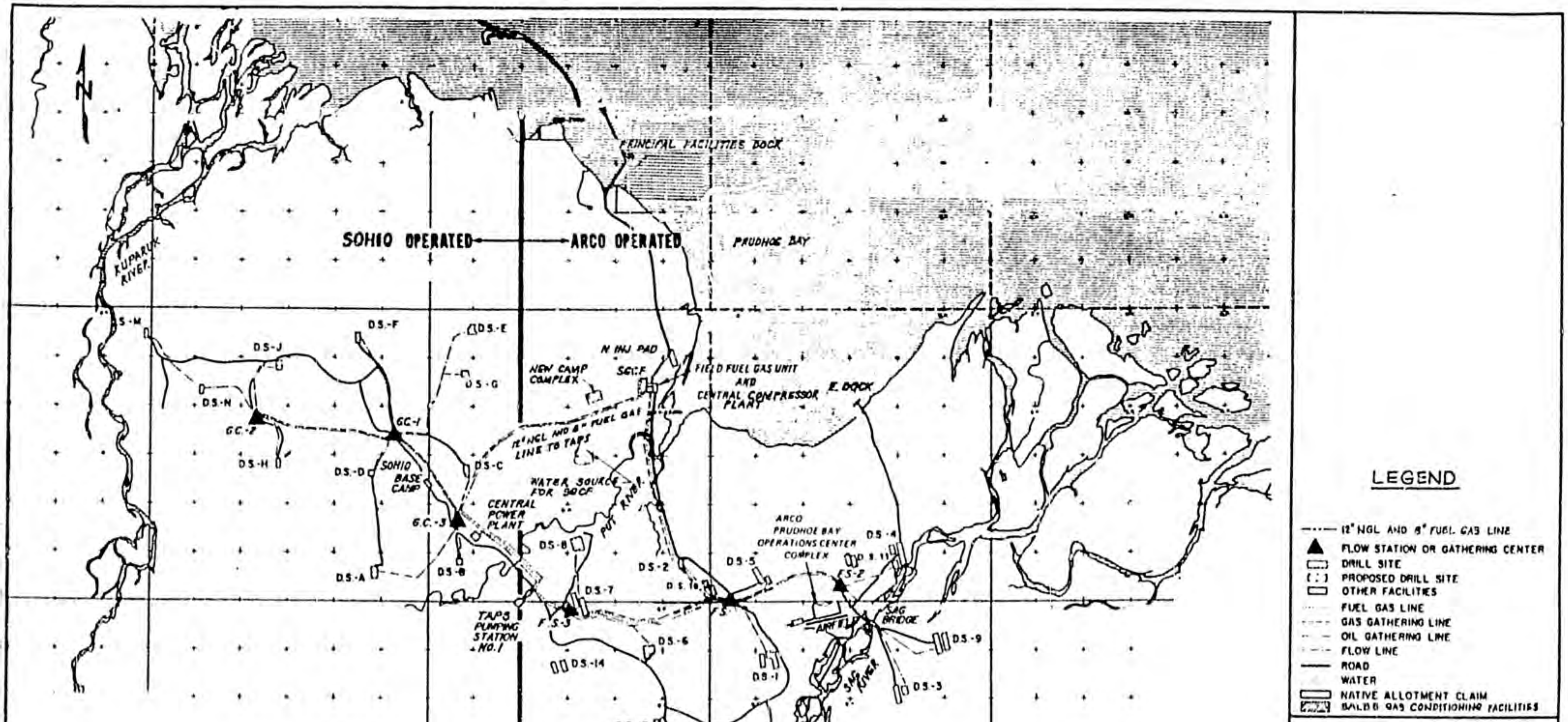


FIGURE 3
SITE LOCATION MAP

INDEX

UNIT 31: (5795-D-31-CA1)
NGL EXTRACTION & CO_2 REMOVAL FACILITIES (TRAIN NO. 1)

MODULE NO.	DESCRIPTION
31-01	SELEXOL ABSORBER
31-02	LOW TEMPERATURE SEPARATION
31-03	GAS COOLING
31-04	SELEXOL INTERCOOLER
31-05	RECYCLE FLASH
31-06	FIELD FUEL & ATMOSPHERIC FLASH
31-07	LOCAL FUEL COMPRESSION
31-08	SELEXOL STRIPPER
31-09	H.P. SELEXOL PUMP
31-10	STRIPPER OVERHEAD COMPRESSION
31-11	PIPEWAY

UNIT 32: (5795-D-32-CA1)
NGL EXTRACTION & CO_2 REMOVAL FACILITIES (TRAIN NO. 2)

MODULE NO.	DESCRIPTION
32-01	SELEXOL ABSORBER
32-02	LOW TEMPERATURE SEPARATION
32-03	GAS COOLING
32-04	SELEXOL INTERCOOLER
32-05	RECYCLE FLASH
32-06	FIELD FUEL & ATMOSPHERIC FLASH
32-07	LOCAL FUEL COMPRESSION
32-08	SELEXOL STRIPPER
32-09	H.P. SELEXOL PUMP
32-10	STRIPPER OVERHEAD COMPRESSION
32-11	PIPEWAY

UNIT 33: (5795-D-33-CA1)
NGL EXTRACTION & CO_2 REMOVAL FACILITIES (TRAIN NO. 3)

MODULE NO.	DESCRIPTION
33-01	SELEXOL ABSORBER
33-02	LOW TEMPERATURE SEPARATION
33-03	GAS COOLING
33-04	SELEXOL INTERCOOLER
33-05	RECYCLE FLASH
33-06	FIELD FUEL & ATMOSPHERIC FLASH
33-07	LOCAL FUEL COMPRESSION
33-08	SELEXOL STRIPPER
33-09	H.P. SELEXOL PUMP
33-10	STRIPPER OVERHEAD COMPRESSION
33-11	PIPEWAY

