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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
RMP

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.

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*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
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. Pipeline 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.

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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 syste

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 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:

Full Capacity Delivery Year	Incremental Delay Cost: Base Case (Dollars)	Incremental Delay Cost: 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

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

Year	Quarter	Base Case		Alternate Case	
		Phased Startup	Full Startup	Phased Startup	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.

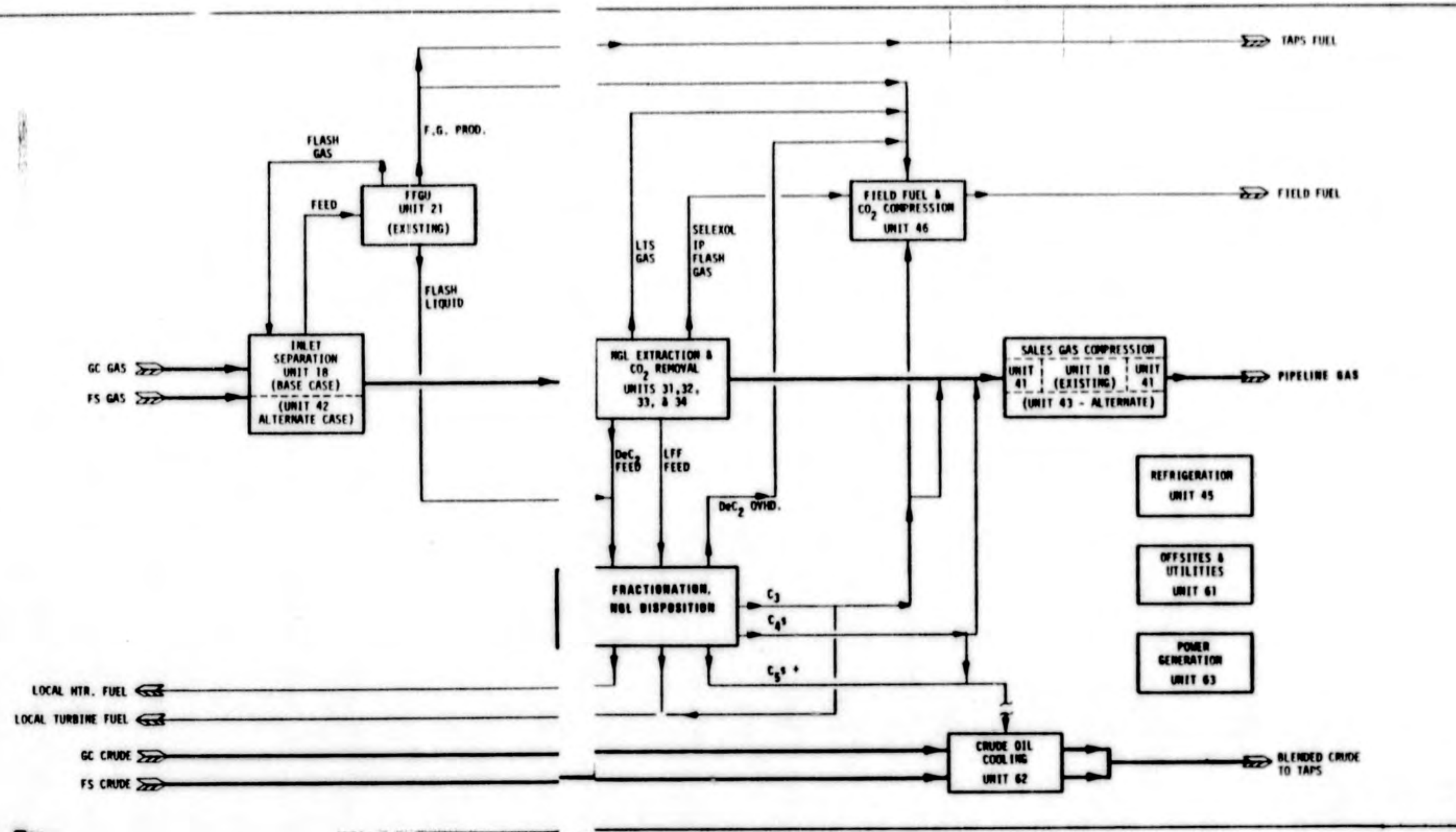
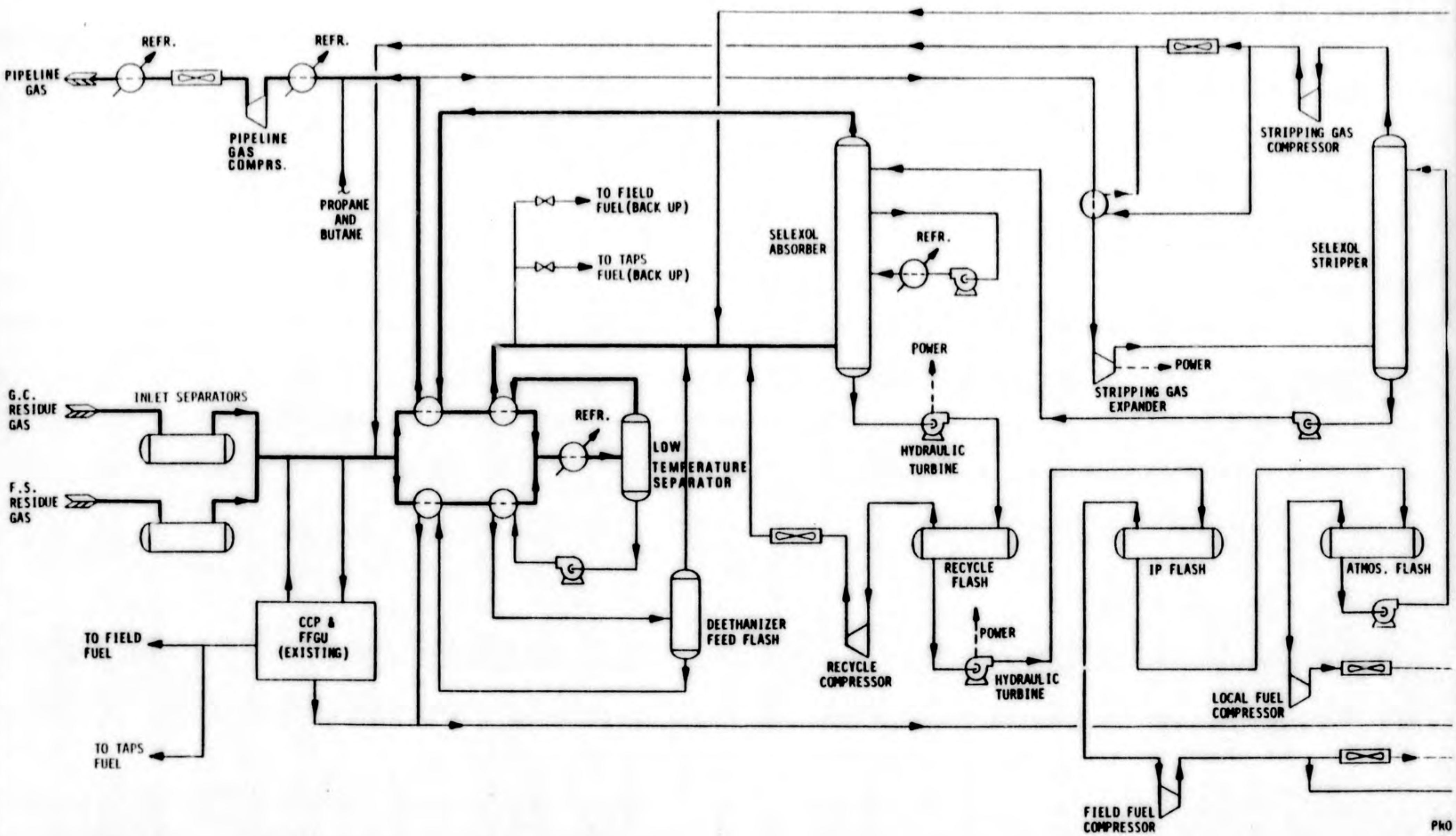


Figure 1
Block Flow Diagram
Sales Gas Conditioning
Facilities


PIPELINE GAS
REFR.
REFR.
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
F.S. RESIDUE GAS
REFR.
LOW TEMPERATURE SEPARATOR
HYDRAULIC TURBINE
RECYCLE FLASH
IP FLASH
ATMOS. FLASH
CCP & FFGU (EXISTING)
TO FIELD FUEL
TO TAPS FUEL
DEETHANIZER FEED FLASH
RECYCLE COMPRESSOR
POWER
HYDRAULIC TURBINE
LOCAL FUEL COMPRESSOR
FIELD FUEL COMPRESSOR

PkO

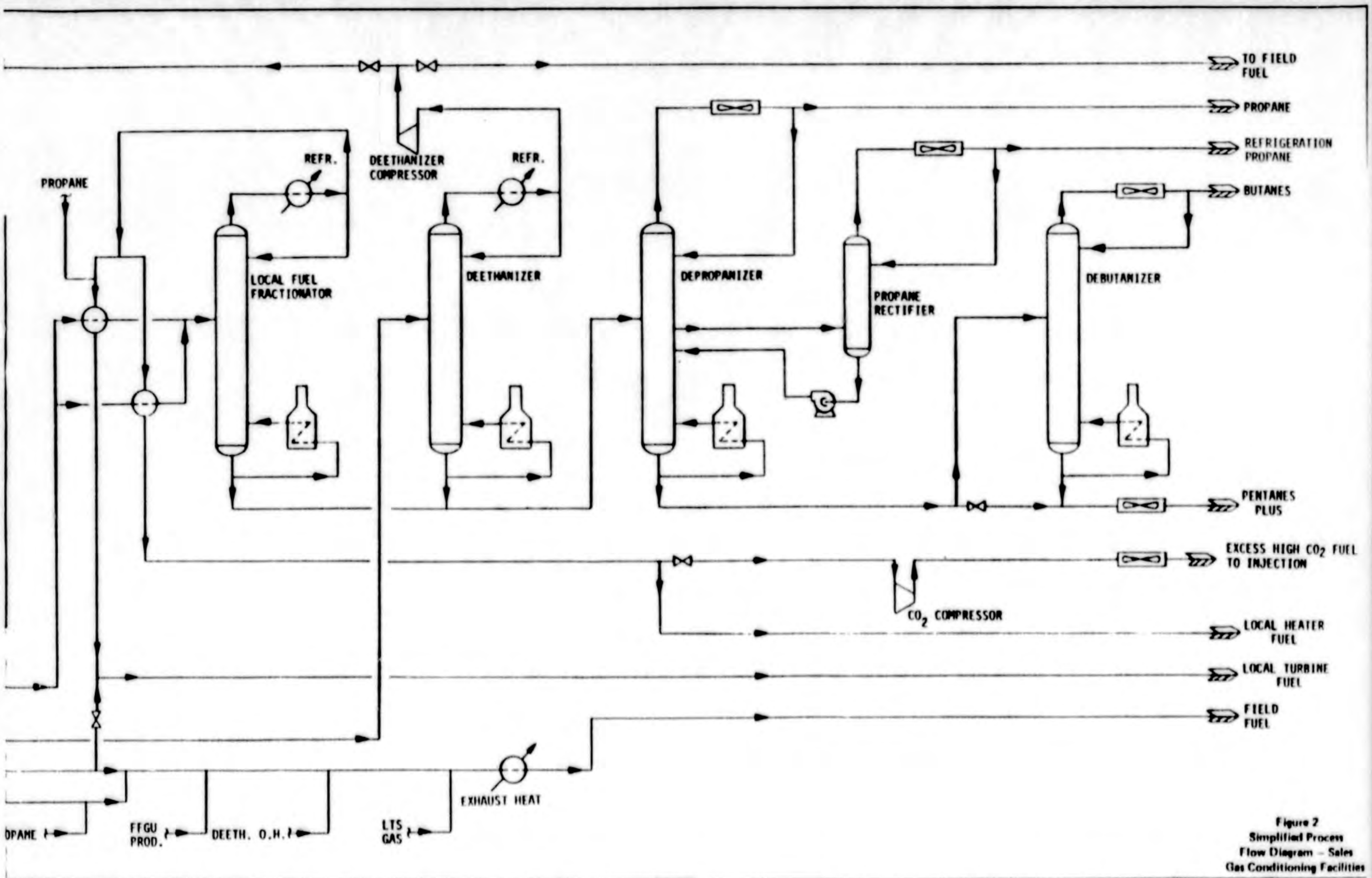


Figure 2
Simplified Process
Flow Diagram – Sales
Gas Conditioning Facilities

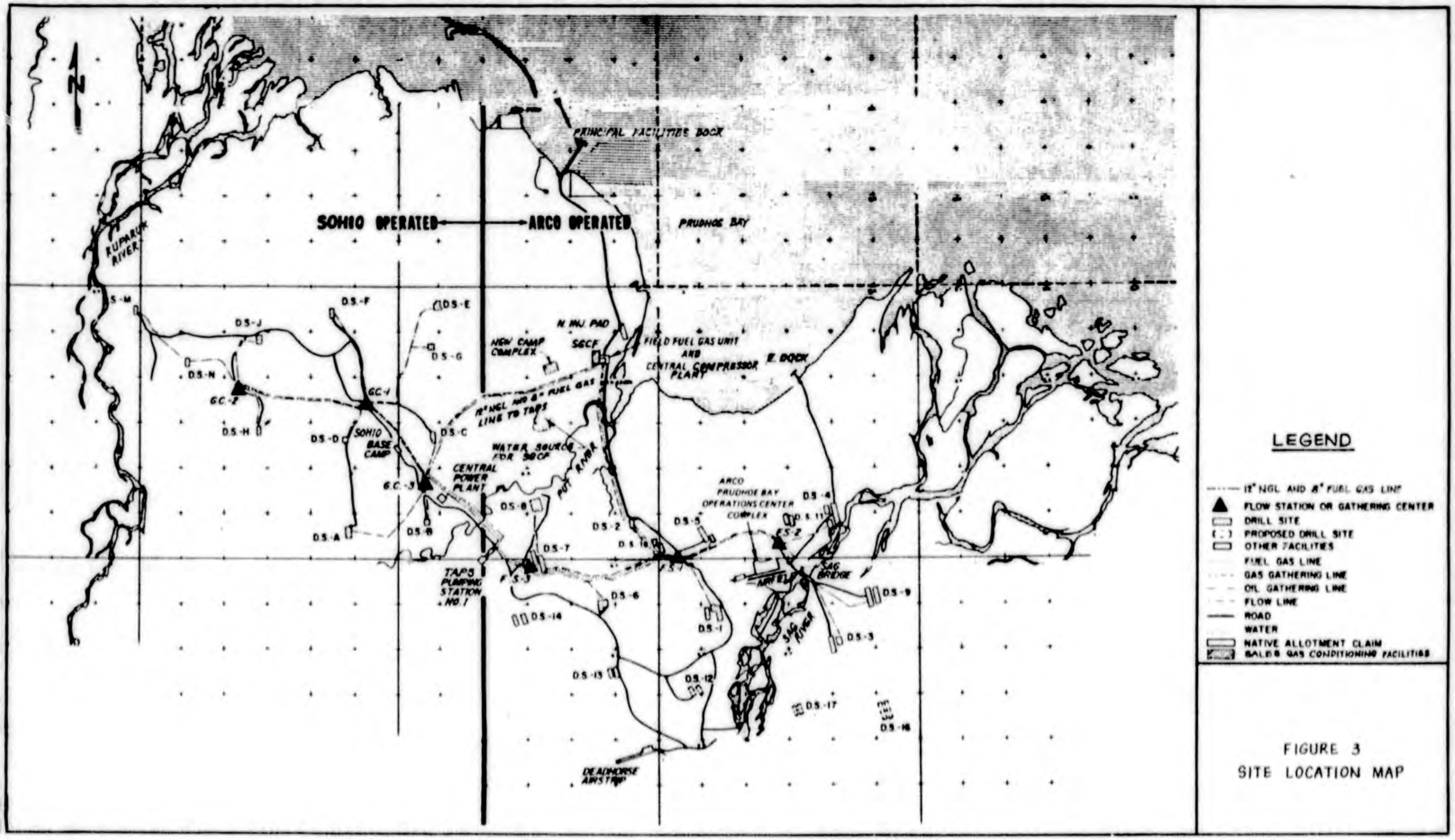


FIGURE 3
SITE LOCATION MAP

UNIT 31. (5795-0-31-GA1)
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 COMPRESSOR
31-11	PIPEWAY

UNIT 32. (5795-0-32-GA1)
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 COMPRESSOR
32-11	PIPEWAY

UNIT 33. (5795-0-33-GA1)
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 COMPRESSOR
33-11	PIPEWAY

UNIT 34. (5795-0-34-GA1)
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 COMPRESSOR
34-11	PIPEWAY

UNIT 41. (5795-0-41-GA1)
SALES GAS COMPRESSION FACILITIES

MODULE NO.	DESCRIPTION
41-01	SALES GAS SUCTION CHILLER
41-02	SALES GAS CHILLER
41-03	CONTROL AND SWITCHGEAR [illegible]

UNIT 45. (5795-0-45-GA1)
REFRIGERATION FACILITIES

MODULE NO.	DESCRIPTION
45-01	ECONOMIZER
45-02	SCRUBBER (45-1001)
45-03	SCRUBBER (45-1002)
45-04	COMPRESSOR (45-1001)
45-05	COMPRESSOR (45-1002)
45-06	UTILIDOR
45-07	SCRUBBER (45-1003)
45-08	SCRUBBER (45-1004)
45-09	COMPRESSOR (45-1003)
45-10	COMPRESSOR (45-1004)
45-11	UTILIDOR
45-12	SCRUBBER (45-1005)
45-13	UTILIDOR
45-14	COMPRESSOR (45-1005)
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. (5795-0-46-GA1)
FIELD FUEL & CO_2 COMPRESSION FACILITIES

MODULE NO.	DESCRIPTION
46-01	FIELD FUEL COMPRESSOR (46-1001 & 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. (5795-0-51-GA1)
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	REBOILER HEATERS (51-1401 & 1402)
51-15	UTILIDOR
51-16	REBOILER HEATERS (51-1403 & 1404)
51-17	UTILIDOR
51-18	REBOILER 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 (51-1315 INJECTION LIQUID COOLER)
51-30	PIPEWAY
51-31	PIPEWAY

UNIT 61. (5795-0-61-GA1)
SERVICE & UTILITY FACILITIES & ON-PLOT INTERCONNECTING PIPEWAY

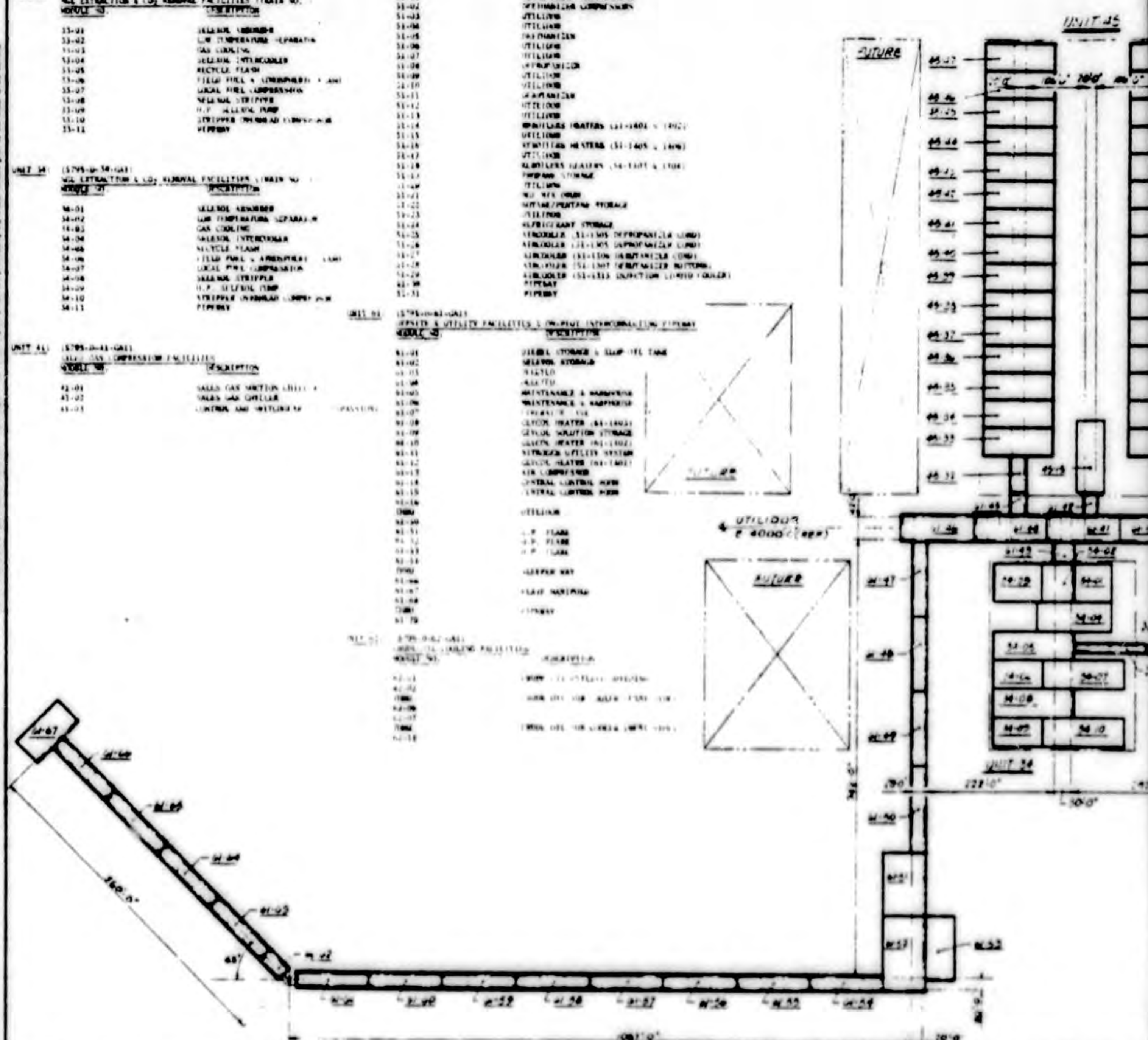
MODULE NO.	DESCRIPTION
61-01	DIESEL STORAGE & SLOP OIL TANK
61-02	SELEXOL STORAGE
61-03	DELETED
61-04	DELETED
61-05	MAINTENANCE & WAREHOUSE
61-06	MAINTENANCE & WAREHOUSE
61-07	[illegible]
61-08	GLYCOL HEATER (61-1401)
61-09	GLYCOL SOLUTION STORAGE
61-10	GLYCOL HEATER (61-1402)
61-11	NITROGEN UTILITY SYSTEM
61-12	GLYCOL HEATER (61-1403)
61-13	AIR COMPRESSOR
61-14	CENTRAL CONTROL ROOM
61-15	CENTRAL CONTROL ROOM
61-16 THRU 61-30	UTILIDOR
61-31	L.P. FLARE
61-32	H.P. FLARE
61-33	H.P. FLARE
61-34 THRU 61-43	SLEEPER WAY
61-44	FLARE MANIFOLD
61-45 THRU 61-70	PIPEWAY

UNIT 62. (5795-0-62-GA1)
[illegible] COOLING FACILITIES

MODULE NO.	DESCRIPTION
62-01	[illegible]
62-02 THRU 62-06	[illegible]
62-07 THRU 62-11	[illegible]

UNIT 63. (5795-0-63-GA1)
ELECTRICAL GENERATION AND [illegible]

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 (STORAGE)



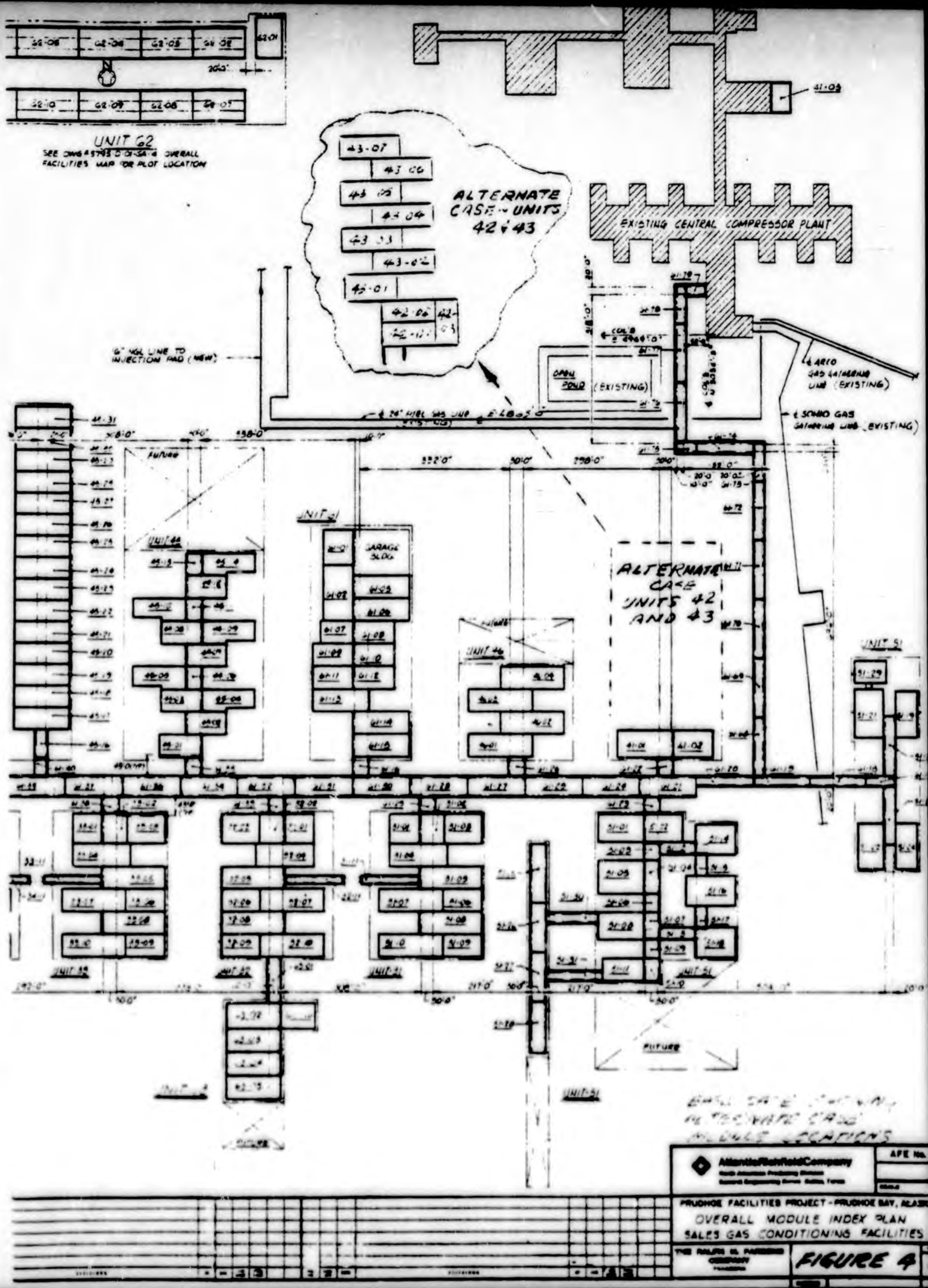

UNIT 62
SEE DWG 5793 D 01-SA-4 OVERALL FACILITIES MAP FOR PLOT LOCATION
ALTERNATE CASE - UNITS 42 & 43
EXISTING CENTRAL COMPRESSOR PLANT
6" NGL LINE TO INJECTION PAD (NEW)
OPEN POND (EXISTING)
℄ ARCO GAS GATHERING LINE (EXISTING)
℄ SOHIO GAS GATHERING LINE (EXISTING)
FUTURE
UNIT 44
UNIT 61
GARAGE BLDG
UNIT 46
ALTERNATE CASE UNITS 42 AND 43
UNIT 51
UNIT 53
UNIT 52
UNIT 21
UNIT 31
UNIT 51
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

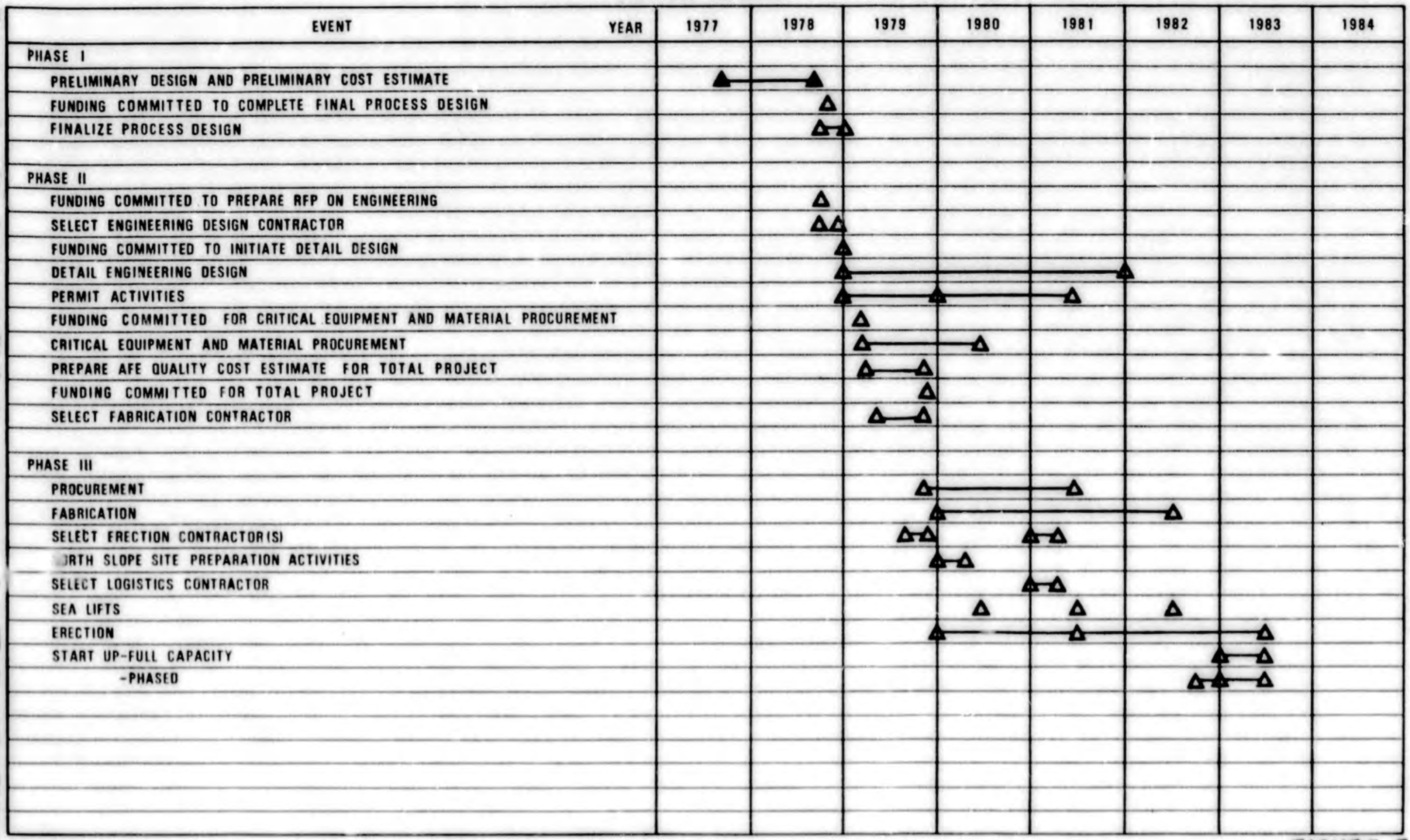
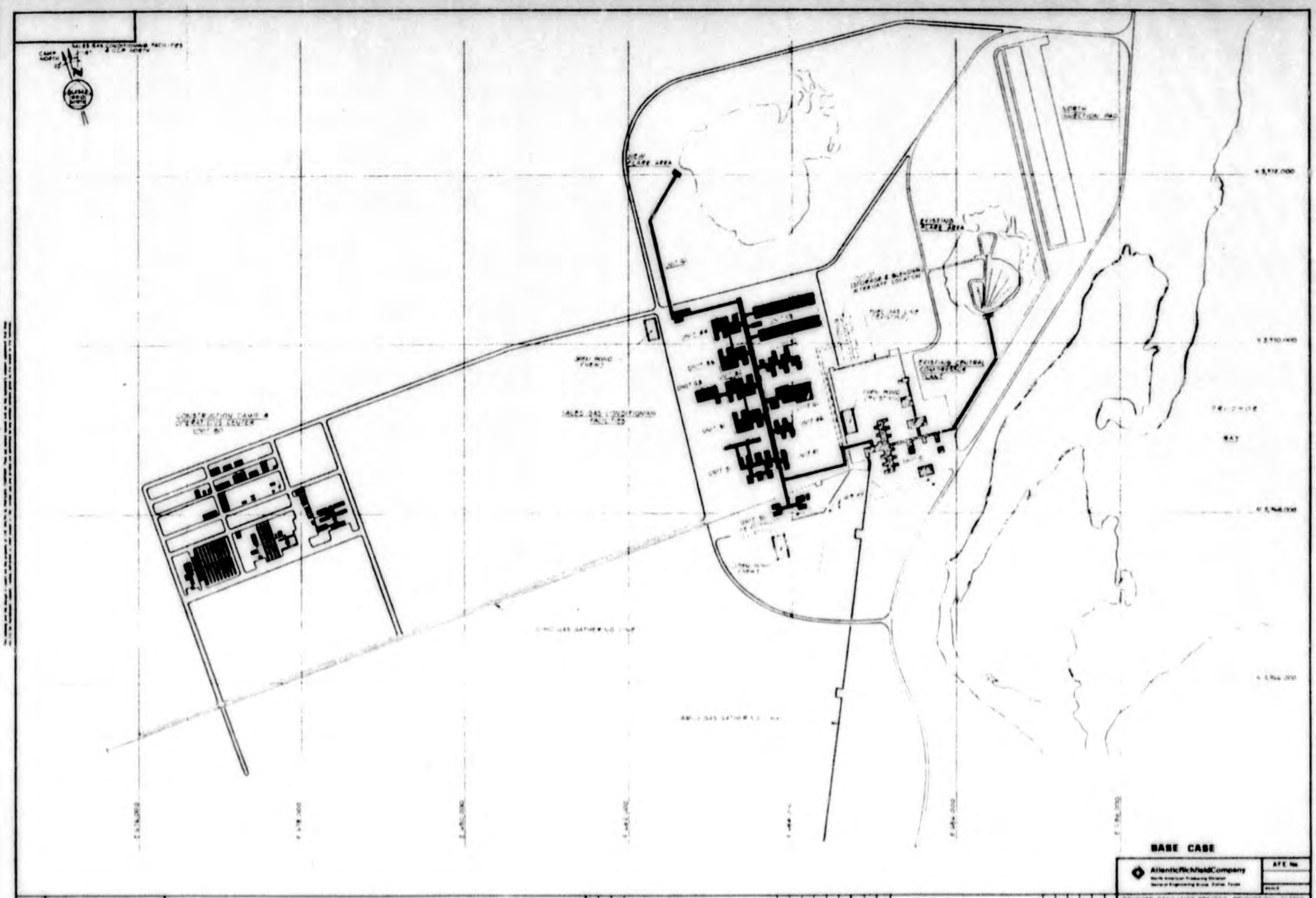


FIGURE 5



BASE CASE

AtlanticRichfieldCompany

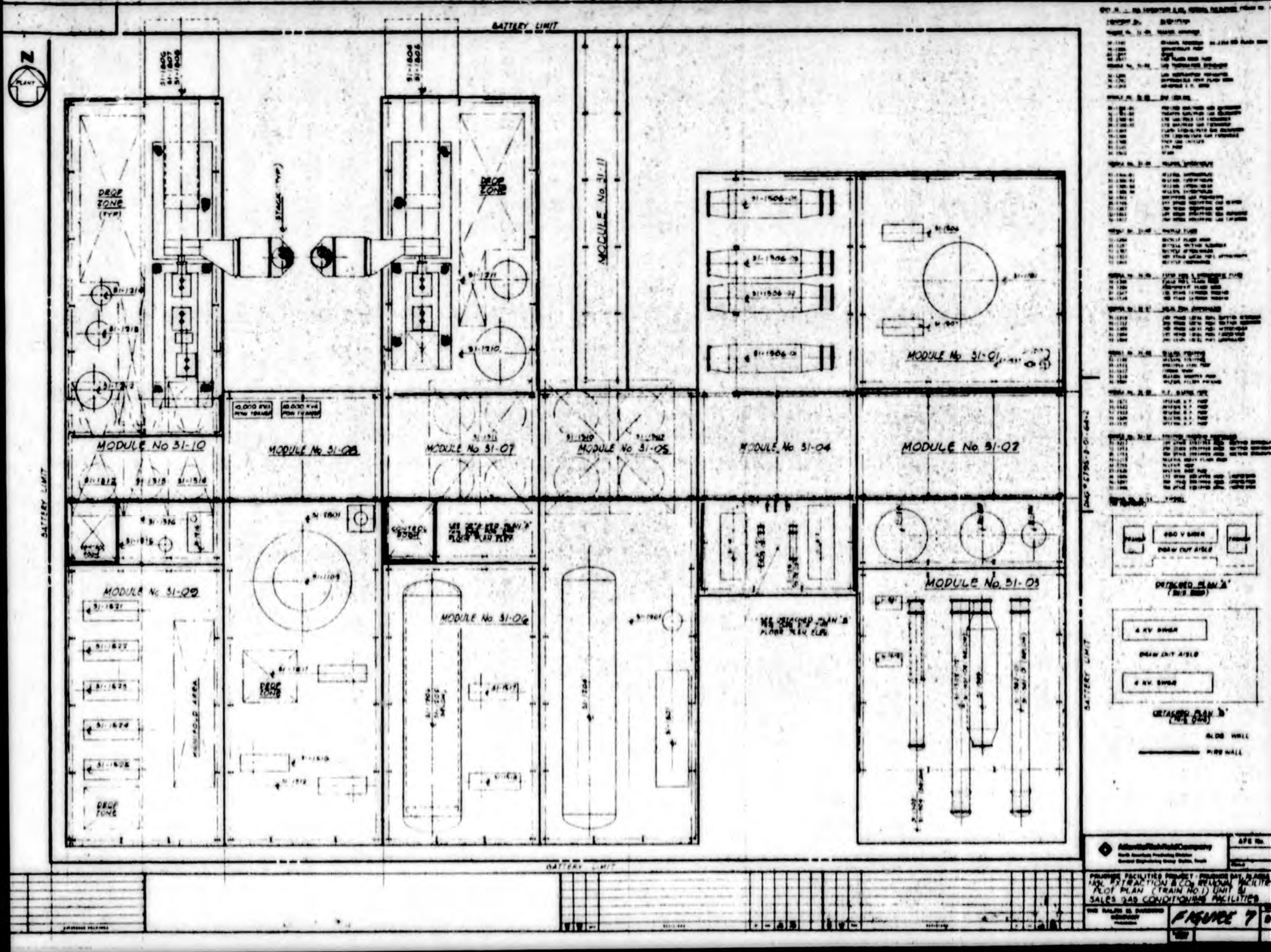
AFE No.

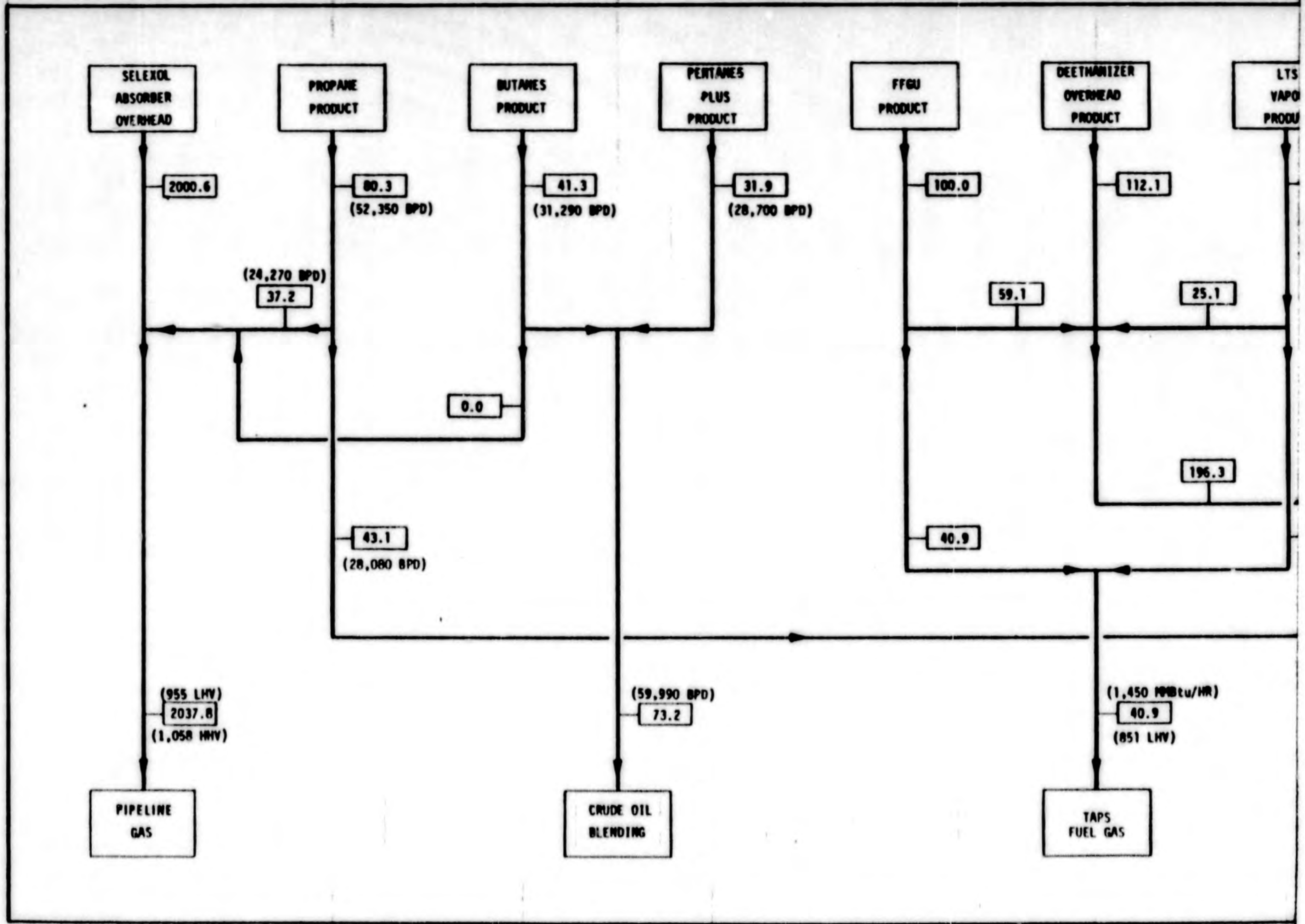
PRUDHOE FACILITIES PROJECT · PRUDHOE BAY, ALASKA

SALES GAS CONDITIONING FACILITIES

THE RALPH M. PARSONS COMPANY

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
CONTROL ROOM
MANIFOLD AREA
DETACHED PLAN 'A'
DETACHED PLAN 'B'
480 V SWGR
DRAW OUT AISLE
4 KV SWGR
FIRE WALL
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