UNIT 34: (5795-D-34-CA1)
NGL EXTRACTION & CO_2 REMOVAL FACILITIES (TRAIN NO. 4)

MODULE NO.	DESCRIPTION
34-01	SELEXOL ABSORBER
34-02	LOW TEMPERATURE SEPARATION
34-03	GAS COOLING
34-04	SELEXOL INTERCOOLER
34-05	RECYCLE FLASH
34-06	FIELD FUEL & ATMOSPHERIC FLASH
34-07	LOCAL FUEL COMPRESSION
34-08	SELEXOL STRIPPER
34-09	H.P. SELEXOL PUMP
34-10	STRIPPER OVERHEAD COMPRESSION
34-11	PIPEWAY

UNIT 41: (5795-D-41-CA1)
SALES GAS COMPRESSION FACILITIES

MODULE NO.	DESCRIPTION
41-01	SALES GAS SUCTION CHILLER
41-02	SALES GAS CHILLER
41-03	CONTROL AND AUXILIARIES (COMPRESSION)

UNIT 45: (5795-D-45-CA1)
REFRIGERATION FACILITIES

MODULE NO.	DESCRIPTION
45-01	ECONOMIZER
45-02	SCRUBBER (45-1001)
45-03	SCRUBBER (45-1002)
45-04	COMPRESSOR (45-1401)
45-05	COMPRESSOR (45-1402)
45-06	UTILIDOR
45-07	SCRUBBER (45-1003)
45-08	SCRUBBER (45-1004)
45-09	COMPRESSOR (45-1403)
45-10	COMPRESSOR (45-1404)
45-11	UTILIDOR
45-12	SCRUBBER (45-1005)
45-13	UTILIDOR
45-14	COMPRESSOR (45-1405)
45-15	SURGE DRUMS
45-16	PIPEWAY
45-17 THRU 45-31	PROPANE CONDENSER
45-32	PIPEWAY
45-33 THRU 45-47	PROPANE CONDENSER

UNIT 46: (5795-D-46-CA1)
FIELD FUEL & CO_2 COMPRESSION FACILITIES

MODULE NO.	DESCRIPTION
46-01	FIELD FUEL COMPRESSOR (46-1401 & 1411)
46-02	CO_2 COMPRESSOR (46-1421 & 1431)
46-03	FIELD FUEL COMPRESSOR (46-1402 & 1412)
46-04	CO_2 COMPRESSOR (46-1422 & 1432)

UNIT 51: (5795-D-51-CA1)
FRACTIONATION & NGL SHIPPING & INJECTION FACILITIES

MODULE NO.	DESCRIPTION
51-01	LOCAL FUEL FRACTIONATION
51-02	DEETHANIZER COMPRESSION
51-03	UTILIDOR
51-04	UTILIDOR
51-05	DEETHANIZER
51-06	UTILIDOR
51-07	UTILIDOR
51-08	DEPROPANIZER
51-09	UTILIDOR
51-10	UTILIDOR
51-11	DEBUTANIZER
51-12	UTILIDOR
51-13	UTILIDOR
51-14	REBOILERS HEATERS (51-1401 & 1402)
51-15	UTILIDOR
51-16	REBOILERS HEATERS (51-1403 & 1404)
51-17	UTILIDOR
51-18	REBOILERS HEATERS (51-1405 & 1406)
51-19	PROPANE STORAGE
51-20	UTILIDOR
51-21	NGL MIX DRUM
51-22	BUTANE/PENTANE STORAGE
51-23	UTILIDOR
51-24	REFRIGERANT STORAGE
51-25	AIRCOOLER (51-1305 DEPROPANIZER COND)
51-26	AIRCOOLER (51-1305 DEPROPANIZER COND)
51-27	AIRCOOLER (51-1306 DEBUTANIZER COND)
51-28	AIRCOOLER (51-1307 DEBUTANIZER BOTTOMS)
51-29	AIRCOOLER (51-1313 INJECTION LIQUID COOLER)
51-30	PIPEWAY
51-31	PIPEWAY

UNIT 61: (5795-D-61-CA1)
OFFSITE & UTILITY FACILITIES & IN-PLOT INTERCONNECTING PIPEWAY

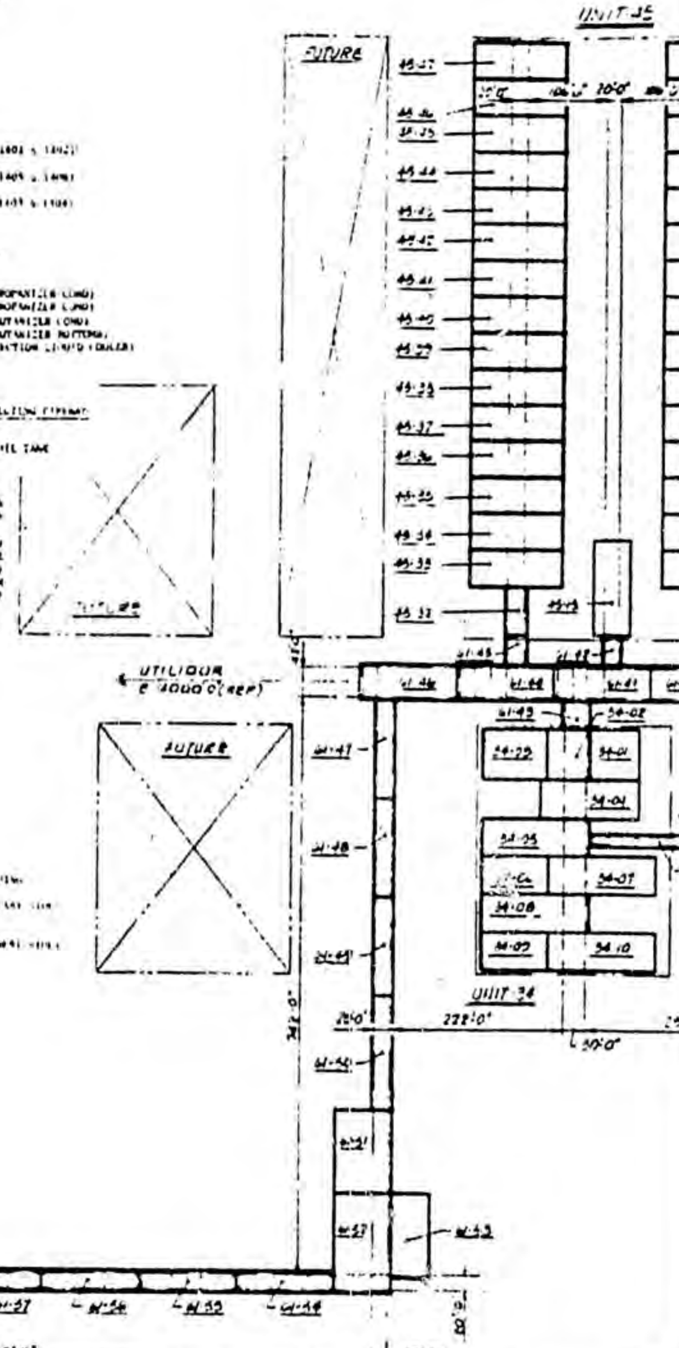
MODULE NO.	DESCRIPTION
61-01	DIESEL STORAGE & SLOP OIL TANK
61-02	SELEXOL STORAGE
61-03	[illegible]
61-04	[illegible]
61-05	MAINTENANCE & WAREHOUSE
61-06	MAINTENANCE & WAREHOUSE
61-07	[illegible]
61-08	GLYCOL HEATER (61-1401)
61-09	GLYCOL SOLUTION STORAGE
61-10	GLYCOL HEATER (61-1402)
61-11	NITROGEN UTILITY SYSTEM
61-12	GLYCOL HEATER (61-1403)
61-13	AIR COMPRESSOR
61-14	CENTRAL CONTROL ROOM
61-15	CENTRAL CONTROL ROOM
61-16 THRU 61-50	UTILIDOR
61-51	L.P. FLARE
61-52	H.P. FLARE
61-53	H.P. FLARE
61-54 THRU 61-66	SLEEPER WAY
61-67	FLARE MANIFOLD
61-68 THRU 61-79	PIPEWAY

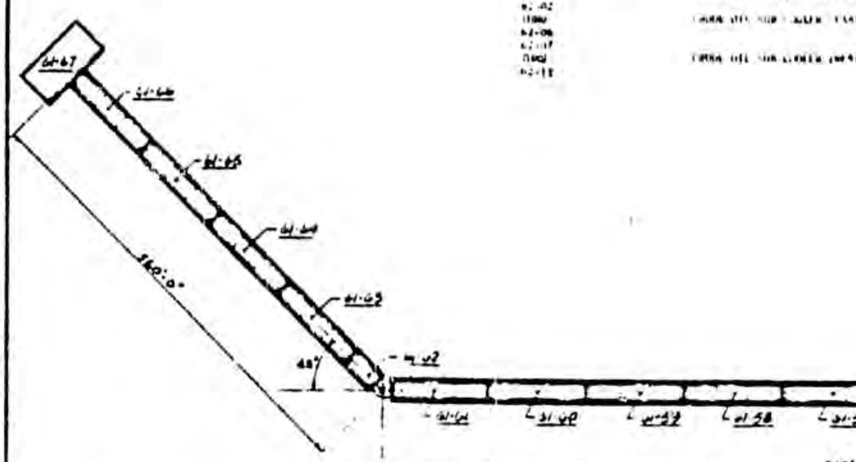
UNIT 62: (5795-D-62-CA1)
CRUDE OIL COOLING FACILITIES

MODULE NO.	DESCRIPTION
62-01	CRUDE OIL UTILITY BUILDING
62-02 THRU 62-06	CRUDE OIL AIR COOLER (EAST SIDE)
62-07 THRU 62-11	CRUDE OIL AIR COOLER (WEST SIDE)

UNIT 63: (5795-D-63-CA1)
ELECTRICAL GENERATION AND DISTRIBUTION

MODULE NO.	DESCRIPTION
63-01	UTILIDOR
63-02	GENERATOR CONTROL BUILDING
63-03	GENERATOR BUILDING NO. 1
63-04	GENERATOR BUILDING NO. 2
63-05	GENERATOR BUILDING NO. 3
63-06	EMERGENCY GENERATORS (250KW)