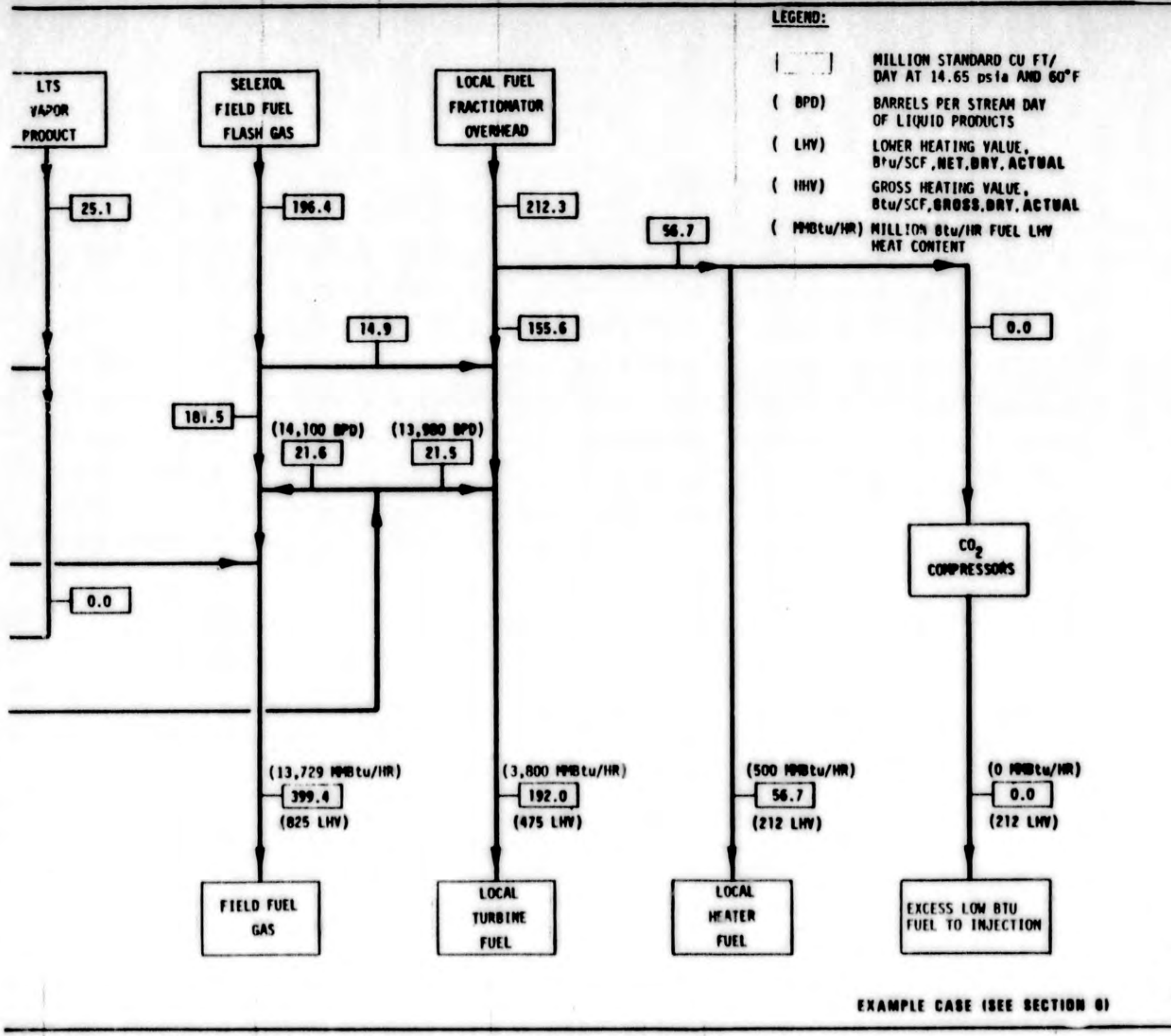


Figure 8
Block Flow Diagram
Of Product And Fuel
Gas Blending

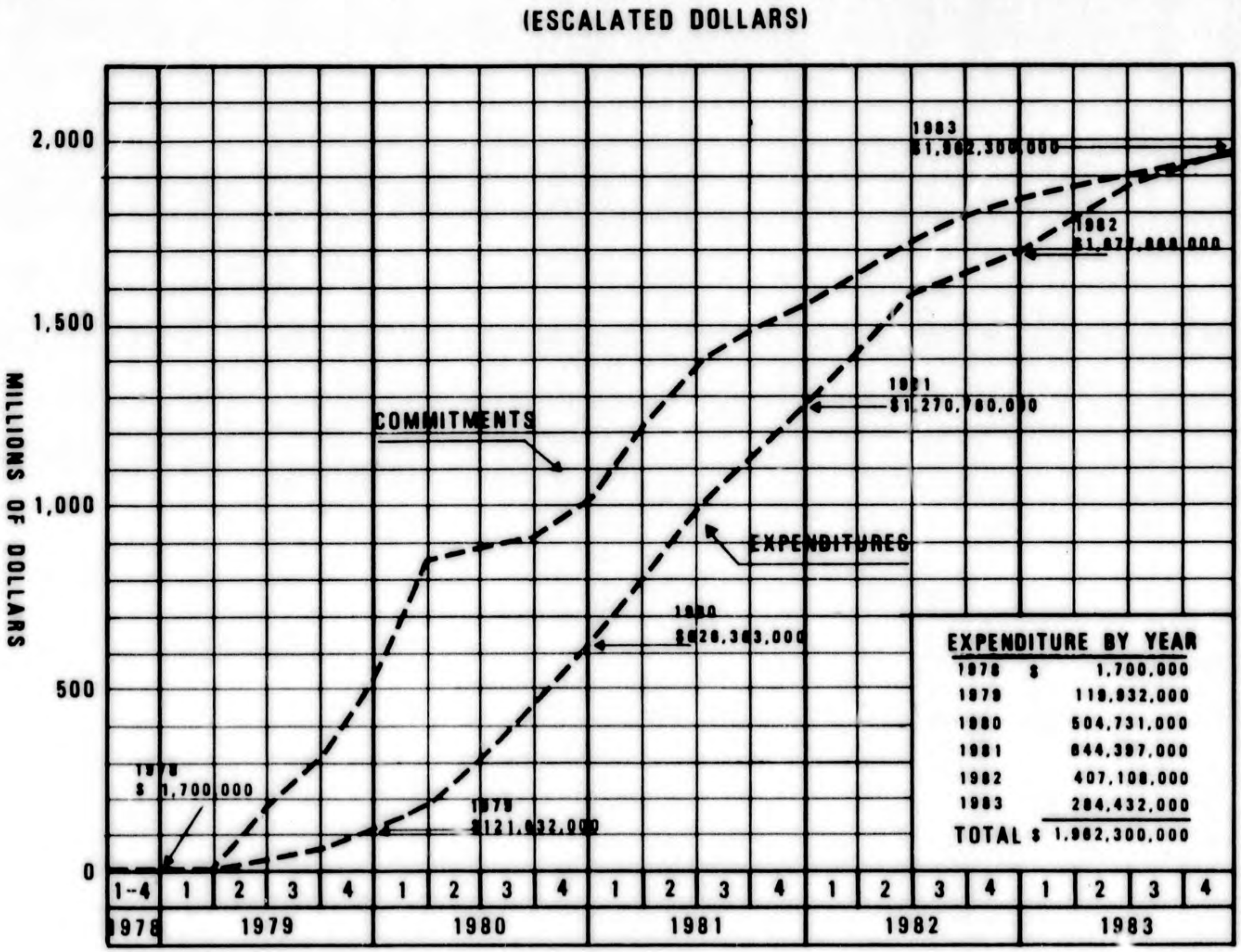


FIGURE 9
SALES GAS CONDITIONING FACILITIES
PATTERN OF COMMITMENTS AND EXPENDITURES
(BASE CASE - FULL-CAPACITY STARTUP)

PLEASE NOTE: THE PRECEDING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT.

PLEASE NOTE: THE FOLLOWING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT

NORTHWEST ALASKAN PIPELINE COMPANY

LUCIUS D. LEGG
VICE PRESIDENT
PROJECT OPERATIONS

PRO-80-2200

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SALT LAKE CITY, UTAH 84110
801-534-3897

3333 MICHELSON DRIVE
IRVINE, CALIFORNIA 92730
714-975-3401

December 30, 1980

507.0

The Honorable Terry Miller
Lieutenant Governor of the
State of Alaska
State Capitol Building
Pouch AA
Juneau, AK 99811

Dear Sir:

Subject: Draft Report on Alternatives for Gas Conditioning Facility Location

During the August 22, 1980, meeting of the Alaskan Northwest Natural Gas Transportation Company Board of Partners, held in Salt Lake City, you requested that a report be prepared on the alternatives to locating the Alaska Gas Conditioning Plant at Prudhoe Bay. The enclosed draft has been prepared by Northwest Alaskan Pipeline Company in conjunction with producer company representatives. It provides a technical and economic evaluation of five alternative cases, four of which involve some degree of gas conditioning in the interior of Alaska. It concludes that the case based on conditioning at Prudhoe Bay is the preferred alternative.

To facilitate review of this draft, we are sending C. E. Behlke a copy of the report and this letter. I will be pleased to meet with him at his convenience to further discuss the draft and to arrange further discussions with the authors prior to final drafting.

At this time, NWA is continuing the design and planning activities necessary for construction of the Alaska segment of the Alaska Natural Gas Transportation System based on a design concept of constructing a single gas conditioning facility at Prudhoe Bay and utilizing a buried chilled 48" diameter pipeline, with an operating pressure of 1260 psig, as prescribed by the

Federal Energy Regulatory Commission. We trust that having studied the enclosed report you will concur in the design concept.

Very truly yours,

L D Legg

L. D. Legg

LDL/bjb

Enclosure

cc: C. E. Behlke w/att.

PRELIMINARY DRAFT
For Discussion Purposes Only

DATE

DEC 22 1980

DRAFT #4

ALASKA NATURAL GAS TRANSPORTATION SYSTEM

REPORT ON ALTERNATIVES FOR GAS CONDITIONING FACILITIES LOCATION

For Discussion Purposes Only

DATE

DEC 22 1980

DRAFT #4

FIGURES AND TABLES

PRELIMINARY DRAFT
For Discussion Purposes Only
DRAFT #4
DEC 22 1980

ALASKA NATURAL GAS TRANSPORTATION SYSTEM

REPORT ON ALTERNATIVES FOR
GAS CONDITIONING FACILITIES LOCATION

I. PURPOSE AND OBJECTIVES

The purpose of this report is to provide information to the State of Alaska which documents the decision of the interested parties to plan for location of gas conditioning facilities for Prudhoe Bay at the North Slope of Alaska in the Prudhoe Bay Field. The decision was based on both economic and technical factors related to the requirements for this multi-billion dollar facility, which will be one of the critical elements in the transportation of natural gas from the Prudhoe Bay Field via the Alaska Natural Gas Transportation System (ANGTS) through Alaska, Canada, and ultimately to the "Lower 48" United States consumers.

II. SUMMARY OF CONCLUSIONS

The economics of various plant locations and the attendant pipeline designs have been developed and are discussed in detail in Section IV. The options considered cover technically feasible cases, although the operational safety and life-cycle economics of higher pressure and/or warm, aboveground gas pipelines are somewhat controversial. The technically feasible cases are as follows:

Case A: Conditioning Plant at Prudhoe Bay - 1260 psig chilled gas pipeline: This is the case required for the Federally approved pipeline design specifications, per the Alaska segment.

Case B: Hydrocarbon dew point control at Prudhoe Bay and CO_2 removal in interior Alaska. A 1260 psig chilled gas pipeline is used and all technically possible facilities are located in interior Alaska.

Case C: All gas conditioning is done in interior Alaska; a 455-mile, 675 psig warm gas pipeline is used to minimize facilities at Prudhoe Bay.

Case D: All gas conditioning is done in interior Alaska; a 455-mile, 1680 psig warm gas pipeline system is used, with compression facilities at Prudhoe Bay.

Case E: All gas conditioning is done in interior Alaska; a 455-mile, 2160 psig chilled gas pipeline is used with compression and cooling facilities at Prudhoe Bay.

The estimated costs are shown in Table 1. Costs of all other cases considered exceed the cost of Case A by 584 to 7539 million 1980 dollars. Risk costs have been estimated and do not significantly change the relative comparisons.

It should be noted that risk costs for possible difficulty in achieving completely successful sealifts to Prudhoe Bay were investigated. Relative to the range of contingencies which confront large projects in any location, the seasonal regularity of the Arctic ice movement is very predictable and the normal Prudhoe Bay sealift plan has a high probability of success. Sealifting to Prudhoe Bay has been conducted since 1970. Only one of the 11 sealifts (1975) has encountered serious delay, and the only major cost effect of that delay was additional construction labor at the North Slope. In the Alaska projects, only one barge has sunk -- a load of pipe enroute from Japan in the open ocean. It should be recognized that interior Alaska also presents some construction delay and cost risks: severe winds, ice fog, closure of roads or railways by weather and traffic accidents, unforeseen soil conditions, etc. The estimated contingency costs for these risks are shown on Table 1.

It should also be noted that all the cost estimates for Cases B through E are "order-of-magnitude"; i.e., estimates for these alternative designs were developed in far lesser detail than the Parsons Study and the ANNGTC application to the FERC, which were the source of Case A costs. However, care has been taken to assure that the estimates are comparable and have included consistent contingency and risk costs.

Table 2 shows estimates of operating and maintenance costs for the gas conditioning facilities. Although these estimates are approximate, they indicate that there are not sufficient operating cost savings to provide any economic incentive to incur higher capital costs of conditioning facilities in interior Alaska.

In summary, the preferred option is Case A, using the approved 48-inch 1260 psig pipeline system and a single gas conditioning facility at Prudhoe Bay. This option has the lowest cost, presents minimum technical risk, and provides the lowest relative potential for overall project delay. A detailed discussion of the relative features of each case is presented in Section V.

III. BACKGROUND

A. Gas Conditioning Requirements

To facilitate understanding of the data submitted herein, a general technical background description of the requirements for gas conditioning is furnished.

Petroleum as produced from wells in oil or gas fields is seldom suitable for long-distance transportation. In the case of oil, gas must be separated from the liquid because the further treating or conditioning of the gas or oil is most economically done in separate facilities. Produced water, usually quite salty, must be separated from the oil to avoid the cost of transporting the useless brine to the refinery. Natural gas as produced is also usually unsuitable for transportation for more than short distances because the gas is saturated with water vapor and will form hydrates at modest pressures and moderately low temperatures. The gas may contain toxic constituents such as

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Hydrogen Sulfide, inert gases such as Nitrogen and Carbon Dioxide, and heavy hydrocarbons which will condense and form liquid slugs in a gas transmission pipeline. Facilities built to upgrade raw crude oil or natural gas to a transportable condition are almost always located in the field near the wells because of safety, operability, and economic factors. The early removal of water reduces the risk of corrosion and hydrate formation, a cost and safety risk; and the removal of toxic material makes the transportation operation significantly more safe. Both water vapor and Carbon Dioxide can create operational problems, especially when both are present, and are usually removed to some small acceptable level (typical pipeline requirements are no more than 3 volume percent CO_2 and 7 pounds of water per million cubic foot of gas) prior to long-distance transmission. The transport of material such as Water, Nitrogen, or Carbon Dioxide is also usually avoided because these materials have no energy value and their inclusion in the transportation system increases the cost of transporting the energy-bearing portion of the stream. Also, these diluents can cause end user interchangeability problems (flame-outs, yellow-tips) and safety hazards. Gas and oil conditioning is usually performed in or near the producing <u>field</u> as opposed to other processing operations which are usually accomplished as close to the <u>market</u> as possible, such as upgrading or refining of raw petroleum into motor fuel, fuel oil, and petrochemicals. This is because the transportation of relatively benign raw petroleum is safer, less complicated, and less expensive than the transportation of finished or intermediate products which have a higher value and may be more volatile, flammable, and/or toxic.

B. <u>Prudhoe Bay Conditioning Requirements</u>

The wellhead gas at Prudhoe Bay has been sampled and analyzed extensively, and three constituents which pose transportation problems have been found: first, the gas is saturated with water vapor (as is virtually always the case); second, the gas contains about 13 volume percent Carbon Dioxide; and third, the gas is relatively "rich" in

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heavy hydrocarbons such as butane, pentane, hexanes, and other components which will condense into the liquid phase at proposed pipeline pressures and temperatures.

Figure 1A shows the expected chemical composition of the gas to be produced. Figure 1B shows a "phase envelope" diagram for this gas to illustrate the vapor and liquid regions of this mixture at various combinations of temperature and pressure. The portion of the chart marked "two-phase region" illustrates conditions where both gas and liquid exist, and shows that at conditions contemplated in the current Alaska Gas Pipeline (30°F, 1260 psig maximim, and -10°F, 1000 psig minimum), the raw or "unconditioned" gas would exist in two phases. This condition poses an unacceptable pipeline risk because condensation and the formation of liquid "slugs" will cause damage to compressors and could cause unacceptable mechanical forces as the "slugs," with large mass relative to the gas, are forced to turn or traverse large elevation changes. Two-phase flow is a particular problem in mountainous areas, where liquid "slugs" occur erratically, are unpredictable in size and velocity, and cause unacceptable pipeline flow surges and pressure fluctuations. A serious problem could arise if condensed liquid accumulated in a section of pipeline on the slopes of the Brooks Range and created enough additional hydraulic head to increase pipeline pressure to the maximum allowable value. The end result of two-phase flow would be severely restricted gas flow rate in a given pipeline.

For a chilled gas pipeline to be able to transport unconditioned gas in the single phase region, the gas pressure must be maintained at quite high levels. When the gas enters more conventional pipelines and pressure is reduced to 1000 psig or below, the unconditioned gas would revert to the unacceptable two-phase condition unless the gas is heated. To eliminate the two-phase problem, the "hydrocarbon dew point" of the gas must be lowered, i.e., enough of the heavier constituents of the mixture (such as pentane, hexanes and heavier hydrocarbons) must be removed so that a single-phase gaseous mixture is obtained which will not partially condense at pipeline conditions. This heavy hydrocarbon removal must be done before the gas enters the lower pressure pipeline.

Figure 1A shows that the Prudhoe Bay gas has a Carbon Dioxide content of nearly 13 percent. This high level of Carbon Dioxide poses two problems in pipeline transportation: first, if water should be allowed to enter the pipeline, corrosive carbonic acid would be formed by the solution of Carbon Dioxide in the water; second, the Carbon Dioxide is of no value at the burner tip, but Carbon Dioxide would take up "transportation space" in the pipeline and requires compression just as does the hydrocarbon gas. To transport 2 billion cubic feet per day of raw Prudhoe Bay hydrocarbon gas, approximately 2.3 billion cubic feet per day of the gas-CO_2 mixture would have to be handled.

The removal of water vapor down to some small level must be accomplished both for corrosion control and to inhibit the formation of hydrates, since conditions suitable for hydrate formation exist at virtually all expected pipeline conditions of temperature and pressure. The produced gas is currently dehydrated to a -40°F water dew point at the separation stations prior to being transported 5-8 miles to the in-field central compressor plant for injection back into the reservoir. For pipeline transportation through the ANGTS, dehydration will continue to be required, and if the gas is subjected to any process which resaturates the gas with water vapor, the gas will again have to be dehydrated before it enters the pipeline.

The preceding discussion describes the technical and economic need to "condition" the Prudhoe Bay gas. Data presented in the remainder of the report will illustrate economic comparisons of various pipeline modes and conditioning facility configurations to meet these requirements.

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VI. CASE STUDIES OF ALTERNATE DESIGNS

Case A (Figure 2A)

A. Objectives

Case A is the current plan and comprises the facilities necessary to prepare gas for entry into a 1260 psig pipeline system for minimum temperature and pressure conditions of -10°F and 1000 psig, respectively.

Carbon Dioxide content of the gas must be reduced to an acceptable economic level. For this case, all conditioning facilities are located at Prudhoe Bay. Pipeline configuration is the 1980 ANNGTC filing.

B. Technical Requirements

Condensation in the pipeline must not take place at -10° and 1000 psig; therefore, enough heavy hydrocarbons must be removed so that the phase envelope of the pipeline gas is adequately below this condition. The process described in the 1978 Parsons Report meets this requirement by removing some of the butane and essentially all of the pentanes and heavier hydrocarbons. Figure 1B illustrates the approximate hydrocarbon dew point curve required. Since no economic uses for light NGLs were assumed to exist, the NGL's removed were either reblended into the product pipeline gas or blended with the effluent Carbon Dioxide stream and used as plant and field fuel.

The gas pipeline specifications require an inlet pressure of 1260 psig and a temperature of 30°F.

C. Facilities Description

The conditioning facilities consist of a refrigeration section to cool the gas and cause heavy hydrocarbons to condense, a CO_2 removal section,

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a fractionation section to separate light hydrocarbons from heavy hydrocarbons so that they may be pumped or compressed as required, pipeline gas compressor facilities to increase the pressure to the pipeline inlet requirement, and gas chilling facilities to cool the pipeline inlet gas to the required pipeline inlet temperature of 30°F.

For comparative purposes, the pipeline facilities for this case are those described for the Prudhoe Bay-Fairbanks (near Fox) section (Mileposts 0.0-455.0) of the July 1980 FERC application by the ANNGTC. In summary, the facilities included in this segment are: (a) the Prudhoe Bay receipt metering station, (b) the four northernmost compressor stations between Prudhoe Bay and Fairbanks proposed for the initial 2.0-2.4 Bcf/day throughput, and 455 miles of 48-inch, 1260 psig buried pipeline.

D. General Discussion of Case

1. Costs: this system has the lowest overall costs.

2. Schedule and Permit Approvals: this system complies with the current schedule and ongoing approval process for the Alaska Segment of the ANGTS; the conditioning facility is essentially as described in the FERC final Environmental Impact Statement.

3. Technical Risks: the systems described in this alternate are all based on commercially proven technologies and the Alaska construction and logistics methods have all been used for facilities at Prudhoe Bay.

4. Personnel Requirements:

   a. Plant Operations: about 438 personnel will be required to operate and maintain the facility at Prudhoe Bay. Over 50 catering and contract personnel will be employed to provide

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services for the Owner employees. Presumably almost all of these personnel will reside in Alaska and work a "7 days on, 7 days off" schedule.

b. Pipeline Operations: the operational plan described in the FERC application would be followed.

5. Environmental and Resource Requirements

a. Plant Impacts: In general, the construction environmental impacts will be short-term, and acceptable as described in the EIS. The facility will have minimal liquid effluents and the air emissions will comply with all applicable state and Federal regulations. The blending of Carbon Dioxide into the fuel system will actually reduce NO_x emissions from gas turbines.

Permafrost impacts will be mitigated by the elevated module and pile method used for other facilities at Prudhoe Bay. The increase in manpower, transportation, and logistics resources now used for North Slope activities will be nominally small relative to the current activity.

b. Pipeline Impacts: all impact information on the Alaska Segment of the ANGTS pertinent to the July 1980 FERC application is public knowledge and is directly applicable to this case.

<u>Case B</u> (Figure 2B)

A. <u>Objectives</u>

The objective of this case was to examine the requirements and costs for a system which utilized the approved 48-inch, 1260 psig chilled pipeline system specifications, with the Carbon Dioxide removal facilities located in interior Alaska in the general vicinity of Fairbanks.

"Pipeline quality" gas enters the pipeline at the outlet of the interior Alaska facilities.

B. Technical Requirements

As can be seen in Figure 1B, the raw gas must undergo hydrocarbon dew point control prior to entering the chilled 1260 psig pipeline at Prudhoe Bay for vapor-phase transmission to interior Alaska. Transportation economics through the remaining pipeline system south dictate that Carbon Dioxide should be removed in Alaska to pipeline specifications. The gas must be dehydrated, compressed to 1260 psig, and cooled to 30°F following the CO_2 removal process.

C. Facilities Description

At Prudhoe Bay, a "straight" refrigeration system is used to cool the gas and liquify the components which are not transportable in the pipeline. The gas has been dehydrated in the field, so no further water removal is necessary. The recovered NGL is fractionated into pentanes and heavier components which can be transported in TAPS and the lighter material is used for fuel. The gas is compressed to pipeline pressure and chilled to 30° prior to entering the pipeline. The pipeline distance to the interior facilities is assumed to be 455 miles, which would locate the interior facility in the vicinity of Fox. This location is assumed only for the comparative perspective of this report; it appears to be as feasible as any other interior site for comparative purposes.

At the interior plant location, gas must be warmed to about 90°F to avoid hydrate formation in the Carbon Dioxide removal system. The gas is contacted by an aqueous "chemical" solvent to remove CO_2. This type of process must be used to minimize the hydrocarbons absorbed and released with the CO_2, since no means are available to conserve hydrocarbons mixed with CO_2 (such as the field fuel system at Prudhoe Bay). The CO_2-free gas is dehydrated to an acceptable

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water content, compressed to 1260 psig, and chilled to 30°F for entry into the pipeline.

To transport the incremental volume of CO_2, approximately 38,000 horsepower of additional compression and 14,000 tons of additional refrigeration would be distributed among the four pipeline compressor stations between Prudhoe Bay and Fairbanks. Additional metering facilities would be required at Prudhoe Bay and probably at the interior CO_2 removal site, but the costs of these facilities were neglected for purposes of this comparative analysis.

D. <u>General Discussion of Case</u>

1. Costs: the system costs are adversly impacted primarily because of additional investment costs and operating costs resulting from separation of the two gas conditioning functions within the resulting duplication of facilities and manpower. An additional 38,000 horsepower of compression would be required to transport the higher gas volumes to interior Alaska because of the CO_2 content.

2. Schedule and Permit Approvals: the ANGTS project schedule would likely be impacted by the time required for approval of the additional compression facilities. The simultaneous construction of two conditioning facilities in two locations would result in large Alaska construction labor requirements which could impact the schedule.

   The FERC EIS does not address two conditioning plant locations constructed and operated simultaneously, and would have to be supplemented with the resulting project delay.

3. Technical Risks: the principal technical risks are the high corrosion potentially associated with transporting gas containing 13 percent CO_2. If water should enter the pipeline from hydrotesting, repairs, or from future tie-ins, and if gas contacts unfrozen

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water, serious corrosion would likely occur. Repairs would probably involve system shutdowns.

A second risk is the operational difficulty of coordinating two facilities which are 455 miles apart, and which must both function effectively and simultaneously to condition the gas.

4. Personnel Requirements:

   a. The facility would require 434 operating, maintenance, and support personnel. Obviously optimum manning could not be achieved with the facilities separated. Some duplication of supervision and maintenance personnel would be required, offset by the lower manpower requirement in interior Alaska.

   b. Construction and operating personnel for the Prudhoe Bay-Fairbanks pipeline segment could increase slightly, but not significantly over Case A. Such changes are considered negligible for purposes of this comparision.

5. Environmental and Resource Requirements: the energy, land use, gravel, and other material resource facilities would exceed those of a single facility; therefore, the environmental impact would undoubtedly be greater. Pipeline compressor stations (north of Fox) would contain more powerful equipment, but environmental impacts and resource requirements over Case A would be minor.