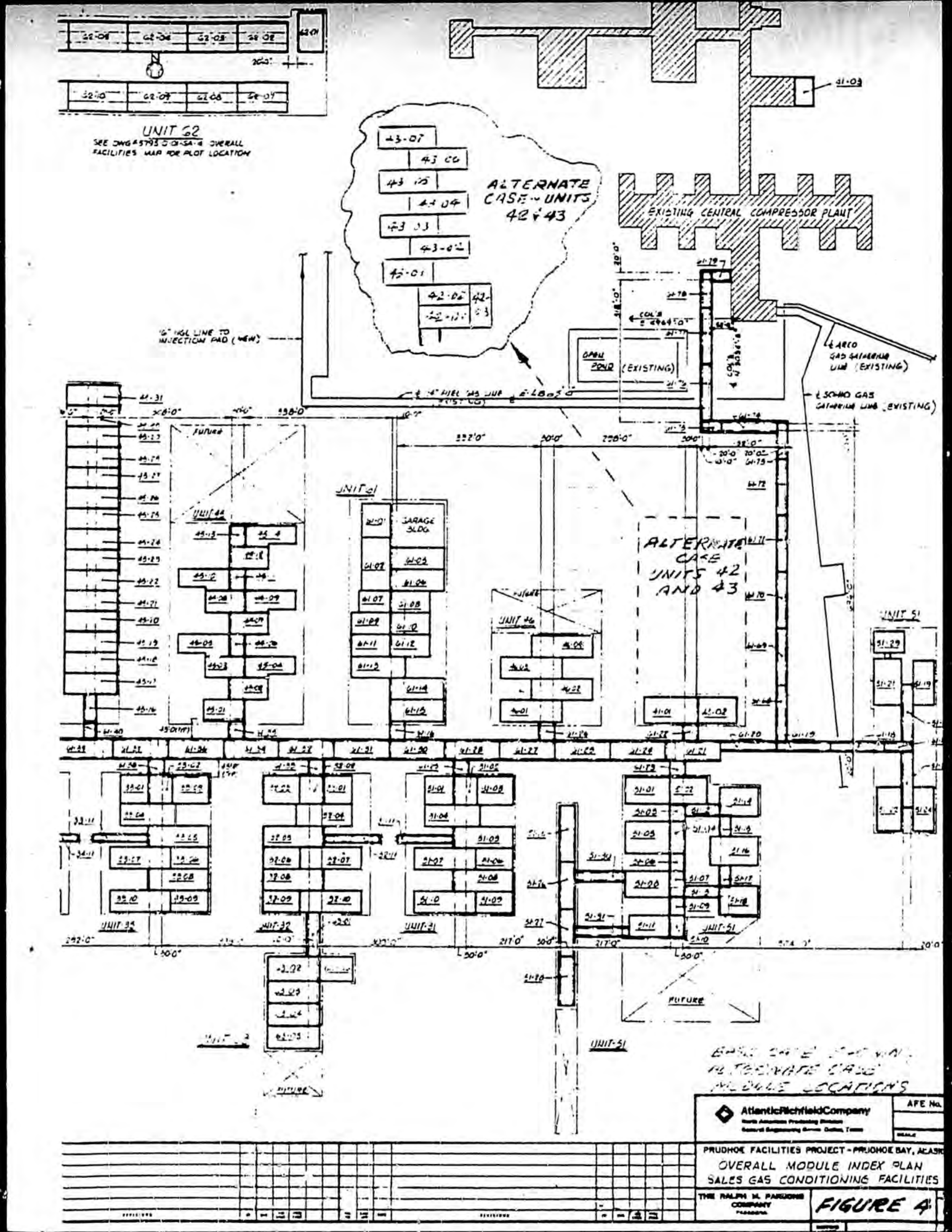

UNIT 62
ALTERNATE CASE - UNITS 42 & 43
EXISTING CENTRAL COMPRESSOR PLANT
OPEN POND (EXISTING)
ARCO GAS GATHERING LINE (EXISTING)
SOHIO GAS GATHERING LINE (EXISTING)
GARAGE BLDG
FUTURE
ALTERNATE CASE UNITS 42 AND 43
UNIT 61
UNIT 51
FUTURE
AtlanticRichfieldCompany
AFE No.
PRUDHOE FACILITIES PROJECT - PRUDHOE BAY, ALASKA
OVERALL MODULE INDEX PLAN
SALES GAS CONDITIONING FACILITIES
THE RALPH M. PARSONS COMPANY
FIGURE 4