Case C (Figure 2C)

A. Objectives

The objective of this case was to examine the requirements and cost for a system which had all gas conditioning facilities located in interior Alaska (in the general vicinity of Fairbanks). The raw gas would be transported to the interior Alaska site for carbon dioxide removal and hydrocarbon dew point control as required for transport in the system south of the conditioning facility.

B. <u>Technical Requirements</u>

The raw gas from the producer gathering system is at a temperature of about 90°F and has a hydrocarbon dew point of about 88°F; the gathering facilities operate at approximately 600 psig. The concept for this alternative is to apply these gathering system design parameters to the Prudhoe Bay-Fairbanks pipeline segment. This necessitates a low pressure, warm gas pipeline adequately insulated to maintain temperature under cold weather shutdown conditions. Above and below ground pipeline construction modes must be employed to cater to permafrost soil conditions (similar to TAPS).

C. <u>Facilities Description</u>

Two 56-inch insulated pipelines are required to provide the nominal 2 BCF/day flow rate at 675 psig maximum operating pressure. The lines are installed on vertical support members and the configuration would be trapezoidal similar to TAPS to provide for thermal expansion and contraction. A new crossing will likely be required at the Yukon River, since the space for two additional lines, each with a total diameter of 5'8", including insulation, probably does not exist. No design exists for this crossing; the cost, although significant, has not been estimated for this study and is omitted. Heating stations must be provided at the compressor stations to maintain gas temperature above the hydrocarbon dew point. Metering facilities and a heating station are provided at Prudhoe Bay.

At the interior Alaska plant site, hydrocarbon dew point control facilities, CO_2 removal facilities, compression facilities for 1260 psig, and gas chilling equipment are installed. The functions are similar to those

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required for Case A, except that a chemical solvent CO_2 removal process and subsequent dehydration, as described in Case B, are required. All NGL removed for gas conditioning would be either burned as fuel or pumped into TAPS. Spare fractionation facilities are required, because at this location no NGL emergency injection (during fractionator outages) into the Prudhoe Bay reservoir is feasible.

D. Advantages and Disadvantages

1. Costs: the costs of this system are extremely high because of the following:

   a. Two 455-mile, 56-inch aboveground pipelines are required.

   b. Plant construction in interior Alaska precludes large modular installations.

   c. Additional facilities for NGL disposition are required.

   d. Additional fuel is required for heating and compression on the pipeline.

2. Schedule and Permit Approvals: a minimum of two years of delay in the ANGTS schedule would occur because of the radical change in pipeline construction and cost from that envisioned in the FERC filing and other Federal applications. A total pipeline redesign effort would contribute to the delay. The number of pipe mills capable of rolling 56-inch pipe in the quantities required is limited. The additional pipeline construction alone would probably require at least one year more than the current schedule. Re-evaluation of the current right-of-way stipulations and TAPS proximity criteria would likely be required.

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3. Technical Risks: the technical risk of transporting raw gas and relying on pipeline heaters to maintain the line above 100°F at all possible flow rates and under all weather conditions in the Arctic is formidable. Additionally, the corrosion risk described in Case B is present, except that at normal operating temperature unfrozen water will exist, making corrosion damage in the CO_2-rich portion of the line inevitable if water enters the line.

4. Personnel Requirements:

   a. Personnel requirements are 362 operating, maintenance, and support personnel required at the Interior Alaska plant site.

   b. The two aboveground pipelines will require considerably more maintenance and surveillance personnel than would be the case with a single buried line. No estimate has been made of these requirements, because none is believed to be warranted due to the cost and schedule infeasibility of this alternative.

5. Environmental and Resource Requirements: the gravel resource requirements will be higher than the buried line cases since more and larger workpads for the pipelines would be necessary. Fuel consumption is higher than other cases because of the heated low pressure line. Construction and operational impacts will be significantly greater than those of a single buried line.

<u>Case D</u> (Figure 2D)

A. <u>Objectives</u>

Case D examines the facilities necessary to compress raw field gas to a 48-inch, 1680-psig, aboveground, warm pipeline system which will handle the gas with no conditioning required at Prudhoe Bay. Carbon

Dioxide removal and hydrocarbon dew point conditioning facilities are located in interior Alaska to prepare the gas for transmission to a 1260-psig buried, chilled line.

B. Technical Requirements

The raw gas must be compressed to 1680 psig and cooled to about 100°F at Prudhoe Bay. The pipeline gas must not be allowed to cool below about 70°F at the compressor suction pressure of 1450 psig. Aboveground pipeline construction with insulated pipe is required. Gas conditioning to prepare the gas for transport in the chilled 1260 psig pipeline is required at the interior Alaska location.

C. Facilities Description

Facilities for compression of the raw gas to 1680 psig and cooling to about 90°F are required at Prudhoe Bay. The 1680 psig pipeline is insulated and installed aboveground in the same manner as TAPS between Prudhoe Bay and the interior Alaska location. At the interior Alaska site, hydrocarbon dew point control is achieved by chilling the gas, and liquids recovered are fractionated and put into TAPS or used for fuel. Carbon Dioxide removal to pipeline specifications is accomplished in a "chemical" solvent process as in Cases B and C except that the conditioning is accomplished at a pressure adequate to provide entry into the 1260 psi pipeline without compression. The gas must be dehydrated, and chilling to 30°F is required prior to entry into the 1260 psi buried, chilled pipeline.

D. General Discussion of Case

1. Costs: the costs are increased by the use of an aboveground pipeline between Prudhoe Bay and the higher construction costs in interior Alaska.

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2. Schedule and Permit Approvals: the change in the pipeline design pressure and revisions to the design and construction method would likely delay the ANGTS schedule by at least one year. The change to an aboveground line would likely require re-evaluation of the current right-of-way stipulations and TAPS proximity critieria.

3. Technical Risk: as described in Cases B and C, the major technical risk is the corrosion potential attendant to transport of raw gas containing 13 percent CO_2.

4. Personnel Requirements:

   a. Personnel requirements are 440 operating, maintenance, and support personnel.

   b. Pipeline personnel requirements would be marginally larger than Case A because of the additional security, inspection, and maintenance for an aboveground pipeline system.

5. Environmental and Resource Requirements: the necessity for a large compression facility at Prudhoe Bay as well as a plant in interior Alaska would increase gravel requirements over a single facility case. The installation of a second aboveground pipeline essentially parallel to TAPS would create additional impacts because of the construction mode.

Case E (Figure 2E)

A. Objectives

The objective of this case was to examine the use of a 42-inch 2160 psig chilled, buried pipeline to transport unconditioned raw Prudhoe Bay gas to an interior Alaska gas conditioning site, and conditioning for a 1260 psi pipeline at the interior site.

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B. Technical Requirements

No gas conditioning is required at Prudhoe Bay, although the gas must be compressed to 2160 psi and chilled to 30°F for the buried line. At the interior Alaska site, CO_2 must be removed and hydrocarbon dew point controlled to allow the gas to be transported in the subsequent 1260 psi chilled line and through the Canadian line segments.

C. Facilities Description

The compression equipment at Prudhoe Bay entails about 200,000 horsepower and the gas must be chilled to 30°F prior to entry into the pipeline. At the 2000 psig ~~opening~~ operating pressure level, a 42-inch diameter pipeline is adequate to transport the required 2.3 Bcf/day of raw gas. At the interior Alaska site, 455 miles from Prudhoe Bay, hydrocarbon dew point control is achieved by chilling utilizing the available pressure drop in expanders. CO_2 is removed by a chemical process similar to Cases B, C, and D. Gas is dehydrated, recompressed to 1260 psi, and chilled to 30°F for entry into the pipeline system.

D. General Discussion of Case

1. Cost: although a smaller diameter pipeline can be used in this case, costs are somewhat higher than Case A because of the greater pipe wall thickness and the higher working pressure valves, fittings, compressors, coolers, etc. Equipment of the large size and high pressure rating which meets Arctic service requirements would likely require some developmental engineering; while these development costs could be significant, none are included in the estimated cost of this case.

2. Schedule and Permit Approvals: since this pipeline system represents a major scope change from the current plan, new approvals would be required from the FERC, DOI, and the Federal Inspector.

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Questions were raised during regulatory hearings relative to the advisability of operating a 1680 psig pipeline, and similar questions would have to be answered for a 2160 psig line. Such resolutions would result in an ANGTS schedule delay of at least one year and probably more.

3. Technical Risks: the transport of gas containing a high concentration of CO_2 would pose the same corrosion risk described in Cases B, C, and D. Pipeline qualification burst tests and/or a decision to utilize crack arrestors would have to be implemented to minimize ductile fracture risk. Large valve and fitting manufacture to the required quality specifications would have a greater risk of schedule delay than more common pressure requirements, and operating risks would be higher due to extensive use of unproven equipment.

4. Personnel Requirements:

   a. Plant personnel requirements are 422 operating, maintenance, and support personnel.

   b. Pipeline operating requirements would be essentially the same as Case A.

5. Environmental and Resoucc Requirements: The construction of facilities at both Prudhoe Bay and at the interior Alaska site would consume more gravel and other resources than a single site.

V. COMPARISON OF ALTERNATES

A. Costs

Investment costs are summarized on Table 1. The costs (excluding risk) of various alternates involving conditioning facilities at an interior Alaska location are estimated to exceed those of the current plan (Case A) by $617 million (for Case B) to $7,597 million (for Case C). When "risk" costs for additional contingency to cover serious weather and construction difficulties are added, the current plan is less costly by $584 million (Case B) to $7,539 million (Case C). These costs are in 1980 dollars and do not include the escalation to the actual date of expenditure. If the regulatory and design delays inherent as a result of major changes to the design were considered, Case B, the next lowest cost alternative to Case A would involve a minimum of one year of delay for the entire ANGTS, including sections of the line in Canada. This delay cost could exceed $2 billion for a $20-billion system at an inflation rate of 10 percent. Cases C, D, and E would undoubtedly involve delays exceeding one year with the commensurate delay costs.

Fuel usage for Case A is estimated to be 10% to 30% lower than the other cases. Case B, the next investment case, would have a fuel cost penalty of about $23 million per year. This is the result of the necessity for using the less efficient "chemical" CO_2 removal process, the additional pipeline fuel required to transport the CO_2 to the interior conditioning plant location, and the restricted availability of integrated heat conservation options when the hydrocarbon dew point and CO_2 removal facilities are at separate locations.

B. Approvals and Permits

A comparison of the deviations from current temporary certificates and other approvals indicates that Case B is the least changed from the current basis; however, Case B still has two important differences: first, the transportation of up to 300 Mmcf per day of CO_2 with the gas creates compressor station design changes with attendant investment and operating cost increases, and creates a technical concern with respect to corrosion potential. The cost-effectiveness of design changes and higher operating risks to transport valueless CO_2 could

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be challenged during the regulatory approval process. Second, the construction and operation of two conditioning facilities at two locations presents additional environmental impacts over a single installation not addressed in the FEIS. Other cases also have obvious design differences which would delay approvals.

C. Schedule

The overall ANGTS schedule delay implications of the cases which differ from the current design are mentioned in A and B, above. In general, the relative schedule delays are conservatively estimated as follows:

Case	Minimum Delay	Comment
A	None	Current plan
B	1 year	Design changes
C	2 years	Design changes & additional construction time
D	1 year	Design and construction changes
E	1 year	Design and technological scope changes

D. Technical Factors and Risks

All cases except Case A pose the problem of mitigating the potential corrosion risk attendant with transport of high CO_2 gas. This risk is most serious in the "warm" pipeline Cases C and D, because water can exist in the liquid form inside the pipeline at normal operating temperatures.

The cases with major facilities at two locations, Cases B, D, and E pose potential operating coordination problems. Case B is the worst example because the hydrocarbon dew point control is done 455 miles from the CO_2 removal. Operational problems at Prudhoe Bay could adversely affect the operation of the CO_2 removal plant at the interior location. "Early warnings" of impending "upsets," or other problems

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are essential to good plant operations and long distances and decentralized operations make this procedure more difficult. The 455-mile separation of facilities would undoubtedly compound operating difficulties.

Cases C and D require aboveground pipeline operations to be conducted at temperatures above the hydrocarbon dew point. The problems of maintaining warm line temperature during an extended shutdown, and/or the disposal of large quantities of condensed NGL in the pipeline would have to be resolved, and would involve costs not considered in these estimates.

The risk of a ductile fracture incident and the potential for damage would be significantly greater for the aboveground construction. The likelihood of fracture initiation would be increased by the higher exposure to either accidental or deliberate external damage.. A propagating fracture could disengage the pipe from the supports and thrust forces caused by the escaping gas could cause relatively long sections of the pipeline to "whip," with resulting damage to the gas pipeline and its supports and possibly to TAPS.

VI. OTHER FACTORS

A. NGL Recovery

As stated previously, the gas from the Prudhoe Bay Sadlerochit reservoir is relatively "rich" in high molecular weight hydrocarbon compounds. These compounds are sometimes extracted from natural gas becausr their separate end-product values may be greater than their heating value in a natural gas-fuel mixture. The compounds are essentially all those components heavier than methane, such as ethane, propane, isobutane, normal butane, etc. They are sometimes called "NGL" (natural gas liquids), or "LPG" (liquified petroleum gas consisting of propane and/or butane), or "Natural Gasoline" (usually pentanes and heavier compounds). It was previously pointed out that some of

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the "NGL" hydrocarbons must be removed to enable safe pipeline transportation. After bulk extraction these compounds (pentanes-and-heavier, plus some of the butane) may be separated and the lower vapor pressure liquids shipped with the crude oil and the higher vapor pressure liquids utilized as fuel. The "lighter" NGL components (ethane and propane) are transportable in the gas and can always be extracted if the necessary economic driving force exists (i.e., if their value as separate commodities is sufficiently higher than their heating value sold as "gas" to justify the required investment and operating costs for the extraction and disposition facilities). NGLs may temporarily be used for fuel or temporarily left in the gas, but are never really "lost." They are simply moving to the use which provides the highest economic value. This situation is reasonably analagous to the refining of crude oil into various products such as motor gasoline, jet fuel, or heating oil; the economics of specific product supply/demand balances dictate the product disposition, and that disposition may change from time to time.

B. Alternate Plant Fuels

It has been suggested that the use of Alaska coal should be considered as fuel for ANGTS gas conditioning facilities in order to release more gas for market. The critical factors in addressing this question are security and reliability of supply, economics, and environmental considerations.

The security of a fuel supply which involves another separate business is far more fragile than a fuel supply which is irrevocably tied to the primary operation; for this reason most field petroleum operations utilize their own production for fuel and most gas pipelines utilize their own gas for fuel. The risk of coal shortages due to production or transportation strikes or weather interruption is relatively high. The ANGTS is unlike electrical power grids having alternate power sources where occasional outages of one or more sources may be tolerable. Prudhoe Bay Gas may represent as much as 10 percent of the gas supply for the U.S. and is essentially irreplaceable in the event of an interruption. Very serious complications would result from an interruption of this gas supply, and the risks of interruption would increase significantly if the ANGTS operation were dependent on coal.

Economics are a factor when costly long distance transportation of fuel by road or rail is required. Employing the transport gas for fuel provides the lowest cost method of moving fuel from the source to the use. A further economic advantage of using gas for fuel is that it is the best engine fuel for low maintenance costs and long life of the gas and refrigeration compressor drivers.

The environmental problems of a coal-fired steam power plant are well-known: particulate, NO_x, and CO emissions into the air, water usage, and liquid effluent disposal. Whether at Prudhoe Bay or at an interior Alaska location, the construction of a coal-fired power plant which meets today's environmental requirements would pose an economic and schedule burden of serious magnitude. These would be even more serious if the power plant emissions impacted the air quality of a Non-Attainment Area.

It is concluded that the interests of the the system owners and the gas consumers are best served by the use of gas for fuel. This system has the lowest investment, the least risk of interruption, and the least environmental impact.

C. Construction Considerations

The construction mode considered in this study for Prudhoe Bay gas conditioning facilities is that utilized for Prudhoe Bay producing facilities. Large equipment modules weighing up to 2000 tons are constructed in a "lower 48" coastal facility suitable for loading onto sea-going barges. Approximately 75 percent (or more, if possible) of the construction labor manhours would be incurred at the lower 48 site.

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Modules are transported to Prudhoe Bay by barge and crawlered to a prepared location. The annual sealifts arrive in Prudhoe Bay in August, and the outdoor construction is completed during August, September, October and November to utilize the better weather. The remaining indoor construction work for hookup, detail finishing, and testing can then be done essentially indoors during the winter. As a result, most of the Prudhoe Bay work is done either outdoors in reasonable weather or indoors during inclement periods.

The Fairbanks work was assumed to be accomplished by packaging the equipment onto small skids which can be transported overland by rail and/or truck to interior Alaska. Because of the size and weight limitations, only a limited amount of packaging can be accomplished prior to shipment. Similar packaging concepts are frequently incorporated into inland facilities in the lower 48. It was assumed that some weather protection can be achieved by air-pressurized enclosures and early construction of buildings. The construction costs are higher in interior Alaska because of several factors: (1) more labor is required in Alaska at higher wage rates, (2) more low-efficiency outside work is required for the land transportable packages, because in most cases enclosures cannot be transported with the equipment and must be built at the site, with increased labor and material costs, (3) full achievement of scale economy is precluded because the transportation constraints require smaller process vessels than may be used in large modules, with the attendant additional piping, valves, instrumentation, foundations, etc., and, (4) schedules become more complex when constrained by interior Alaska weather and transport requirements resulting in increased construction time.

It has been suggested that weather factors in a central Alaska location would allow more efficient construction. This was not the experience on TAPS. Pump Stations 3 and 4, north of the Atigun Pass, cost essentially the same as Pump Stations [illegible] 9, and 12 which are in central and southern Alaska.

Further, it is felt that the construction mode utilized for Prudhoe Bay has minimized the impact of outdoor bad weather construction by utilizing very large pre-enclosed facilities to mitigate the effects of the harsh climatic conditions on construction labor efficiency. The experience in the complexities of construction by the producers at Prudhoe Bay and by TAPS -- in Alaska and on very large projects -- has illustrated that simplistic approaches to project cost estimates, cost comparison of alternatives, construction planning, and project execution are not adequate. Estimates using only factored Gulf Coast or other lower 48 construction costs may contain very large margins for error.

D. Personnel and Operating Factors

The construction cost estimates assume that a peak of about 800-1000 construction workers would be employed at the Prudhoe Bay location for Case A, and peak labor forces 25 to 50 percent higher in the cases with facilities in interior Alaska. For operations, about the same number of operating positions and employees would exist for Cases A, B, D, and E, since the cases with split facilities would incur some duplication of positions offset by the lower manpower requirement in interior Alaska.

In the cases with facilities in interior Alaska, the savings for operating costs of wages, camps, and catering are offset by the insurance and tax burdens of the higher capital costs.

Pipeline operating costs have not been calculated for this study, but directionally the higher capital cost cases would have higher operating costs. Cases C and D which transport unconditioned gas will undoubtably have significantly higher pipeline costs because of the corrosion potential and the additional equipment required for heating, slug catching, and liquid disposal.

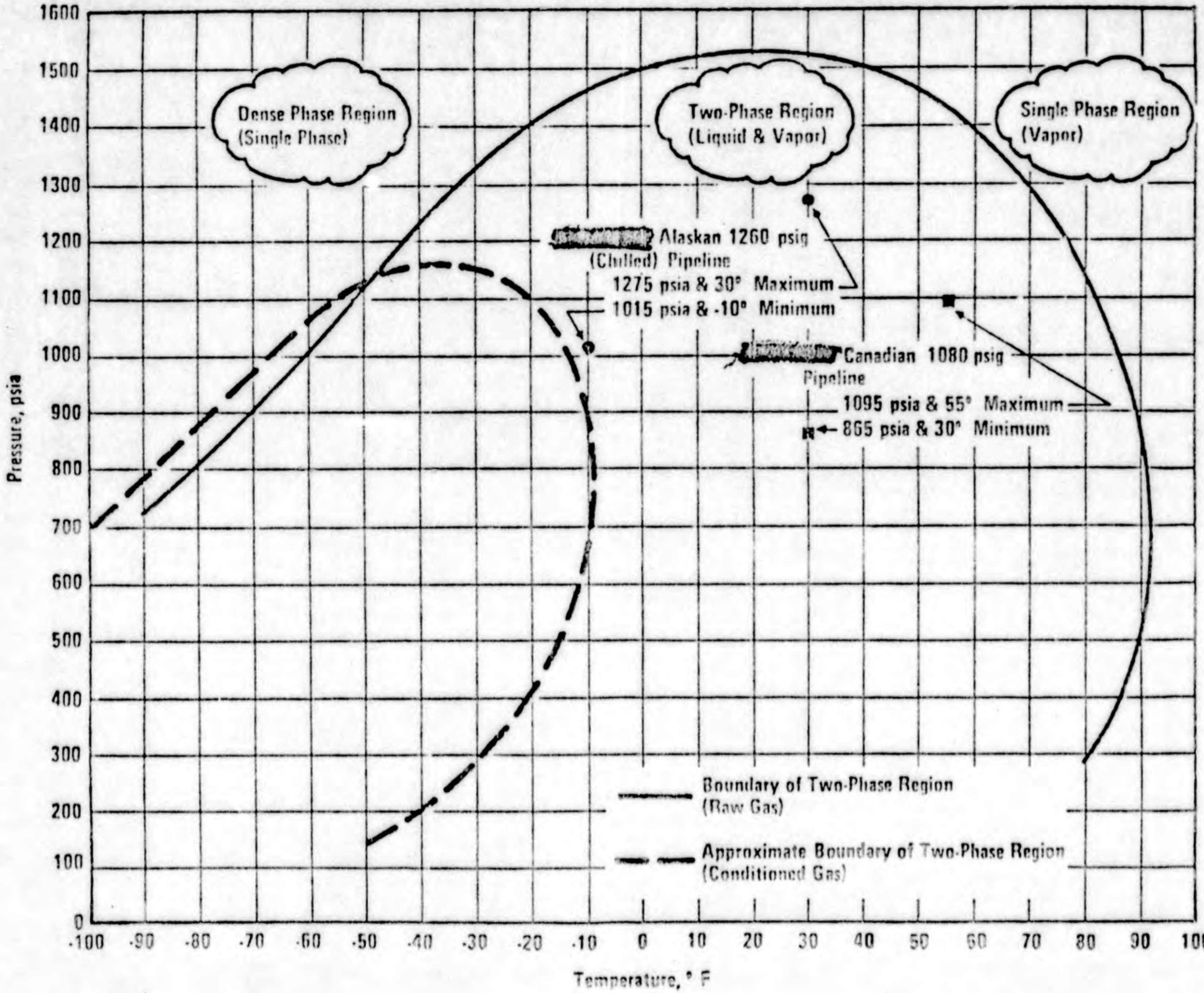
FIGURE 1A

Prudhoe Bay Raw Separator Gas Composition

Component	Mole Percent
Hydrogen Sulfide	(8 ppmv)
Carbon Dioxide	12.63
Nitrogen	0.47
Methane	74.18
Ethane	6.47
Propane	3.48
Isobutane	0.49
Normal Butane	1.17
Isopentane	0.26
Normal Pentane	0.49
Hexanes	0.19
Heptanes	0.12
Octanes	0.04
Nonanes	0.01