FIGURE 5

SALES GAS CONDITIONING FACILITIES

IMPLEMENTATION PLAN

MILESTONE SCHEDULE

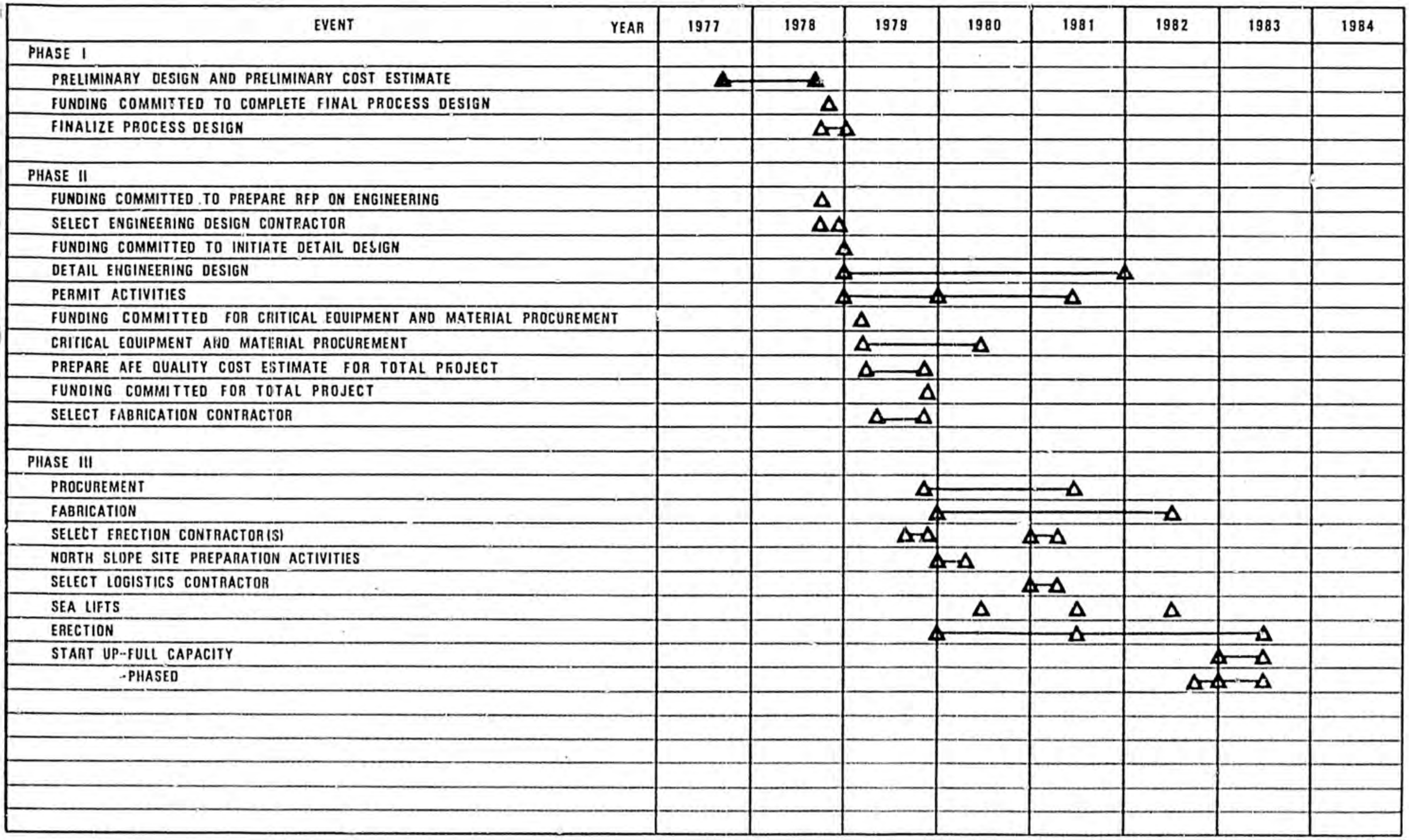
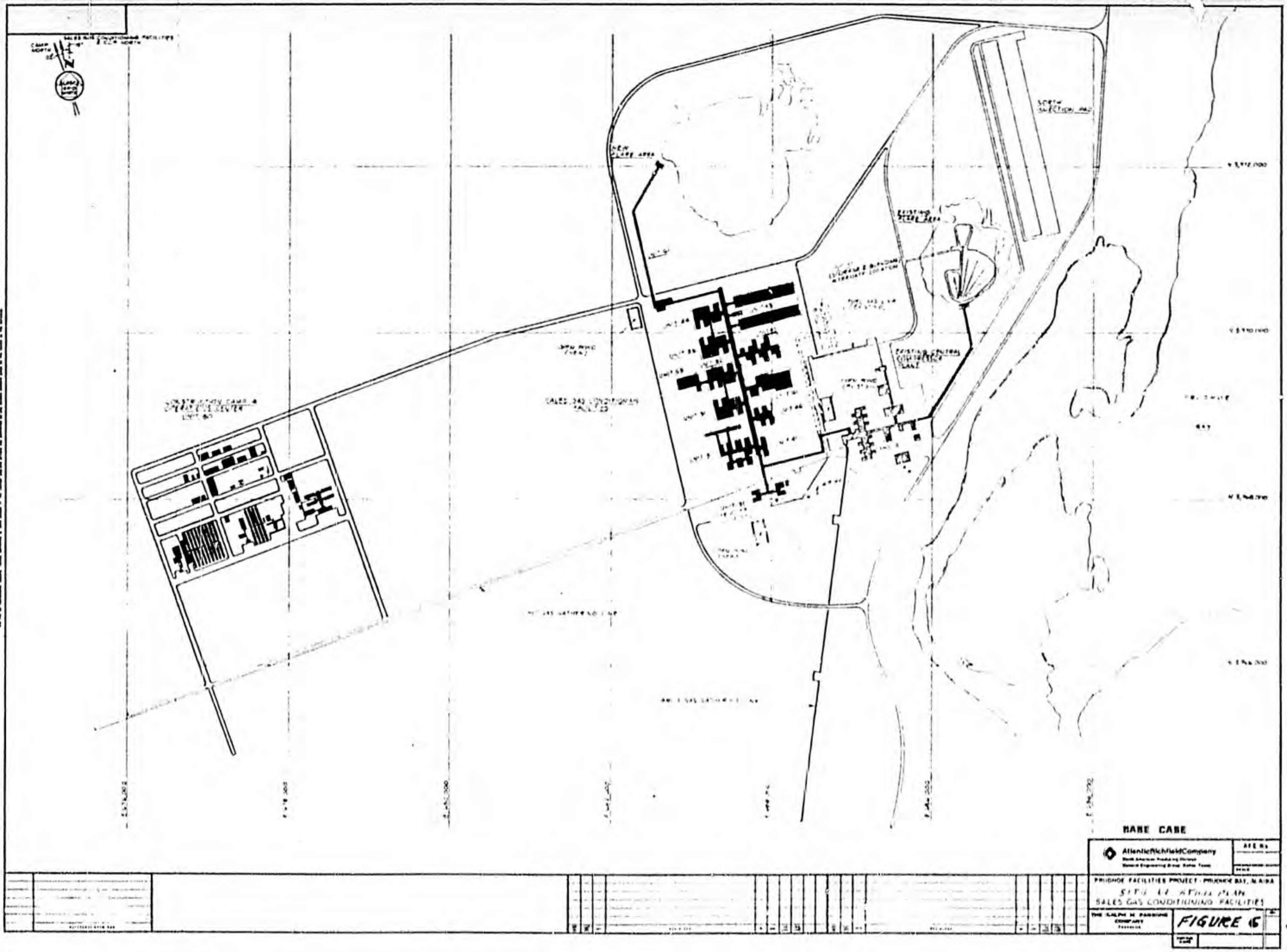
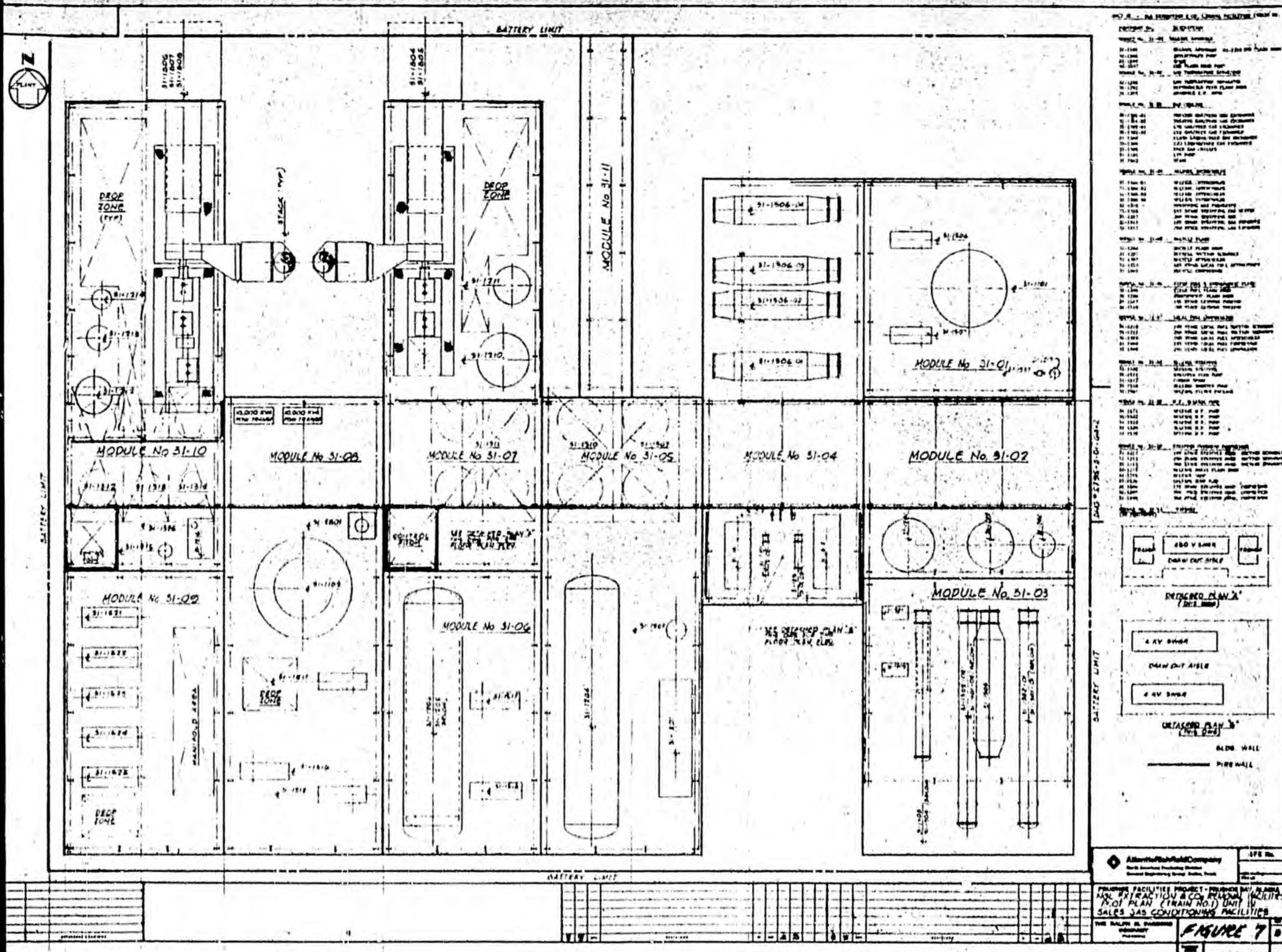
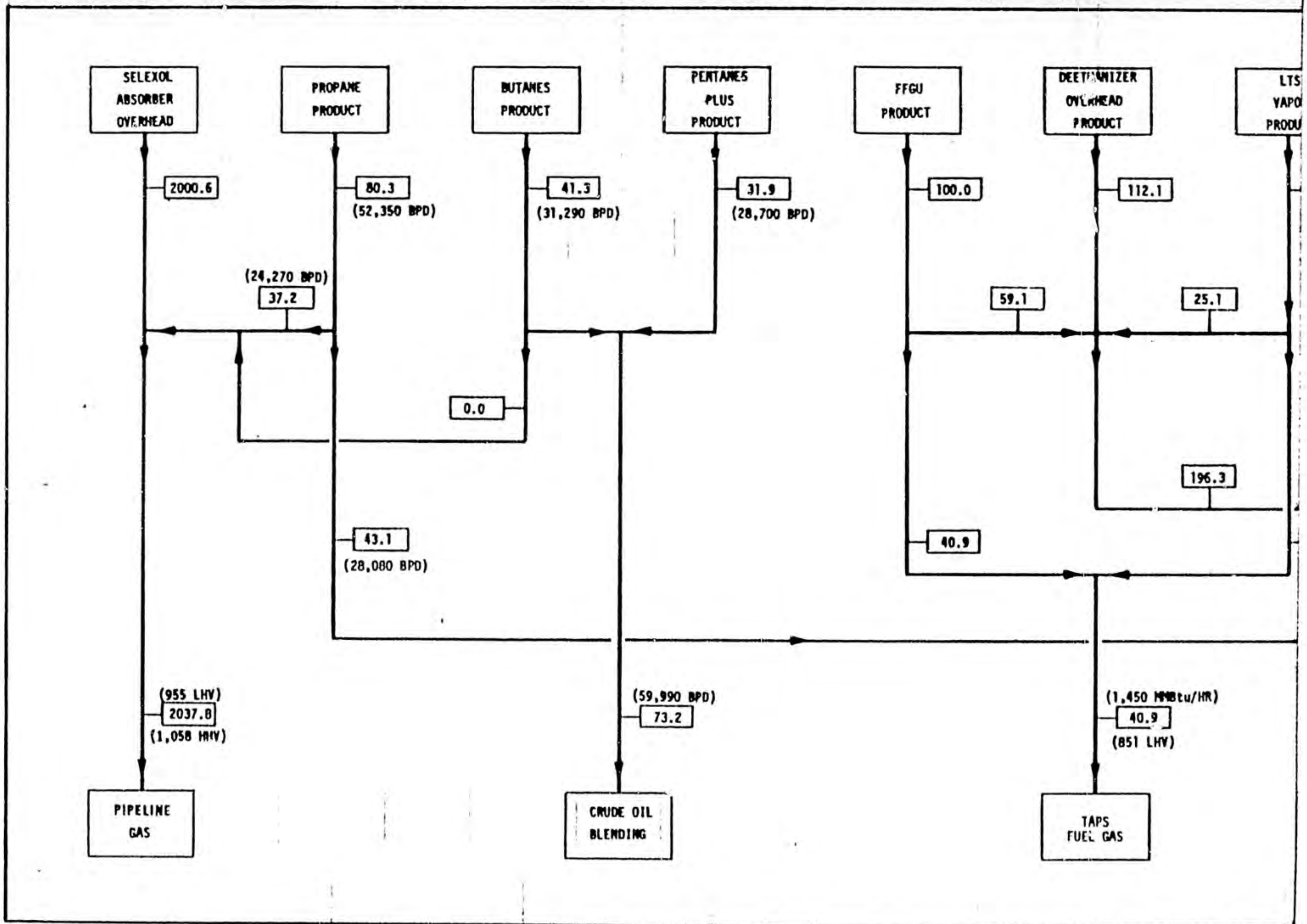