Figure 1-B

PHASE ENVELOPE FOR PRUDHOE BAY (SADLEROCHIT) ASSOCIATED GAS



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DATE
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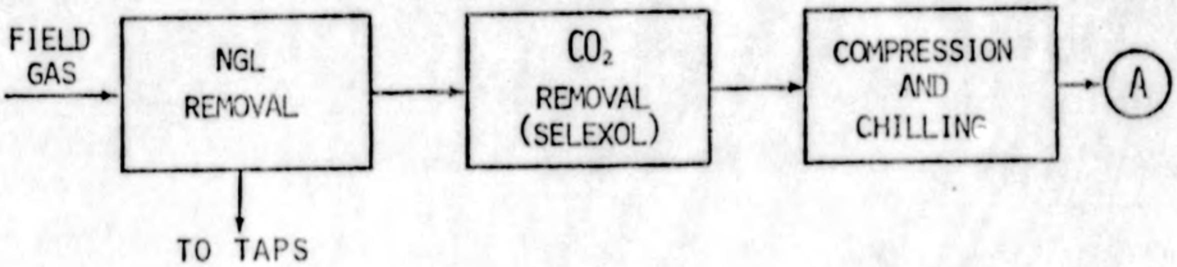
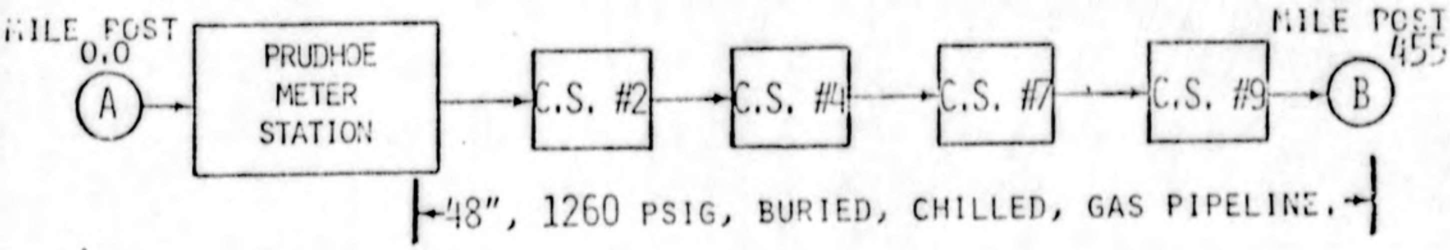
FIGURE 2A

CASE A: ALL GAS CONDITIONING AT PRUDHOE BAY
PRESENT PIPELINE SYSTEM DESIGN

PRUDHOE BAY FACILITIES



PRUDHOE BAY - FAIRBANKS PIPELINE SEGMENT



INTERIOR ALASKA FACILITIES

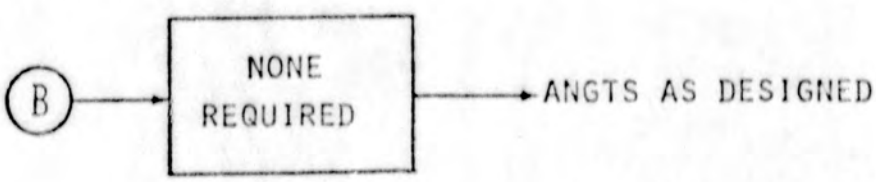


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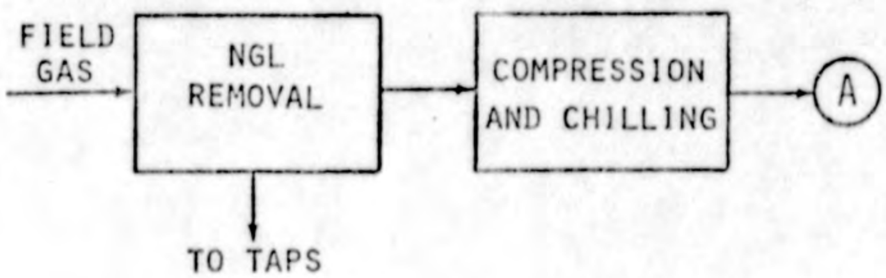
FIGURE 2B

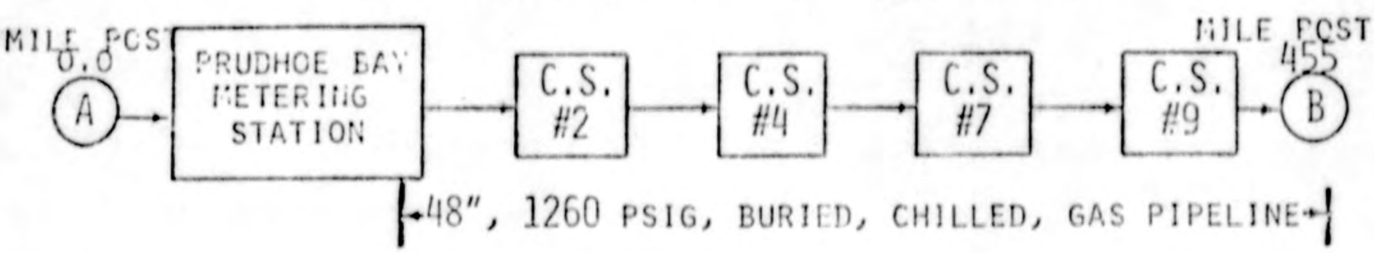
CASE B: CO_2 REMOVAL IN INTERIOR ALASKA

ADDITIONAL PIPELINE COMPRESSION
TO TRANSPORT 13% CO_2 TO FAIRBANKS

PRUDHOE BAY FACILITIES



PRUDHOE BAY - FAIRBANKS PIPELINE SEGMENT



INTERIOR ALASKA FACILITIES

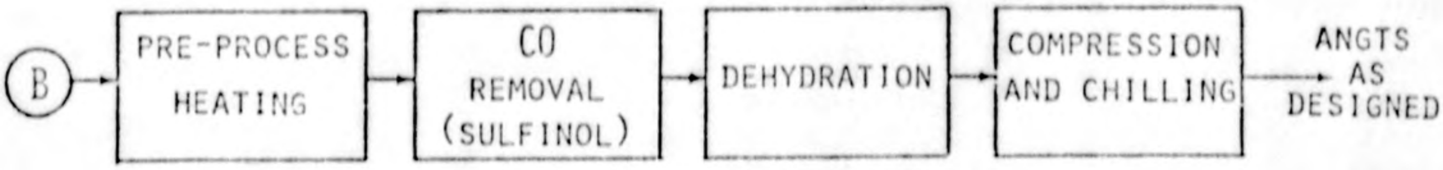


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FIGURE 2C

CASE C: ALL GAS CONDITIONING IN INTERIOR ALASKA LOW PRESSURE, WARM GAS PIPELINE TO FAIRBANKS

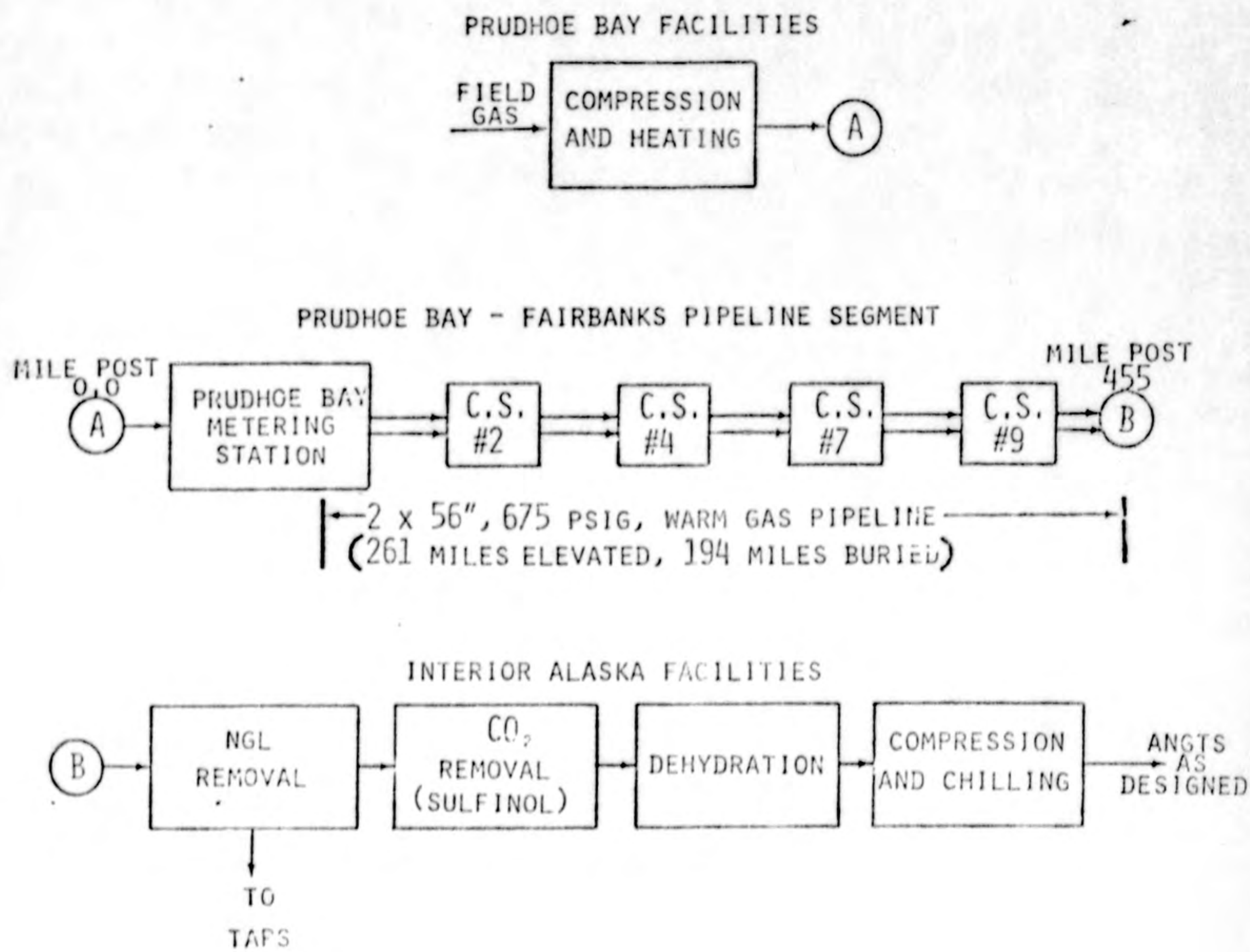

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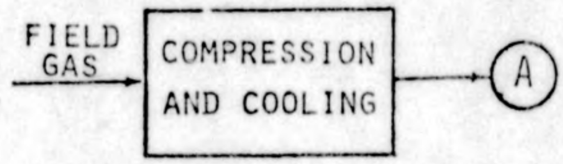
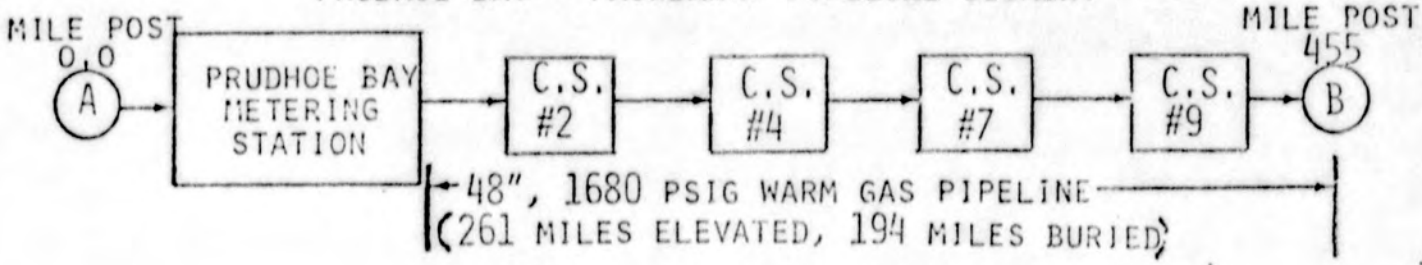
FIGURE 2D

CASE D: ALL GAS CONDITIONING IN INTERIOR ALASKA
HIGH PRESSURE WARM GAS PIPELINE

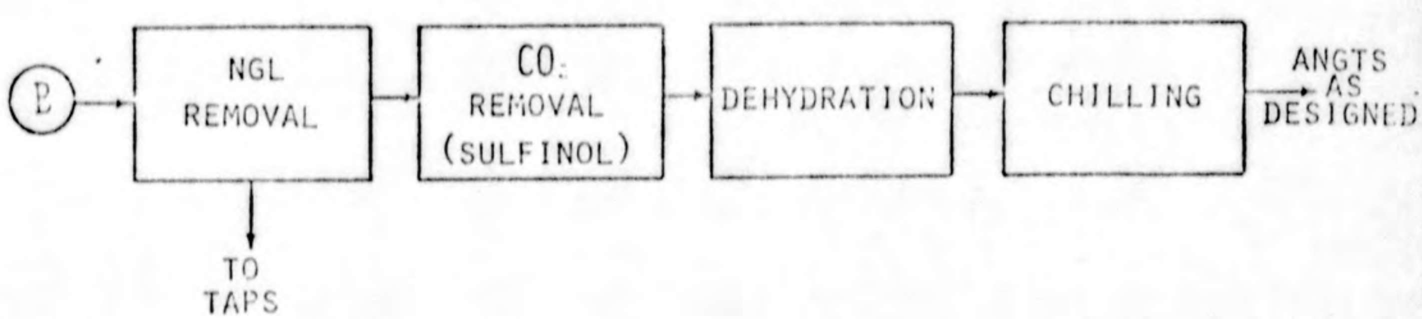
PRUDHOE BAY FACILITIES



PRUDHOE BAY - FAIRBANKS PIPELINE SEGMENT



INTERIOR ALASKA FACILITIES



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DATE
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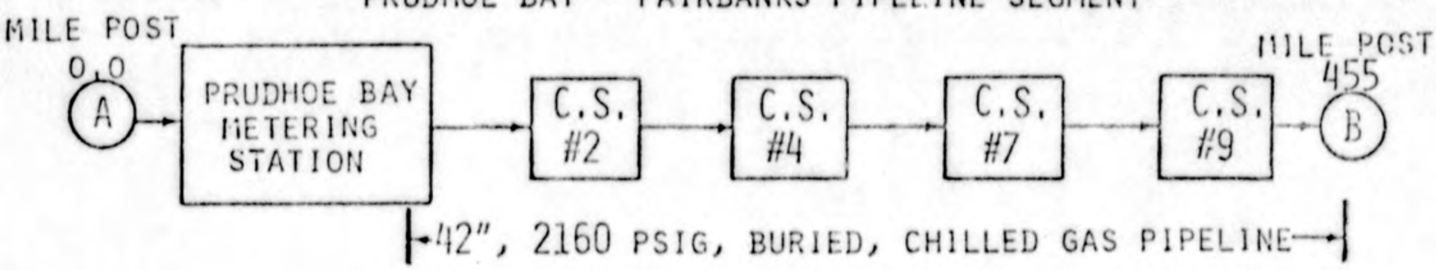
FIGURE 2E

CASE E: ALL GAS CONDITIONING IN INTERIOR ALASKA
HIGH PRESSURE CHILLED GAS PIPELINE

PRUDHOE BAY FACILITIES

PRUDHOE BAY - FAIRBANKS PIPELINE SEGMENT



INTERIOR ALASKA FACILITIES

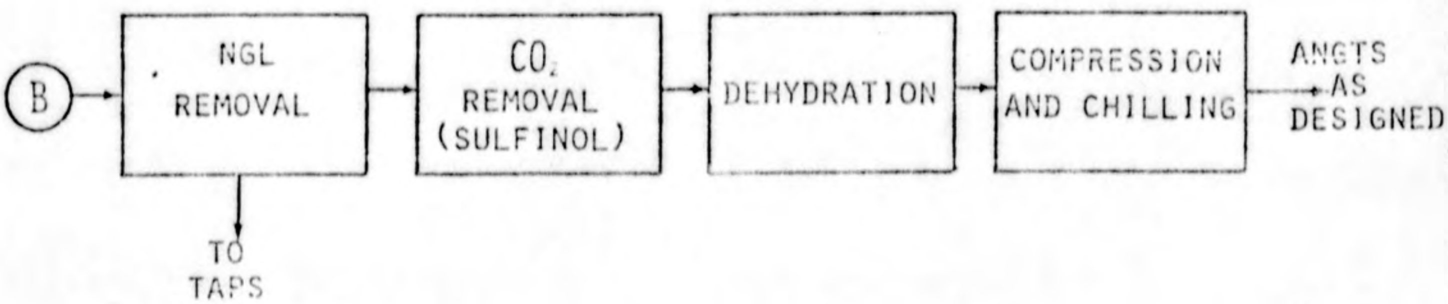


TABLE 1

Summary of Estimated Investment Costs and Fuel Consumption

	Cases				
	A	B	C	D	E
Investment Costs, $ Million (1980)					
Prudhoe Bay Facilities	2300	1100	-	360	585
Pipeline Segment (455 miles)	5250	5317	11717	6214	5495
Interior Alaska Facilities	-	1740	3430	3025	2945
Subtotal Investments	7550	8167	15147	9599	9231
Risk Costs - Prudhoe Bay	230	110	-	36	59
Risk Costs - Interior	-	87	172	151	147
Subtotal Risk Costs	230	191	172	187	206
Grand Total Investment Costs	7780	8364	15319	9786	9437
Fuel Consumed, Billion Btu/day					
Prudhoe Bay Facilities	93	66	-	43	59
Pipeline Segment	15	19	18	19	12
Interior Alaska	-	55	102	68	66
Total	108	140	120	130	137
Estimated Fuel Value, $ Million (1980) Per Year[1]					
Total	74	97	83	90	94

[1] Based on current (July 1980) NGPA allowed field gas price for Alaska. Does not include costs of transportation to interior Alaska other than pipeline fuel, in Cases B, C, D, and E.

TABLE 2

Summary of Estimated Operating and Maintenance Costs and Manpower Requirements for Alaska Gas Conditioning Facilities

	Cases				
	A	B	C	D	E
Operating Costs, $ Million (1980) Per Year					
Operations Labor[1]	19.0	18.3	14.9	18.6	17.4
Maintenance Labor[1]	31.6	29.3	23.4	28.4	28.3
Maintenance Materials	45.1	47.1	45.1	46.0	46.2
Chemicals and Supplies	1.9	2.0	1.9	1.9	2.0
Camps and Catering	8.9	5.4	-	2.7	2.7
Taxes and Insurance[2]	46.0	55.0	69.0	68.0	71.0
Overhead Above Field Level	21.2	22.3	21.2	21.8	22.0
Total	173.7	179.4	175.5	187.4	189.6
Total Field Personnel Employed	438	434	362	440	422

1 Prudhoe Bay costs based on 12-hour shifts working "seven and seven" schedule. Interior Alaska costs based on 8-hour shift without camp, catering, or travel expense.

2 Based on 2 percent of total investment cost.

PLEASE NOTE: THE PRECEDING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT.

D.C. Circuit Dismisses Complaint by Earth Resources Co. and State of Alaska Against FERC Order Approving Design and System Capacity of Alaskan Segment of ANGTS

On 1/3/80 the U.S. Court of Appeals for the D.C. Circuit dismissed complaints filed on 10/5/79 by Earth Resources Co. and the State of Alaska et al. seeking reversal of the FERC's order of 8/6/79 granting application of Alaskan Northwest Natural Gas Transportation Co. (CP78-123 et al.) to approve the design specifications set forth in the President's Decision of September 1977 for the Alaskan segment of the Alaska Highway Pipeline Project. The Court concluded that it does not have jurisdiction to rule on these complaints because the Alaskan Natural Gas Transportation Act prevents review of the Commission's decision for reasonableness or substantial support on the record. Earth Resources Co. of Alaska v. FERC, No. 79-2191 and State of Alaska et al. v. FERC, No. 79-2193.

Specifications for Canadian portions of the ANGTS were adopted by the NEB in February 1978. Over objections of the U.S. Government which favored a 48-inch, 1680 psig system, the NEB prescribed a 48-inch, 1260 psig system from the Alaskan-Yukon border to Whitehorse (as proposed by the project sponsors) and selected a 56-inch, 1080 psig system for the segment from Whitehorse, Yukon to Caroline, Alberta to accommodate reserves from the Mackenzie Delta region (which may be attached to the ANGTS via construction of the so-called Dempster lateral).

In its request that the Commission approve the design specifications in the President's Decision, Alaskan Northwest said it had considered various alternatives and concluded that the system as originally proposed by Alcan Pipeline -- 48-inch with a pressure of 1260 psig -- is the best economic selection for delivery of volumes up to 3.4 Bcf/d. There has been no new evidence, Alaskan Northwest observed, showing that that level of expansibility above the expected 2.0 Bcf/d delivery rate from presently proven reserves is inadequate.

In its 8/6/79 order granting Alaskan Northwest's application, the Commission stated that the President had decided that the diameter of the pipeline will be 48 inches, and created "a predisposition" in favor of the 1260 psig system by his statement that the approved facilities were those included in the Alcan proposal -- which involved operations of a pipeline at a maximum pressure of 1260 psig. Furthermore, the Commission noted, the President also said that the capacity of the system should be adequate for an average daily throughput of up to 2.4 Bcf/d and up to 3.2 Bcf/d with increased compression. Such requirements, the Commission observed, would be satisfied by a combination 48-inch pipe and 1260 psig pressure.

The FERC emphasized the importance of the operating pressure because of its relationship to the ability of the gas stream to carry natural gas liquids. In this connection, the Commission noted concern by the State of Alaska and Earth Resources about the ability of the gas stream to carry gas liquids in view of the state's desire to preserve the option of developing a world-class petrochemical industry. Alaska emphasized that an operating pressure of 1260 psig -- in conjunction with other factors such as the standard for carbon dioxide content and the type of process used for its removal -- could preclude development of a petrochemical industry in Alaska. The state also questioned the location of a conditioning facility at Prudhoe Bay, contending that alternative sites should be given serious consideration.

The Commission, however, stated that it preferred to consider the complex liquids carrying issue in the context of deciding the carbon dioxide standard in the RM78-12 rulemaking proceeding (involving incentive rate of return for the project). Moreover, the Commission added, consideration of that issue in this proceeding would result in delay of the project because it is an essential predicate to refining the design of the system which, in turn, has a direct bearing on the sponsors' ability to prepared detailed cost estimates and obtain financing. "Thus, a delay in determining the pressure could have serious and wideranging consequences in delaying the entire project." (See REPORT NOS. 1146, pp9-11; 1156, pp15-16; 1199, pp32-33; 1231, pp11-12; 1240, App. p39.)

In its decision, the D.C. Circuit asserted that this case "reflects exactly the sort of time-consuming challenge that Congress designed ANGTA to prevent during the approval and construction of the Alaska Natural Gas Pipeline. The order challenged here is nothing but a routine step in the process of issuing a certificate for construction of the pipeline. It is clearly within the Commission's statutory authority and offers no basis for a challenge on due process grounds. . . . In order to accept jurisdiction and step in to overturn this action of the Commission, we would have to ignore the ANGTA in precisely the type of situation where it most compellingly applied. This would produce exactly the result the Congress tried to prevent."

More specifically, the Court explained, the ANGTA prevents a review of agency decisions on the pipeline system for reasonableness or substantial support on the record. Instead, "the courts are strictly limited to reviewing claims of denial of constitutional rights, or action in excess of statutory jurisdiction, authority, or limitations, or action short of statutory right."

The Court rejected the complainants' claim that the FERC violated due process by deciding the pressure issue in isolation from related issues such as carbon dioxide content and conditioning plant design, thereby foreclosing options and rendering any future hearing on the related issues meaningless. In order to invoke the due process guarantee of a hearing, the Court said, "one must show that a government action infringes upon a liberty or property interest." And if the complaint concerns administrative action, due process hearing requirements apply only to adjudicatory proceedings which concern particular effects on an individual or a small group of persons. "Complaints have met neither prerequisite." First, the Court explained, they really do not assert any damage from the FERC action to the environment, to existing property interests or to any liberty interests. The most they can point to is a possible impairment of Alaskan development which they -- "but not necessarily everyone" -- favor. Moreover, the desire of Alaska to develop a petrochemical industry which is alleged to be frustrated by the Commission's order "is far removed from the sort of legitimate claim to entitlement necessary to demonstrate a liberty or property interest." To put the matter in broader perspective, the Court stressed, the Commission's decision on the pressure issue "is a prospective, legislative resolution of a general policy issue, by rulemaking, which in contrast to an adjudication does not give rise to the usual due process rights associated with a hearing. Therefore, complaints have not stated a valid due process objection to the Commission's actions." Instead, they actually attack the Commission's decision to proceed on an issue-by-issue basis which goes to the reasonableness of the Commission's exercise of its discretion. Hence, the Court has no jurisdiction under ANGTA to review.

Next, the D.C. Circuit answered the complainants' further argument that the Commission's order approving the design specifications was not issued within the normal framework of Section 7(c) certificate applications and it thereby exceeded statutory limitations. Northwest did not apply for a certificate, they argued, but only an order on particular issues and the Commission did not require the usual detailed supporting exhibits. These objections, the Court concluded, "fly in the face of the ANGTA," which authorized the Commission to issue a certificate "forthwith." The Court reiterated that the procedures adopted to that end by the Commission were within its discretion and beyond jurisdiction to review. "Whether the FERC has fully considered factors and alternatives bearing on public necessity and convenience is an issue going to reasonableness and the substantiality of support for the order; this type of issue, ANGTA expressly forecloses for consideration." For the same reason, the Court added, it cannot review the claim that the Commission improperly accorded a presumption to the 1260 psig pressure level based on the President's Decision. The statutory preclusion of the Court's review of these issues "is crystal clear."

Finally, the D.C. Circuit rejected the contention that the FERC was required to prepare an environmental impact statement in accordance with the National Environmental Policy Act. First, the Court observed, Congress set forth in the ANGTA "specific and far-reaching restrictions on judicial review of the decisions related to the pipeline system." Specifically, the Act provided that Congressional ratification of the President's pipeline system decision will be conclusive as to the legal and factual sufficiency of the EIS submitted by the President with respect to the approved system. "This provision goes on to state that no court has jurisdiction to consider the sufficiency of such statements under NEPA. The effect of this section is place review in the Congress instead of the courts. By ratifying the President's 1977 decision, Congress has approved the EISs for the pipeline system."

The Court noted that in an attempt to avoid the force of the Act's preclusion of judicial review on NEPA questions, the complainants argued that if the Commission permitted another pressure for the pipeline, a different location for the conditioning plant would have been permitted as desired by Alaska. In effect, it was suggested that the FERC decided the issue of a conditioning plant location without preparing a final EIS; and since the conditioning plant is outside the scope of the pipeline system approved in the President's Decision, the Act's limited review provisions do not apply. The problem with this theory, the Court said, is that the issue of conditioning plant location was not a subject of the application and the order under review here, and the Commission did not take any major federal action on the conditioning plant. Instead, it took an entirely different action which happens to affect future consideration of the conditioning plant issue. The actual decision of the Commission that the complainants claim is a violation of NEPA is the pipeline pressure decision which is shielded from judicial review as regards the adequacy of the EIS.

Furthermore, the Court added, to require a separate EIS for the pipeline pressure issue would delay eventual construction by months and perhaps years. "The interrelationship between issues, which is the foundation of the complainants' argument, could make the delay even longer. . . . Such concerns underly the Commission's decision to proceed with separate issues and Congress' decision to shield the decisionmaking process from judicial review when constitutionally permissible." Even with provisions of the ANGTA to expedite pipeline construction, the Court noted, it has already taken two years since the President's Decision for the Commission to approve a pressure level, and final certificates and commencement of construction are still further in the future. "In this light, if there is any shortcoming in the Commission proceedings, it is certainly not in a lack of deliberation or in denial of time and opportunity for interested parties to express their views."

The Court's decision was signed by Circuit Judge Wilkey for the panel completed by Circuit Judges Robb and Mikva.

FERC Orders Northern Border Pipeline to Refile Certification Cost Estimates for Prebuild ANGTS Facilities

On 1/4/80 the FERC granted a motion filed on 11/29/79 by the Staff for an order directing Northern Border Pipeline Co. to refile certification cost estimates for the prebuilding of the Eastern Leg of the Alaska Highway Pipeline Project. The Commission agreed with the Staff that more detail is required to conform Northern Border's cost estimate with the cost format criteria previously established by the Commission in a procedural order issued 9/6/79.

In September 1977 the President approved the Alcan-Foothills proposal to construct the ANGTS. The decision included a condition requiring the FERC to review certification cost estimates filed by the sponsors to determine whether they materially and unreasonably exceed the estimates filed with the Commission in March 1977. The Commission then established specific requirements for review of cost estimates as part of the final certification process for the ANGTS in Order No. 31 (RM78-12) issued 6/8/79. Also, the Commission fixed values for the Incentive Rate of Return mechanism to be applied to the Alaskan and Northern Border segments for the purpose of controlling construction cost overruns and decided most tariff issues for the same two segments. The Commission directed that certification cost and schedule estimates must be filed, according to cost estimate formats approved by the Commission, to determine if they materially and unreasonably exceed comparable capital cost estimates filed by Alcan. In its 9/6/79 order, the Commission set out procedures for the submission and examination of cost estimates by Northern Border (and Pacific Gas Transmission Co.). Specifically, the Commission adopted cost estimate formats recommended by the Alaskan Delegate and decided that its review thereof for the prebuild facilities would proceed concurrently with the ongoing adjudicatory proceeding (Northwest Alaskan Pipeline Co., CP78-123 et al.). The Commission observed that pursuant to Order No. 31, it must determine a Center Point for the IROR mechanism. Also Northern Border sponsors may request a change in the Center Point value and file necessary material and support thereof.

On 11/29/79 the FERC Staff filed a motion requesting, among other things, that the Commission reject Northern Border's certification cost estimates, and require refiling thereof in compliance with the 9/6/79 order or pursuant to an alternative format suggested by Staff.

In support, the Staff argued that Northern Border's filing should be rejected because it does not permit a comparison of the March 1977 estimate as required by the President's Decision nor highlight reasons for changes which were to be solicited by the Alaskan Delegate's formats. Contrary to the express purpose of these formats, the Staff said, it is not possible to determine from Northern Border's filing the reasons for, among other things, changes in design and revised estimates of labor productivity. This is so, Staff explained, because Northern Border did not recast the March 1977 estimate into the formats of the certification cost estimate but instead started with the certification cost estimate and recast it into a two-year construction program instead of a one-year construction program. "Thus, rather than showing any increases in the quantities of resources required or any changes that are due to design modifications or changes in expected productivity rates, the proffered filing merely reflects the difference in cost that is due to Northern Border's decision to follow an accelerated construction schedule."

The Staff also suggested an alternative format whereby Northern would apply inflation indices to "bottom line" 1975 total dollar costs to produce a 1979 dollar figure for that estimate.

Finally, Staff objected to Northern Border's suggestion that it will use a formula specified in Order No. 31 for its Center Point determination and not seek a change thereof assuming the Commission does not make any changes to its entire cost estimate filing. Any election regarding a change in Center Point, Staff said, must be made when the certification cost estimate is filed. Hence, if the FERC directs refiling, Northern Border should also be ordered to make its election at that time. In a subsequent response in opposition to Staff's motion, Northern Border emphasized that the description "March 1977 estimate" is a misnomer as applied to it since Northern Border itself did not prepare or file any cost estimate at that time. What was filed in March 1977, Northern Border said, was an Alcan estimate. At that time, Northern Border explained, it was an intregal part of the Arctic Gas Project and was in an adversary competitive position with Alcan. In March 1977, Alcan "adopted" Northern Border and made it an intregal part of its proposal as the Lower 48 Eastern Leg. Cost estimates then presented by Alcan included an estimate for Northern Border which was not new at that time but had been the same cost estimate filed in May 1976 in the El Paso Alaska case and was based on 7/1/75 dollars.

As to Staff's contention that certain of its exhibits lack sufficient detail, Northern Border argued that Staff previously agreed at a conference on 9/21/79 that the exhibits were acceptable. However, Northern Border stated that workpapers to support these exhibits are available.

In its order, the Commission agreed with the Staff that Northern Border's filings did not provide sufficient documentation for analysis. The Commission noted that its cost criteria require that all cost estimates be explained in full detail. Specifically, the March 1977 cost estimates must be presented in such a manner as to show derivation of the original estimates by labor and material categories, and also 1975 costs are to be shown. Then the 1977 estimate must be recast into a Work Breakdown Structure format of the certification cost estimate and, in as many exhibits as needed, information for each WBS element is to be provided down to the lowest level. As to the certification cost estimate, the Commission continued, its criteria require (also in as many exhibits as are needed) details for each estimate at the lowest level of WBS as appropriate for a "thorough understanding of the composition and derivation of the estimate." The comparison of the two estimates should contain an explanation of all variances.

The FERC recognized that a March 1977 cost estimate for Northern Border's prebuild portion of the ANGTS does not exist per se and that analytical techniques must be used to derive such a value. However, the Commission emphasized that the derived estimate must correspond as nearly as possible to one which would have been made in 1977 if the Northern Border prebuild portion had been considered then. The Commission noted Staff's objections to the fact that Northern Border did not recast the March 1977 estimate into the formats of the certification cost estimate but instead started with the construction cost estimate and recast it into a two-year construction program instead of a one-year construction program. The Commission concluded that a one-year construction schedule should be assumed in the preparation of the 1977 estimate and that for purposes of preparing the studies, the two estimates should reflect a consistent set of assumptions as to the design of facilities. For example, the same size compressor stations and measurement stations should be assumed for both estimates. Design changes for pipe type and the number of communication system sites are to be considered as design changes.

The Commission directed Northern Border to prepare a new March 1977 estimate using assumptions of the 1979 certification estimate to derive the estimate where no known design changes are involved.

In response to Staff's request for clarification of Order Nos. 31 and 31-B to require Northern Border to make any election to change its Center Point at the time of filing its certification cost estimate, the Commission noted that it gave the project sponsors two choices as to how the Center Point would be determined. The first was to use a formula based on a comparison of the certification cost estimate and the estimates in the President's Decision. Under the second, the sponsors could request a Center Point without reference to the formula if a major change occurred in the project resulting in a total estimated cost which would exceed the estimates in the Decision. This request was to be included in the certification cost estimate submission.

Accordingly, the Commission said, if Northern Border does not choose to use the Center Point formula it has approved, it may elect to assess the likelihood of abnormal events that could increase costs and determine the impact thereof on costs. And this information should be submitted as part of Northern Border's certification cost estimate. Accordingly, if Northern Border desires to request a change in its Center Point, it should make this selection at the time of filing the required revisions to its March 1977 estimate herein.

The Commission directed the Law Judge to establish a schedule for filing answering and rebuttal testimony and a date for hearing on Northern Border's revised filings and methodology. 1/

Distributors and Pipelines Differ on Proposed Policy Regarding Distributor Access to OCS Gas

On 1/7/80 a public hearing was held on a proposed FERC rulemaking (RM80-11) to adopt a statement of policy with respect to distributor access to Outer Continental Shelf gas. Pipeline and distributor spokesmen differed on various aspects of the proposed rulemaking, particularly as to whether the Commission has the authority to direct unwilling interstate pipelines to transport distributor OCS gas.

This rulemaking was initiated pursuant to Section 603 of the Outer Continental Shelf Lands Act Amendments of 1978 which directed the FERC to publish a statement of policy -- within one year from enactment if possible -- to encourage expanded participation by local distribution companies in OCS lease acquisition and natural gas development through facilitating transportation of distributor-owned gas from OCS leases to distributor service areas. Section 603 also directed that the statement of policy set forth standards to be applied in considering certificate applications for such transportation, identify conditions which may be attached to such certificates, and specify criteria or requirements for determining whether an application of a local distribution company qualifies under the statement of policy.

The proposed rulemaking sets out procedures and standards to govern interstate pipeline transportation of "eligible OCS natural gas" (produced either from a lease entered into after 4/20/77 or from an interest acquired prior to the date of the final rule in a pre-4/20/77 lease) to an "eligible recipient" (a regulated distributor) from an "eligible OCS producer" (a regulated distributor, a producing affiliate of such distributor, or a joint venture in which either participates).

1/ On 1/10/80 Law Judge Lotis directed that direct, answering and rebuttal evidence be submitted on 1/15, 2/1 and 2/13/80, respectively, and that the hearing commence on 2/25/80.

CONDITIONING PLANT
(GASLINE)

EXXON COMPANY, U.S.A.
1800 AVENUE OF THE STARS • LOS ANGELES, CALIFORNIA 90067 (213) 552-5400

PRODUCTION DEPARTMENT
WESTERN DIVISION

January 21, 1980

Heating Values Used in
Parsons Study Report

Ms. Connie Barlow
811 Basin
Juneau, Alaska 99801

As you requested in the meeting in Houston on January 14, attached are the heating values used in the September, 1978 Parsons Study Report. As the footnote on the attached table indicates, these heating values are from the 1972 edition of the Gas Processors Suppliers Association Engineering Data Book.

If you have any questions, or we can be of additional assistance, please call.

Very truly yours,

James L. Shanks, Jr.
Division Gas Engineer

GJM:ae
Attachment

HEATING VALUES[1] USED IN PARSONS STUDY REPORT

(Btu/SCF)

	LHV	HHV
C_1	909.1	1009.7
C_2	1617.8	1768.8
C_3	2316.1	2517.4
iC_4	3001.1	3252.7
nC_4	3010.4	3262.1
iC_5	3698.3	4000.3
nC_5	3707.5	4009.5
C_6	4403.7	4756.1
C_7	5100.2	5502.9
C_8	5796.7	6249.7
C_9+	6493.3	6996.6
H_2S	588.0	637.0

1. Heating values are from the Engineering Data Book of the Gas Processors Suppliers Association (1972).

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1-16-80
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4-4

File 623 Liquids
OR GAS CONDITIONING PLANT

GAS PLANT LIQUIDS
POTENTIAL, RECOVERED, AND DISPOSITION
PRUDHOE BAY GAS CONDITIONING PLANT

Product	Potential In Inlet Gas MB/D	Components Recovered as Liquid Feed in the Conditioning Plant(1) % of Inlet	Components Recovered as Liquid Feed in the Conditioning Plant(1) MB/D	Commercial Liquid Products from Fractionator Train Total(2) MB/D	Disposition Cond. Plt. Fuel MB/D	Disposition Field Fuel MB/D	Disposition Crude Oil MB/D	Disposition Sales Gas MB/D	Components in Sales Gas Residue Gas MB/D	Components in Sales Gas Total MB/D
Ethane	111.4	20	22.8	-	-	-	-	-	56.5	58.1
Propane	61.8	77	47.5	52.4	14.0	14.1	0	24.3	1.1	23.3
I-Butane	10.3	93	9.6	} 31.2	} 0	} 0	} 0	} 31.2	0.2	9.4
N-Butane	23.7	95	22.5						0.3	21.9
Pentane plus	27.3	99	26.9	28.6	0	0	28.6	0	0.2	1.0
TOTAL	234.5		129.3	112.2	14.0	14.1	28.6	55.5	58.3	113.7

(4) + (14)

(1) Feed to fractionator train: Liquid from low temperature separator plus liquid from local fuel fractionator. Not a commercial product - contains about 13% methane which is not included in these volumes.

(2) There are three commercial products produced: Propane, mixed-butanes, and pentane-plus (natural gasoline). These commercial products are not pure components; for example, the propane product contains some ethane which accounts for more ethane in the total sales gas than is indicated by adding the numbers shown on this table.

HRG/lh
6/22/79

PRUDHOE BAY CONDITIONING PLANT
MAXIMUM FUEL CASE STREAM COMPOSITIONS

	(1) INLET GAS		(2) LTS FEED (INLET GAS & SELEXOL REGENERATOR OVHD)		(3) LTS VAPOR		(4) LTS LIQS (DeC_2 FEED)	
	%	MB/D	%	MB/D	%	MB/D	%	MB/D
C_1	74.00	-	72.20	-	76.88	-	19.90	-
C_2	6.53	111.4	6.35	114.3	5.59	91.5	14.80	21.9
C_3	3.52	61.8	3.66	67.9	1.95	32.8	22.87	34.9
iC_4	0.50	10.3	0.50	11.0	0.14	2.8	4.53	8.2
nC_4	1.18	23.7	1.18	25.0	0.24	4.6	11.64	20.3
iC_5	0.27	6.2	0.26	6.3	0.02	0.5	2.88	5.8
nC_5	0.49	11.4	0.48	11.6	0.03	0.6	5.46	11.0
C_6	0.19	5.1	0.18	5.1	-	0.1	2.21	5.0
C_7	0.12	3.2	0.11	3.3	-	-	1.38	3.2
C_8	0.04	1.2	0.04	1.2	-	-	0.50	1.2
C_9	0.01	0.2	0.01	0.2	-	-	0.07	0.2
H_2S	0.00	-	0.00	-	-	-	-	-
CO_2	12.68	-	14.57	-	14.65	-	13.72	-
N_2	0.47	-	0.46	-	.50	-	0.02	-
TOTAL	100.0	234.5	100.00	245.9	100.00	132.9	100.00	111.7
MMSCFD	2686		2836		2578		233	
BTU/CF	1055		1036		939		2117	
PSIA	576		565		540		495	
°F	98		96		18		86	

PRUDHOE BAY CONDITIONING PLANT
MAXIMUM FUEL CASE STREAM COMPOSITIONS

	(5) DeC_2 OVHD TO FIELD LEVEL		(6) DeC_2 BTMS		(7) SELEXOL ABSORBER FEED		(8) SELEXOL ABSORBER OVMD	
	%	MB/D	%	MB/D	%	MB/D	%	MB/D
C_1	42.33	-	0.00	-	76.88	-	94.28	-
C_2	28.13	20.0	2.98	2.4	5.59	91.5	4.47	56.5
C_3	0.41	0.3	42.57	35.3	1.95	32.8	0.08	1.1
iC_4	-	-	8.55	8.4	0.14	2.8	0.01	0.2
nC_4	-	-	22.02	20.9	0.24	4.6	0.02	0.3
iC_5	-	-	5.48	6.0	0.02	0.5	-	0.1
nC_5	-	-	10.42	11.4	0.03	0.6	0.01	0.1
C_6	-	-	4.23	5.2	-	0.1	-	-
C_7	-	-	2.65	3.4	-	-	-	-
C_8	-	-	0.96	1.3	-	-	-	-
C_9	-	-	0.13	.2	-	-	-	-
H_2S	.01	-	-	-	-	-	-	-
CO_2	29.08	-	-	-	14.65	-	0.50	-
N_2	.04	-	-	-	0.50	-	0.62	-
TOTAL	100.00	20.3	100.00	94.6	100.00	132.9	100.00	58.3
MMSCFD	112		127		2578		1994	
BTU/CF	935		3174		939		1035	
PSIA	545		480		540		530	
^oF	2		249		18		86	

PRUDHOE BAY CONDITIONING PLANT
MAXIMUM FUEL CASE STREAM COMPOSITIONS

	(9) SELEXOL ABSORBER BTMS		(10) SELEXOL REGENERATOR OVHD		(11) SELEXOL FLASH VAPOR		(12) LFF FEED	
	%	MB/D	%	MB/D	%	MB/D	%	MB/D
C_1	15.53	-	39.99	-	19.66	-	1.30	-
C_2	9.85	39.3	3.07	2.9	15.32	19.0	8.61	13.0
C_3	8.40	34.5	6.32	6.2	7.58	9.7	10.18	15.8
iC_4	0.67	3.3	0.60	0.7	0.38	0.6	0.74	1.4
nC_4	1.22	5.7	1.04	1.2	0.57	0.8	1.26	2.2
iC_5	0.25	1.4	0.10	0.1	0.03	0.1	0.11	0.2
nC_5	0.42	2.3	0.13	0.2	0.04	0.1	0.14	0.3
C_6	0.11	0.7	0.01	-	-	-	0.02	-
C_7	0.07	0.4	0.01	-	-	-	-	-
C_8	0.02	0.1	-	-	-	-	-	-
C_9	-	-	-	-	-	-	-	-
H_2S	0.01	-	0.01	-	-	-	-	-
CO_2	63.42	-	48.44	-	56.41	-	77.65	-
N_2	0.02	-	0.27	-	0.01	-	-	-
TOTAL	100.00	87.7	100.00	11.3	100.00	30.3	100.00	33.0
MMSCFD	628		150		196		238	
BTU/CF	642		681		694		498	
PSIA	540		565		75		350	
^oF	19		79		13		131	

PRUDHOE BAY CONDITIONING PLANT
MAXIMUM FUEL CASE STREAM COMPOSITIONS

	(13) LFF OVHD		(14) LFF BTMS		(15) DeC_3FEED (DeC_2, BTMS & LFF BTMS)		(16) DeC_3 OVHD (PROPANE)	
	%	MB/D	%	MB/D	%	MB/D	%	MB/D
C_1	1.46	-	-	-	-	-	-	-
C_2	9.04	12.1	5.17	0.9	3.35	3.3	6.41	3.3
C_3	2.29	3.2	74.16	12.6	47.96	48.0	91.58	47.9
iC_4	-	-	6.74	1.4	8.24	9.8	1.44	0.9
nC_4	-	-	11.45	2.2	20.21	23.2	0.56	0.3
iC_5	-	-	0.98	0.2	4.71	6.3	-	-
nC_5	-	-	1.29	0.3	8.86	11.7	-	-
C_6	-	-	0.14	-	3.54	5.3	-	-
C_7	-	-	0.03	-	2.21	3.4	-	-
C_8	-	-	-	-	0.80	1.3	-	-
C_9	-	-	-	-	0.11	0.2	-	-
H_2S	-	-	-	-	-	-	-	-
CO_2	87.21	-	0.03	-	0.01	-	0.01	-
N_2	-	-	-	-	-	-	-	-
TOTAL	100.00	15.3	100.00	17.7	100.00	112.2	100.00	52.4
MMSCFD	212		26		153		80	
BTU/CF	232		2651		3085		2484	
PSIA	325		340		220		200	
^{o}F	-4		156		175		90	

PRUDHOE BAY CONDITIONING PLANT
MAXIMUM FUEL CASE STREAM COMPOSITIONS

	(17) DeC_3 BTMS (DeC_4 FEED)		(18) DeC_4 OVHD (Butanes)		(19) DeC_4 BTM (Pentanes - Plus)		(21) PROPANE TO LOCAL FUEL	
	%	MB/D	%	MB/D	%	MB/D	%	MB/D
C_1	-	-	-	-	-	-	-	-
C_2	-	-	-	-	-	-	6.41	0.9
C_3	0.10	-	0.18	-	-	-	91.58	12.8
iC_4	15.70	8.9	27.72	8.9	0.17	-	1.44	0.2
nC_4	41.78	22.8	69.58	21.4	5.84	1.4	0.56	0.1
iC_5	9.89	6.3	2.15	0.8	19.89	5.5	-	-
nC_5	18.59	11.7	.38	0.1	42.13	11.5	-	-
C_6	7.41	5.3	-	-	17.00	5.3	-	-
C_7	4.63	3.4	-	-	10.62	3.4	-	-
C_8	1.67	1.3	-	-	3.84	1.3	-	-
C_9	0.22	0.2	-	-	0.52	0.2	-	-
H_2S	-	-	-	-	-	-	-	-
CO_2	-	-	-	-	-	-	-	-
N_2	-	-	-	-	-	-	-	-
TOTAL	100.00	59.8	100.00	31.2	100.00	28.6	100.00	14.0
MMSCFD	73		41		32		21	
BTU/CF	3745		3277		4350		2484	
PSIA	75		60		70		640	
^oF	161		90		90		90	