FIGURE 5

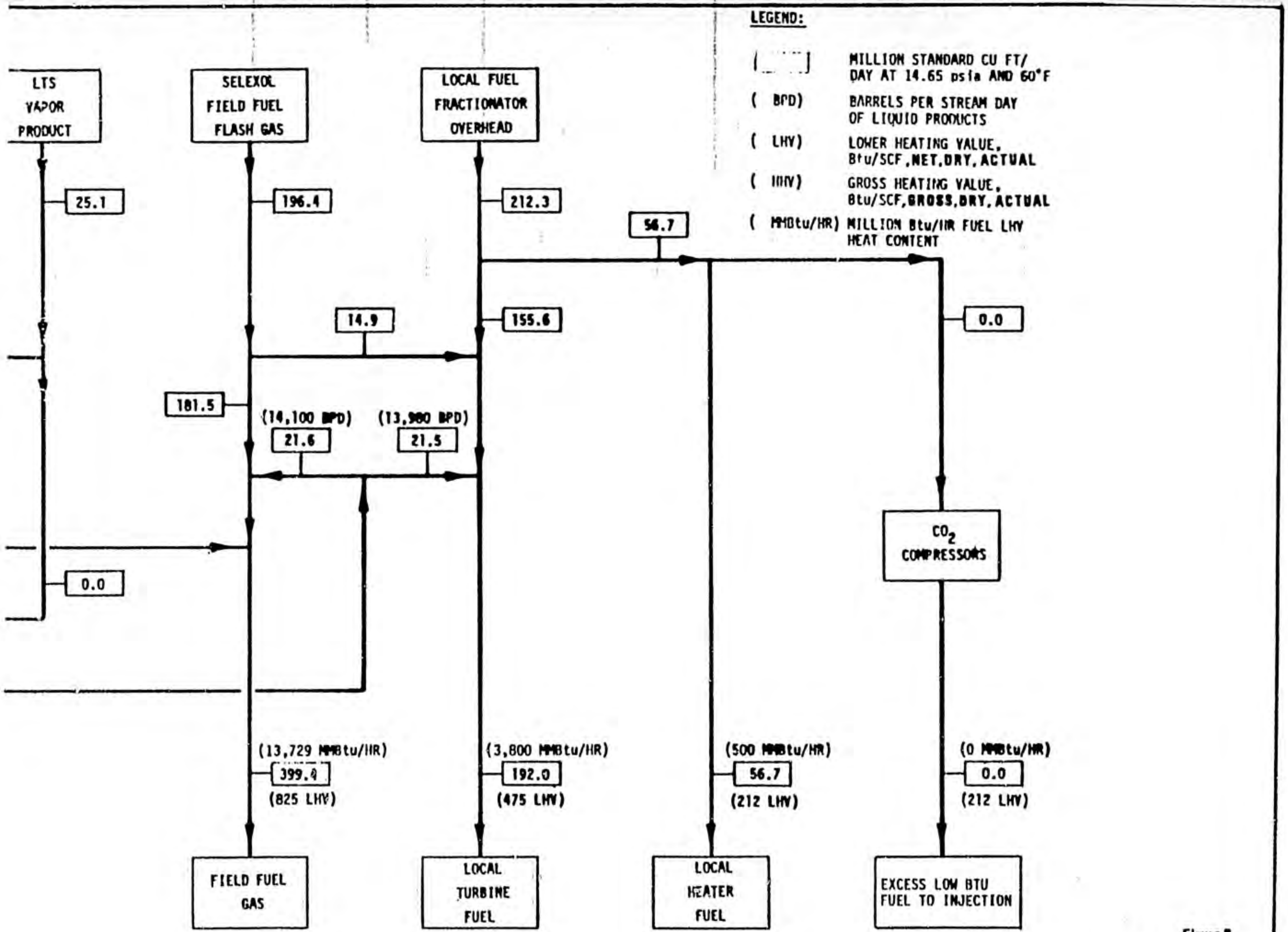

BASE CASE
AtlanticRichfieldCompany
PRUDHOE FACILITIES PROJECT · PRUDHOE BAY, ALASKA
SALES GAS CONDITIONING FACILITIES
FIGURE 6