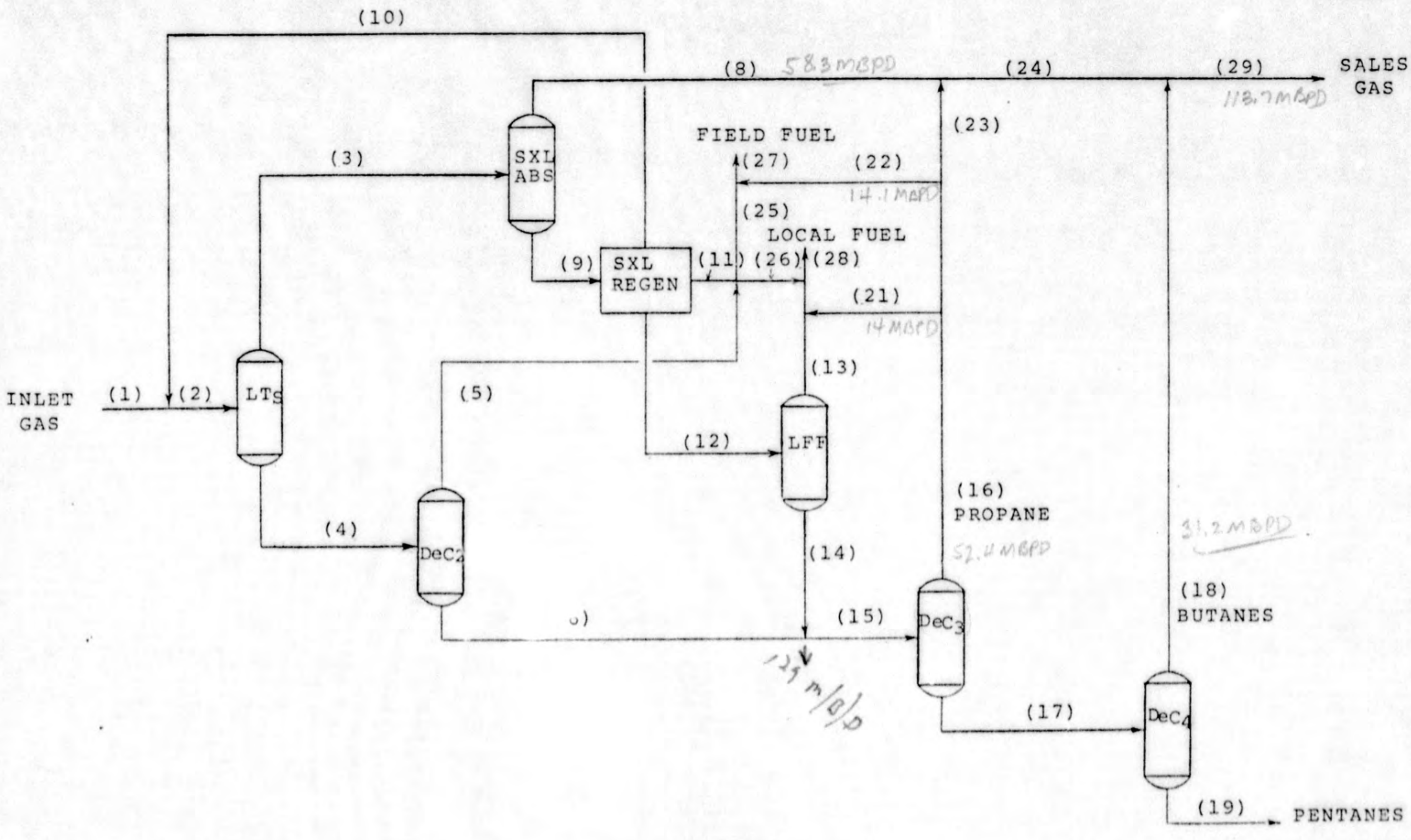
PRUDHOE BAY CONDITIONING PLANT
MAXIMUM FUEL CASE STREAM COMPOSITIONS

	(22) PROPANE TO FIELD FUEL		(23) PROPANE TO SALES GAS		(24) SALES GAS WITH PROPANE		(25) SELEXOL FLASH VAPOR TO FIELD FUEL	
	%	MB/D	%	MB/D	%	MB/D	%	MB/D
C_1	-	-	-	-	92.56	-	19.66	-
C_2	6.41	0.9	6.41	1.5	4.50	58.1	15.32	17.6
C_3	91.58	12.9	91.58	22.2	1.75	23.3	7.58	9.0
iC_4	1.44	0.2	1.44	0.4	0.04	0.6	0.37	0.5
nC_4	0.56	0.1	0.56	0.2	0.03	0.5	0.56	0.8
iC_5	-	-	-	-	0.01	0.1	0.04	0.1
nC_5	-	-	-	-	0.01	0.1	0.05	0.1
C_6	-	-	-	-	-	-	-	-
C_7	-	-	-	-	-	-	-	-
C_8	-	-	-	-	-	-	-	-
C_9	-	-	-	-	-	-	-	-
H_2S	-	-	-	-	-	-	-	-
CO_2	-	-	-	-	0.49	-	56.41	-
N_2	-	-	-	-	0.61	-	0.01	-
TOTAL	100.0	14.1	100.00	24.3	100.00	82.6	100.00	28.0
MMSCFD	22		37		2032		181	
BTU/CF	2484		2484		1061		694	
PSIA	640		640		525		520	
$^{\circ}F$	90		90		75		208	

PRUDHOE BAY CONDITIONING PLANT
MAXIMUM FUEL CASE STREAM COMPOSITIONS

	(26) SELEXOL FLASH VAPOR TO LOCAL FUEL		(27) FIELD FUEL		(28) LOCAL FUEL		(29) SALES GAS W/ PROPANE & BUTANES	
	%	MB/D	%	MB/D	%	MB/D	%	MB/D
C_1	19.65	-	37.74	-	2.42	-	90.72	-
C_2	15.30	1.4	16.24	41.1	9.18	14.5	4.41	58.1
C_3	7.59	0.7	8.84	23.0	10.31	16.7	1.72	23.3
iC_4	0.37	-	0.27	0.8	0.51	0.3	0.59	9.4
nC_4	0.55	0.1	0.33	1.0	0.08	0.1	1.41	21.9
iC_5	0.06	-	0.02	0.1	-	-	0.05	0.8
nC_5	0.06	-	0.03	0.1	-	-	0.01	0.2
C_6	-	-	-	-	-	-	-	-
C_7	-	-	-	-	-	-	-	-
C_8	-	-	-	-	-	-	-	-
C_9	-	-	-	-	-	-	-	-
H_2S	-	-	-	-	-	-	-	-
CO_2	56.43	-	36.40	-	77.85	-	0.49	-
N_2	-	-	0.13	-	-	-	0.60	-
TOTAL	100.00	2.3	100.00	66.1	100.00	31.6	100.00	113.8
MMSCFD	15		398		248		2071	
BTU-CF	695		912		454		1106	
PSIA	515		520					
^oF	90		81					

PRUDHOE BAY SALES GAS CONDITIONING PLANT
MAXIMUM FUEL CASE (DeC_2 OVHD TO FIELD FUEL)



2-4

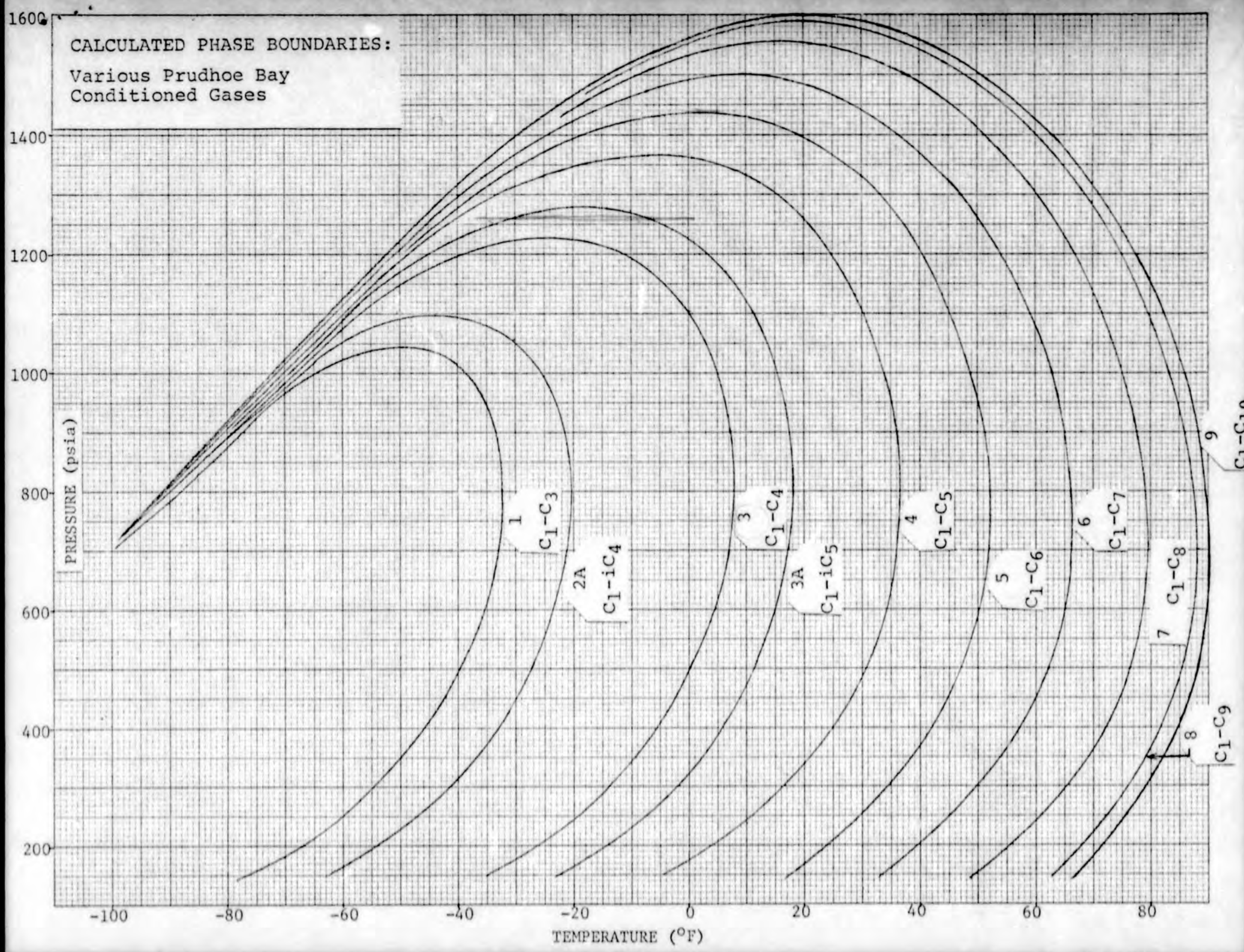
VARIOUS PRUDHOE BAY CONDITIONED GAS COMPOSITIONS
(Mole Percent)

Component	Unconditioned Separator Off-Gas	1 C_1-C_3	2A C_1-iC_4	3 C_1-C_4	3A C_1-iC_5	4 C_1-C_5	5 C_1-C_6	6 C_1-C_7	7 C_1-C_8	8 C_1-C_9	9 C_1-C_{10}
N_2	0.484	0.564	0.561	0.554	0.553	0.551	0.550	0.549	0.549	0.549	0.549
CO_2	12.659	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000
C_1	74.706	87.053	86.594	85.554	85.341	84.964	84.818	84.742	84.695	84.679	84.676
C_2	6.428	7.491	7.451	7.362	7.343	7.311	7.299	7.292	7.288	7.287	7.287
C_3	3.340	3.892	3.872	3.826	3.815	3.799	3.793	3.789	3.787	3.786	3.786
$i-C_4$	0.450	---	0.522	0.515	0.514	0.512	0.511	0.511	0.510	0.510	0.510
$n-C_4$	1.038	---	---	1.189	1.186	1.181	1.179	1.178	1.177	1.177	1.177
$i-C_5$	0.217	---	---	---	0.248	0.247	0.247	0.246	0.246	0.246	0.246
$n-C_5$	0.383	---	---	---	---	0.435	0.435	0.434	0.434	0.434	0.434
C_6	0.148	---	---	---	---	---	0.168	0.168	0.168	0.168	0.168
C_7	0.081	---	---	---	---	---	---	0.091	0.092	0.092	0.092
C_8	0.047	---	---	---	---	---	---	---	0.054	0.054	0.054
C_9	0.016	---	---	---	---	---	---	---	---	0.018	0.018
C_{10}	0.003	---	---	---	---	---	---	---	---	---	0.003
Mol. Wt.	22.7	18.5	18.7	19.2	19.3	19.5	19.7	19.8	19.8	19.8	19.9
Heating Value* (Btu/cf)	1027	1095	1106	1131	1138	1150	1156	1160	1163	1164	1164

2%

*Gross, Wet, Actual @ 60°F., 14.73 psia

5/5/78


CALCULATED PHASE BOUNDARIES:
Various Prudhoe Bay Conditioned Gases
PRESSURE (psia)
TEMPERATURE (°F)
1600
1400
1200
1000
800
600
400
200
-100
-80
-60
-40
-20
0
20
40
60
80
1 C_1-C_3
2A C_1-iC_4
3 C_1-C_4
3A C_1-iC_5
4 C_1-C_5
5 C_1-C_6
6 C_1-C_7
7 C_1-C_8
8 C_1-C_9
9 C_1-C_{10}

PLEASE NOTE: THE FOLLOWING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT

GASLINE — CONDITIONING PLANT

ARLON R. TUSSING & ASSOCIATES, INC. / 2720 Rainier Bank Tower / Seattle, Washington 98101 / (206) 447-0321

December 3, 1979

Mr. Ray Booth
Assistant Manager
Natural Gas Department
Exxon Company USA
800 Bell Street
Houston, Texas 77001

Dear Mr. Booth:

You are undoubtedly aware of the recent preoccupation of the public and of state officials in Alaska with the availability of natural gas liquids from Prudhoe Bay for industrial use within the state. Unfortunately, most of the current deliberation and debate on this question is taking place without a real understanding of the technical and economic tradeoffs.

The Alaska Department of Natural Resources has engaged Arlon R. Tussing and Associates, Inc. to write a report explaining to state officials, legislators, and the public what the technical and economic relationships are among the choice of gas pipeline pressure, location and technology of the gas conditioning plant, choice of field and pipeline fuel, and the availability and cost of natural gas liquids for use at different locations in Alaska. Our challenge is to reduce a number of extremely complex economic and engineering issues to a form in which non-specialists can deal with them intelligently.

Our investigation has now proceeded about as far as it can on the basis of published materials, because many of the available documents fail to set out clearly the assumptions upon which their scientific and engineering judgments are made. The following is a list of the questions we have so far been unable to resolve. Our biggest need from you with respect to these questions is to know what assumptions underlie your company's judgment (and that of the others) about the ability of pipeline systems of various pressures to transport different hydrocarbon mixtures.

1220 19th Ave. N.W., Suite No. 201, Washington D.C. 20036 / 880 H St., Suite No. 210, Anchorage, Alaska 99501

We do not know how well these issues can be resolved by correspondence or by telephone, but Connie C. Barlow (who has principal responsibility for this project) and/or I would be willing to travel to a place and at a time of your convenience, in order to discuss them in depth. Ms. Barlow will be calling you toward the end of this week to explore how we ought to proceed.

Your help will be greatly appreciated.

Sincerely,

Arlon R. Tussing

ART/bas

Enclosure

cc Connie Barlow
Robert Maynard
Mary Halloran

QUESTIONS FOR THE PRODUCERS

Question #1:

With respect to determining the actual volumes of methane and gas liquids that will become available for shipment ...

(a) What is the expected composition and daily production rate of total raw gas (less whatever is reinjected to maintain reservoir pressures)? How much uncertainty accompanies anybody's best guesses about this matter?

(b) How is the daily production rate and composition expected to change through time, particularly when oil production begins to drop off? (Specifically, is there a significant difference in composition of solution gas versus gas cap gas, and is the ratio of produced volumes of solution and gas cap gas affected by time?)

(c) Of these produced hydrocarbons how much of each will be consumed to meet North Slope total fuel requirements? How will field fuel requirements change through time as oil production and transportion drop?

(d) Hence, what volumes of methane, ethane, propane, butane, and pentanes+ will actually be available through time for delivery to some transportion system?

Question #2:

Concerning upset conditions in the gasline ...

(a) For each pipeline pressure specification under consideration (1080 psig in Canada and 1260, 1440, and 1680 psig in Alaska), what temperature and pressure combinations constitute "upset" conditions, "normal minimum operating" conditions, or whatever terminology you use to signify the most troubling conditions (from the standpoint of condensation) that gas may encounter during the journey south?

(b) Are the phase characteristics of probable gas mixtures such that the temperature and pressure combinations of most concern represent the <u>lowest</u> temperature and the <u>lowest</u> pressure likely to be encountered?

(c) At least in the documents we have seen, there appears to be some difference of opinion on the answer to the above questions. For example, I believe Exxon has indicated that a 0 degree F upset temperature is appropriate in Alaska, and 30 degree F for the sections south of Whitehorse, while Sohio lists -10 degrees and +42 degrees respectively. With regard to pressure, Sohio appears to assume that a 1680 psig line will have a 1300 psia upset while Arco assumes a 1200 psia upset. What might be the reasons behind these differences of opinion (e.g. compressor spacing, which compressor station is assumed out, duration of outage, immediacy of response in reducing throughput, normal station outlet temperature of the gas, ambient air or ground temperature, etc.)?

<u>Question #3</u>:

With respect to the ability of TAPS to carry gas liquids ...

(a) What is the lowest TAPS inlet temperature that might <u>safely</u> be considered? What are the physical considerations that determine this (i.e. what worst case assumptions about ambient air temperature and duration do you make with regard to TAPS shutdown or pump station outage?) Are there absolute physical limitations on the minimum inlet temperature, or is it primarily an economic matter of the costs and agravation of frequent pigging?

(b) Approximately how much would it cost to add chilling facilities to bring the inlet temperature down to the level indicated above?

(c) Are light hydrocarbons such as butane and pentane useful in reducing oil pipeline viscosity and friction, thereby increasing throughput capabilities with existing equipment? Would one part of additional butane, therefore, require leaving behind maybe less than one part of heavier hydrocarbons if the line were already functioning at capacity? Given the likely timing of gas production for sale, and given the likely oil production scenario, is it a realistic prospect that throughput limitations of the line and pump stations might make it impossible to ship butanes and pentanes separate.. from the raw gas unless oil shipments are cut back?

(d) What volumes of oil prouction for each year of field operation should one assume when trying to ascertain TAPS throughput capabilities? To your knowledge, did the other parties commenting on the ability of TAPS to carry light hydrocarbons assume similar production schedules?

(e) Given the oil production scenario noted above and the volume of butanes and pentanes implied in your answer to question 1(. , how much butane and pentane could be transportate hrougn TAPS, (noting time as a variable) at the minimum inlet temperature you have specified in your answer to question 3(a) at the existing TAPS inlet temperature?

(f) Currently, what (if anything) is being done at Valdez to ensure that oil shipped through TAPS (at 140 degrees F and 14.7 psia) meets tanker and lower 48 storage specifications for vapor pressure? How would those activities change if the TAPS inlet temperature were lowered to the point mentioned in your answer to question 3(a)?

<u>Questions #4:</u>

Concerning North Slope Field Fuel requirements ...

(a) Given the lik oil and gas production schedules you have indicate reviously, how will the required amounts of field el change through time?

(b) What are th physical and economic principles that limit flexibility in choosing the BTU content, dewpoint, and overall composition of field fuel? (e.g., the need to ensure a certain level of precision in combustion characteristics, dewpoint limitations imposed by exposure to ambient air temperatures en route, BTU specifications of existing equipment and the cost of changing or expanding that range.)

<u>Miscellaneous questions</u> ...

- When you speak of BTU's do you mean <u>gross</u> or <u>net</u> BTU's? When calculating the BTU values of various compositions of gases, what assumptions do you use with respect to the BTU content per mcf and per barrel for each hydrocarbon type (C_1 through C_5) What phase diagrams to you use for the following pure substances: C_1, C_2, C_3, C_4, C_5, CO_2, H_2O?

- How does the addition of CO_2 affect the dewpoint characteristics of any particular gas composition, <u>assuming that the composition of the hydrocarbon portion of the mixture remains unchanged</u>?

- On a very approximate basis, how much more money would it cost to isolate most of the produced ethane from the CO_2 rich stream during the conditioning process? Does this represent, say, the added cost of using a Sulfinol or similar process with required dehydration facilities? What other options are available?

- Has anybody done any work yet on the added cost of transporting a gas of 13% CO_2 compared to the cost of transporting a 1% or 3% mixture, holding the pipeline operating pressure constant?

- General questions on the effect of CO_2 content on the hazards of carbonic acid and hydrate corrosion and erosion at the pipeline operating pressures and temperatures under consideration.

PLEASE NOTE: THE PRECEDING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT.

PLEASE NOTE: THE FOLLOWING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT

GASLINE: CONDITIONING PLANT

SOHIO THE STANDARD OIL COMPANY
MIDLAND BUILDING, CLEVELAND, OHIO, 44115

January 23, 1980

Ms. Connie C. Barlow
Arlon R. Tussing & Associates, Inc.
811 Basin
Juneau, Alaska 99801

Prudhoe Bay Gas Sales

Dear Ms. Barlow:

In reply to the questionnaire attached to Arlon Tussing's letter of December 3, 1979 to Mr. Paul Phillips, attached is Sohio's response.

Many of the answers, particularly those to question #1, draw on data derived from the Parsons Study of September, 1978. While this is the best data we have, we would emphatically point out that the Parsons design is by no means definitive. The main objectives of the Parsons study were to get some idea of the cost and lead time of constructing a gas conditioning plant that would enable an assumed set of pipeline specifications to be met. Time did not permit adequate optimization of the plant design, and perhaps more important, overall optimization of the conditioning plant/pipeline system was not considered. The validity of Parsons' results should be viewed in this context.

We hope the attached answers are helpful to your effort, and will be happy to try to answer any queries you may have.

Very truly yours,

DJPritchard

D. J. Pritchard

DJP:ejh

Question #1:

With respect to determining the actual volumes of methane and gas liquids that will become available for shipment ...

(a) What is the expected composition and daily production rate of total raw gas (less whatever is reinjected to maintain reservoir pressures)? How much uncertainty accompanies anybody's best guesses about this matter?

(b) How is the daily production rate and composition expected to change through time, particularly when oil production begins to drop off? (Specifically, is there a significant difference in composition of solution gas versus gas cap gas, and is the ratio of produced volumes of solution and gas cap gas affected by time?)

(c) Of these produced hydrocarbons how much of each will be consumed to meet North Slope total fuel requirements? How will field fuel requirements change through time as oil production and transportion drop?

(d) Hence, what volumes of methane, ethane, propane, butane, and pentanes+ will actually be available through time for delivery to some transportion system?