BATTERY LIMIT
DROP ZONE (TYP)
STACK (TYP)
DROP ZONE
MODULE No 31-11
MODULE No 31-01
MODULE No 31-10
MODULE No 31-08
MODULE No 31-07
MODULE No 31-05
MODULE No 31-04
MODULE No. 31-02
MODULE No 31-09
MODULE No 31-06
MODULE No. 31-03
MANIFOLD AREA
CONTROL ROOM
DETACHED PLAN 'A'
DRAW OUT AISLE
DETACHED PLAN 'B'
BLDG. WALL
FIREWALL
AtlanticRichfieldCompany
NGL EXTRACTION & CO2 REMOVAL FACILITIES
PLOT PLAN (TRAIN NO.1) UNIT 31
SALES GAS CONDITIONING FACILITIES
FIGURE 7



SELEXOL ABSORBER OVERHEAD
PROPANE PRODUCT
BUTANES PRODUCT
PENTANES PLUS PRODUCT
FFGU PRODUCT
DEETHANIZER OVERHEAD PRODUCT
LTS VAPO PRODU
2000.6
80.3
(52,350 BPD)
41.3
(31,290 BPD)
31.9
(28,700 BPD)
100.0
112.1
(24,270 BPD)
37.2
59.1
25.1
0.0
196.3
43.1
(28,080 BPD)
40.9
(955 LHV)
2037.8
(1,058 HHV)
(59,990 BPD)
73.2
(1,450 MMBtu/HR)
40.9
(851 LHV)
PIPELINE GAS
CRUDE OIL BLENDING
TAPS FUEL GAS

3



EXAMPLE CASE (SEE SECTION 8)

Figure 8
Block Flow Diagram
Of Product And Fuel
Gas Blending

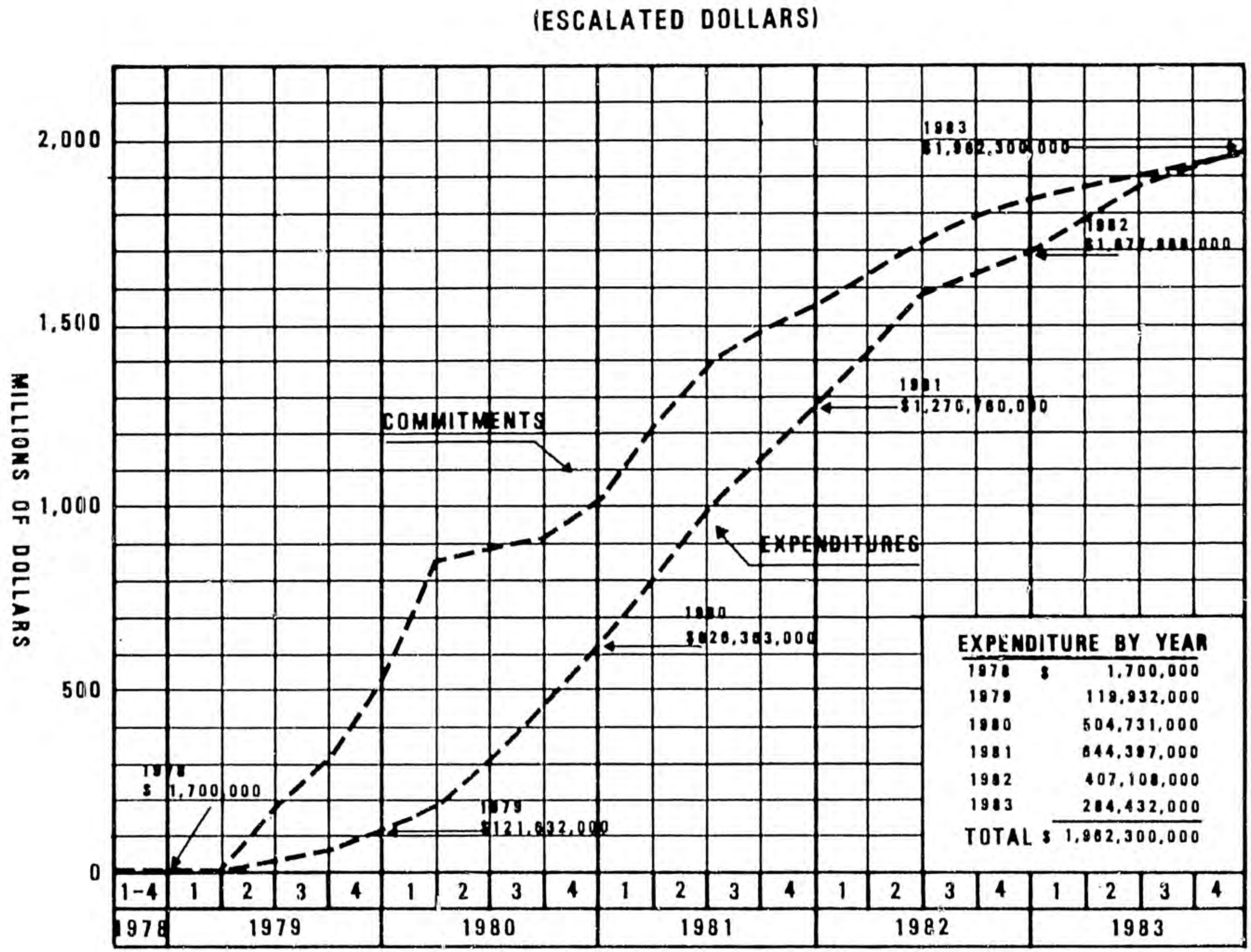


FIGURE 9
SALES GAS CONDITIONING FACILITIES
PATTERN OF COMMITMENTS AND EXPENDITURES
(BASE CASE - FULL-CAPACITY STARTUP)