RESPONSE #1

(a) An expected composition of produced gas in the early years of gas sales (taken from the Parsons Report) is given on table 1. The raw gas production rate used in the Parsons Report is 2.8 BCFD, although the current conservation order limits this production rate to 2.7 BCFD.

(b) It is expected that the initial gas production rate can be maintained for about 25 years. Gas composition will initially get leaner with time. At lower reservoir pressures in the latter stages of depletion this process can reverse itself, and the gas will tend to become richer and approach its initial composition. It is impossible to predict produced gas compositions at different times with any certainty.

Nevertheless, to give a feel for the gas compositions expected later in the field life, it is estimated that the ethane and propane contents will drop to minima of about 5.5% and 2.5% respectively.

(c) The Parsons Report analyses a maximum and a minimum fuel requirements case, and addresses the requirements for various components to meet these fuel demands under the Parsons design and the assumed raw gas composition. Field fuel estimates are further discussed in the response to question 4.

(d) This question has not been addressed except under the maximum and minimum fuel cases in the Parsons Report. The availability of sales volumes of various components in these two cases is given in table 2.

Question #2:

Concerning upset conditions in the gasline ...

(a) For each pipeline pressure specification under consideration (1080 psig in Canada and 1260, 1440, and 1680 psig in Alaska), what temperature and pressure combinations constitute "upset" conditions, "normal minimum operating" conditions, or whatever terminology you use to signify the most troubling conditions (from the standpoint of condensation) that gas may encounter during the journey south?

(b) Are the phase characteristics of probable gas mixtures such that the temperature and pressure combinations of most concern represent the <u>lowest</u> temperature and the <u>lowest</u> pressure likely to be encountered?

(c) At least in the documents we have seen, there appears to be some difference of opinion on the answer to the above questions. For example, I believe Exxon has indicated that a 0 degree F upset temperature is appropriate in Alaska, and 30 degree F for the sections south of Whitehorse, while Sohio lists -10 degrees and +42 degrees respectively. With regard to pressure, Sohio appears to assume that a 1680 psig line will have a 1300 psia upset while Arco assumes a 1200 psia upset. What might be the reasons behind these differences of opinion (e.g. compressor spacing, which compressor station is assumed out, duration of outage, immediacy of response in reducing throughput, normal station outlet temperature of the gas, ambient air or ground temperature, etc.)?

RESPONSE #2

(a)(b)&(c) This is a question that can only be definitively answered by the pipeline operators after some design work has been done. The definition of "upset conditions" depends on:

- Pipeline <u>operating</u> pressure (remembering that 1260 psia etc., are <u>maximum</u> pressures).

- **Pipeline operating temperature**
- **Compression ratio and compressor station spacing**
- **Soil temperatures and heat transfer**
- **Definition of what constitutes an "upset" (e.g. one compressor station shutdown for one month in mid-winter). This is dependent on the degree of backup and sparing in the pipeline design.**

It is suspected that the designs of neither the Canadian section nor the Alaskan section (for 1260 psia) are sufficiently advanced for the pipeline operators to have defined their upset conditions. The 1440 psia and 1680 psia proposed Alaskan designs are even less likely to have been evaluated with any precision.

In view of the above, projections to date on upset conditions are rough estimates, based in some part on Alcan's data. In this context the noted variances of 100 psi and 10°F or so between the various companies are not out of line.

We believe that an upset pressure of 200-400 psi below the maximum operating pressure is a reasonable educated guess. The difference between maximum operating pressure and upset pressure would tend to be wider at higher maximum operating pressures.

An upset operating temperature in the range 0°F to -10° F seems reasonable to us for permafrost areas (for practical matters, North of Whitehorse) and around 20 to 30°F south of Whitehorse. Note that the 42°F quoted as Sohio's Canadian upset temperature presumes that heaters will be installed at the critical Canadian compressor station #3; and that in the absence of such heaters we quoted 32°F. It is not known whether it is planned to install the necessary heaters.

The dew point specifications set out in the initial tariff filings of the pipeline sponsors are set out below:

Alaskan Northwest	Foothills (converted from original metric) North of 60°N	Foothills South of 60°N
-5°F @ 800 psia	-4°F @800 psia	
-10°F @ 1000 psia	-9°F @ 1000 psia	15°F @ 800 psia
-18°F @ 1100 psia	-18° F @ 1100 psia	

Presumably these dew points represent something close to the sponsors' worst case or upset conditions.

It is generally true that the conditions of concern represent the lowest pressure and lowest temperatures.

Question #5:

With respect to the ability of TAPS to carry gas liquids ...

(a) What is the lowest TAPS inlet temperature that might safely be considered? What are the physical considerations that determine this (i.e. what worst case assumptions about ambient air temperature and duration do you make with regard to TAPS shutdown or pump station outage?) Are there absolute physical limitations on the minimum inlet temperature, or is it primarily an economic matter of the costs and agravation of frequent pigging?

(b) Approximately how much would it cost to add chilling facilities to bring the inlet temperature down to the level indicated above?

(c) Are light hydrocarbons such as butane and pentane useful in reducing oil pipeline viscosity and friction, thereby increasing throughput capabilities with existing equipment? Would one part of additional butane, therefore, require leaving behind maybe less than one part of heavier hydrocarbons if the line were already functioning at capacity? Given the likely timing of gas production for sale, and given the likely oil production scenario, is it a realistic prospect that throughput limitations of the line and pump stations might make it impossible to ship butanes and pentanes separated from the raw gas unless oil shipments are cut back?

(d) What volumes of oil prouction for each year of field operation should one assume when trying to ascertain TAPS throughput capabilities? To your knowledge, did the other parties commenting on the ability of TAPS to carry light hydrocarbons assume similar production schedules?

(e) Given the oil production scenario noted above and the volume of butanes and pentanes implied in your answer to question 1(d), how much butane and pentane could be transportated through TAPS, (noting time as a

variable) at the minimum inlet temperature you have specified in your answer to question 3(a) at the existing TAPS inlet temperature?

(f) Currently, what (if anything) is being done at Valdez to ensure that oil shipped through TAPS (at 140 degrees F and 14.7 psia) meets tanker and lower 48 storage specifications for vapor pressure? How would those activities change if the TAPS inlet temperature were lowered to the point mentioned in your answer to question 3(a)?

RESPONSE #3

Having recognized the importance of the type of questions posed in 3, Sohio commenced a study earlier this year to determine the NGL carrying capacity of the oil transportation system. This study is due to be completed in early 1980, and the responses below give an indication of the results to date.

(a) There is a TAPS temperature in the region of about 100-105° below which wax begins to form. The pipeline has been operated at lower temperatures (e.g. during startup) but it is believed that two factors will act to set the miminum inlet temperature for NGL blending purposes at around 100°F.

- At the current inlet temperature of about 140°F, the oil cools during its passage through TAPS. However, TAPS pressure/temperature analyses predict that under high throughput summer operation, at an inlet temperature of 100-110°F, there is little or no cooling. At 100°F the oil actually gains temperature in certain pipeline sections and emerges at Valdez at essentially the same temperature as it left the North Slope. Clearly there is no benefit in cooling an oil NGL mix on the North Slope for stabilization, only to have the mixture subsequently warm up again.

- Various downstream jurisdictions impose vapor pressure limitations on hydrocarbons stored in floating roof tanks. The climates of the downstream locations, combined with their vapor pressure limitations suggest that cooling on the North Slope below about 100°F would achieve no practical benefit.

(b) The cost of crude cooling facilities to cool to 1.5 MMBD of crude to about $110^{o}F$ was estimated in the Parson's Study to be about $60 MM (1978 terms).

(c) Light hydrocarbons do indeed reduce viscosity. However, cooling the oil to accommodate light hydrocarbons tends to increase viscosity.
Given:

- the likely timing of gas sales (1985 or 1986 at the earliest).
- the likely onset of Prudhoe Bay oil decline in the same time frame.
- the flexibility of increasing TAPS throughput by injection of Drag Reductior Additive.
- the downstream vapor pressure constraints.

It is unlikely that NGL blending and/or oil production will be constrained by TAPS capacity, and if it is it will only be for a period of a year or so.

(d) Expected oil production volumes are shown on Fig 1 (attached). We believe this production profile does not drastically conflict with the views of other Prudhoe Bay Unit owners.

(e) Our present opinion is that the pentanes plus can be accommodated in the crude oil stream for much of the field life, since the vapor pressure characteristics of the pentanes plus/crude oil mix are little different from crude oil by itself. Some butanes may be mixed with the oil during early high throughput years.

(f) Vapor pressure control does not currently take place at Valdez. To control vapor pressure at this point would be a massive undertaking for the following reasons:

- The oil stream to be handled is large.
- Limited space is available for additional facilities in the terminal area.
- Sufficient storage would be required to accumulate cargo sized batches of the NGL's products.
- A fleet of Jones Act NGL tankers would have to be assembled to ship the NGL's for market, probably on the Gulf Coast.

Questions #4:

Concerning North Slope Field Fuel requirements ...

(a) Given the likely oil and gas production schedules you have indicated previously, how will the required amounts of field fuel change through time?

(b) What are the physical and economic principles that limit flexibility in choosing the BTU content, dew-point, and overall composition of field fuel? (e.g., the need to ensure a certain level of precision in combustion characteristics, dewpoint limitations imposed by exposure to ambient air temperatures en route, BTU specifications of existing equipment and the cost of changing or expanding that range.)

RESPONSE #4

(a) Fig. 2 (attached) gives a perspective of field fuel usage with time. The figure is scaled in billions of BTU/Day, which is equivalent to MMCFD of 1000 net BTU/SCF gas.

The figure portrays a range. The lower curve is derived from an in-house study made in early 1979; this study was conducted primarily for other purposes and the fuel gas estimates being a secondary objective were not determined rigorously. The upper curve starts with the maximum fuel demand from the Parsons Report and is somewhat arbitrarily declined similar to the lower curve.

Fuel gas estimated at this stage are necessarily coarse:

- Timing and scope of major PB oil production facilities have not been finalized.
- Facility requirements during the oil decline phase cannot be determined at this point.
- No account has been taken of possible enhanced/tertiary oil recovery at Prudhoe Bay.

(b) BTU content of fuel is determined by the specification of the existing equipment at about 900 BTU/SCF $\pm$ 10%. Future equipment can in principle be tailored to the fuel available, subject to the constraints discussed in the remainder of this response.

The dewpoint of a fuel is dictated by the lowest temperature expected to be encountered by that fuel. This temperature is,in turn,dictated by the distance of the fuel user from the fuel source, pipeline size and insulation, etc. In principle,a field fuel system can be designed to cater for a high dewpoint by installation of heating systems, adequate insulation, circulating systems (for shutdown periods) adequate scrubbers (to catch liquid slugs) etc. However, such a system would obviously cost more than a simple system and would also run contrary to operational preferences for simplicity and reliability.

Composition is dictated primarily by the need to meet the conflicting requirements of heating value and dewpoint. An additional constraint for turbine fuel is that of flame stability, which is impaired by higher CO_2 contents. The 75% CO_2 local turbine fuel in the Parson's Report is felt to be close to this limit.

Other considerations for field fuel systems are simplicity, reliability and economics. Ideally there would be one quality of field fuel supplied through one cost effective system from several backup sources which can readily furnish fuel under any likely operating mode. The existing system comes close to meeting these criteria.

Simplicity, reliability and economics go hand in hand. A simple system is likely to be reliable. Reliability will reduce costly outages and enhance economics. It must be stressed that any field fuel system must be capable of sustaining oil production during periods when the gas conditioning plant or gas pipeline are shut down.

Miscellaneous questions ...

- When you speak of BTU's do you mean gross or net BTU's? When calculating the BTU values of various compositions of gases, what assumptions do you use with respect to the BTU content per mcf and per barrel for each hydrocarbon type (C_1 through C_5) What phase diagrams to you use for the following pure substances: C_1, C_2, C_3, C_4, C_5, CO_2, H_2O?

RESPONSE

In general one uses net BTU's (lower heating values) when discussing fuel usage, whereas gross BTU's (higher heating values) are used for sales streams in line with FERC and industry practice. The Parsons Report is consistent in this respect, and in some places quotes both.

BTU values are taken from the BTU tabulation in the API Technical Data Book, 3rd Edition. Phase diagrams for the Parsons Report were calculated based on Parsons modified version of the Soave Redlich Kwong method. It was recognized that the high proportions of CO_2 in most of the streams cast doubt on the validity of Parsons' equilibrium constants ., and various field gas and liquid samples were taken at Prudhoe Bay in an attempt to determine these constants experimentally. Revised constants were used and while these were the best data available, doubt remains as to their accuracy.

- How does the addition of CO_2 affect the dewpoint characteristics of any particular gas composition, assuming that the composition of the hydrocarbon portion of the mixture remains unchanged?

RESPONSE

Under thepressures, temperatures and compositions contemplated, increased CO_2 content will tend to reduce the dewpoint slightly.

- **On a very approximate basis, how much more money would it cost to isolate most of the produced ethane from the CO_2 rich stream during the conditioning process? Does this represent, say, the added cost of using a Sulfinol or similar process with required dehydration facilities? What other options are available?**

RESPONSE

This has not been addressed by Sohio. However, there are three reasons why there is no simple answer to this question:

- Any plant design is likely to be complex, especially in the finely tuned areas which are peripheral to the main process.
- The answer would depend on what is meant by "most" of the ethane.
- Depending on the motive for removing the ethane, the same objective may be achieved in another manner.

One could speculate that any such ethane removal may well cost less than a sulfinol plus dehydration process. This speculation is based simply on the observation that sulfinol would require dehydrating a 2BCFD stream whereas the CO_2 rich stream at approximately 400 MMCFD is considerably smaller.

- Has anybody done any work yet on the added cost of transporting a gas of 13% CO_2 compared to the cost of transporting a 1% or 3% mixture, holding the pipeline operating pressure constant?

RESPONSE

Northwest Alaskan Pipeline has done such a study, dated February, 1979, simultaneously with a Supplementary Report to the Parsons Study. The Northwest Study and the Parsons Supplementary Study compared the effects of 3% and 13% CO_2 on respectively, the pipeline and the conditioning plant. Two cases were presented for each of the higher CO_2 contents, one of which aimed at BTU equivalency of the sales stream with the 1% Parsons base case, while the other aimed at volumetric equivalency with the 1% base case.

Northwest chose to compare only the volumetric equivalency cases in their analysis, a choice which we have criticized. Since the primary objective of the venture is to deliver BTU's to the U.S. consumer, it makes sense to us to compare alternate cases on the basis of BTU parity. This latter approach also makes more economic sense since over the range in question (2.0-2.3 BCFD) the cost per MCF of transporting gas through the pipeline system is declining.

- General questions on the effect of CO_2 content on the hazards of carbonic acid and hydrate corrosion and erosion at the pipeline operating pressures and temperatures under consideration.

RESPONSE

Neither CO_2 corrosion nor hydrate formation can occur in the absence of free water. Gas containing CO_2 can be and is safely pipelined in the U.S.; the breadth of CO_2 contents currently being pipelined range from nominal amounts to almost pure CO_2.

The requirement that free water be absent is reflected in the stringency of anticipated pipeline dewpoint specifications in the range of $-40^{o}F$.

TABLE 1

ANTICIPATED PRODUCED GAS COMPOSITION

Component	Mol%
CO_2	12.6
N_2	0.5
C_1	74.2
C_2	6.5
C_3	3.5
C_4	1.6
C_5	1.1
	100.0

TABLE 2

PRODUCT DISPOSITION

(MBD except C_1 in MMSCFD)

Parsons Maximum Fuel Case

Component	Supply	Taps and Field Fuel	GCP Fuel	Sales Volumes
C_1	2077	184	6	1887
C_2	116	43	15	58
C_3	63	23	16	24
C_4	35	2	1	32
C_5+	28	-	-	28

Parsons Minimum Fuel Case

Component	Supply	Taps and Field Fuel	GCP Fuel	Reinjection	Sales Volumes
C_1	1936	54	17	1	1864
C_2	107	16	18	4	69
C_3	60	17	10	1	32
C_4	33	1	1	-	31
C_5+	27	-	-	-	27

FIGURE 1

ESTIMATED PRUDHOE BAY OIL PRODUCTION PROFILE

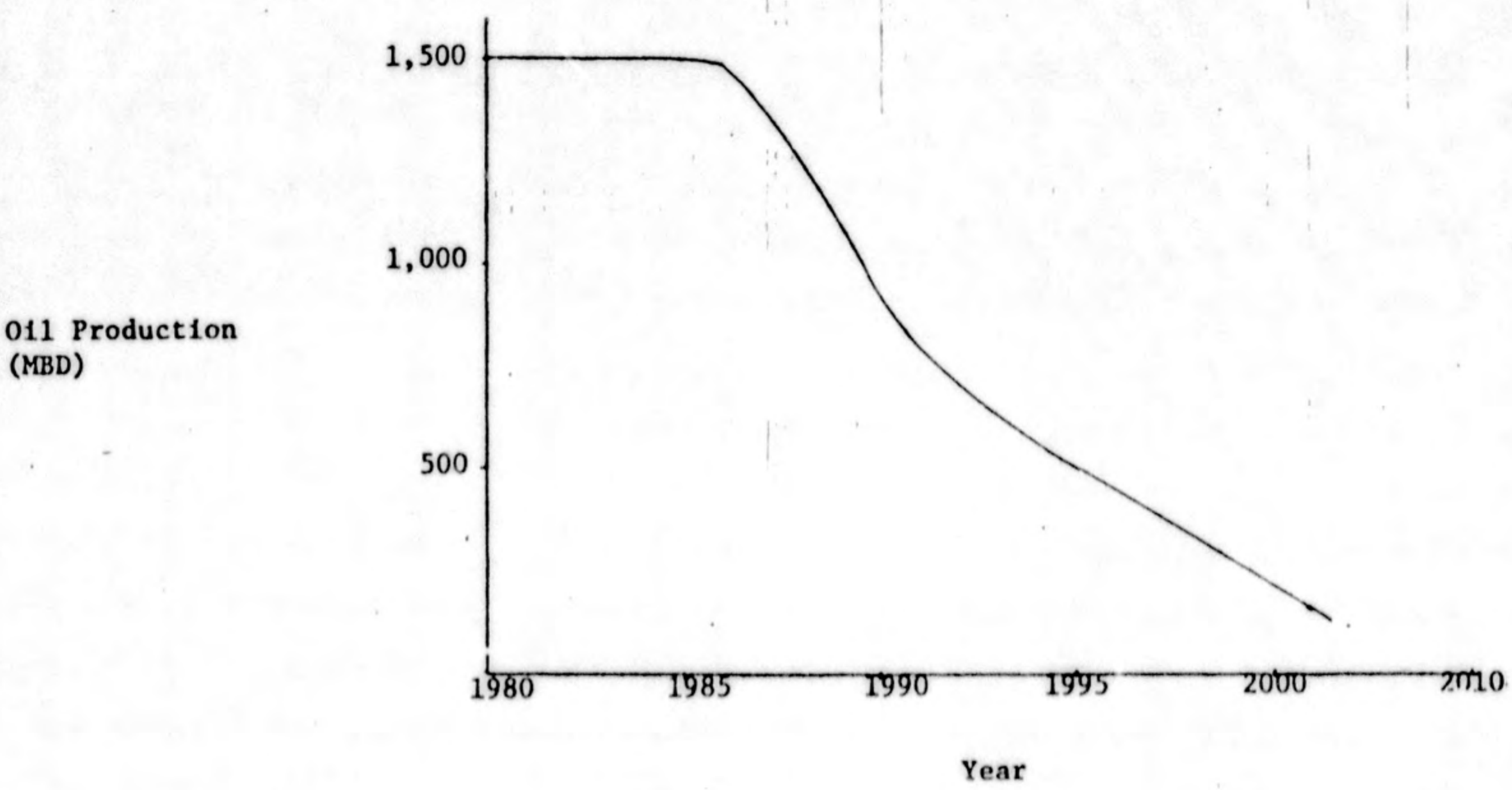
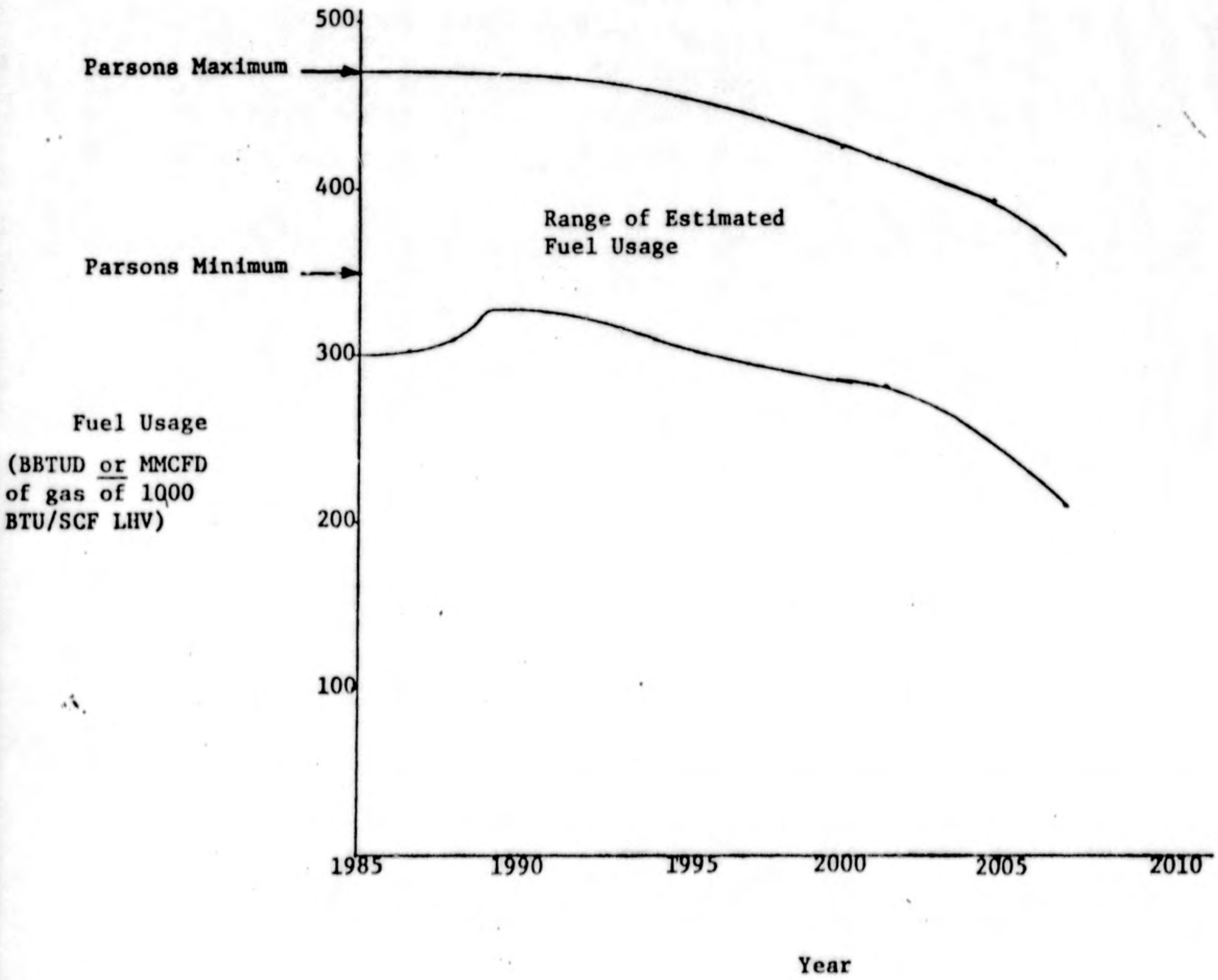


FIGURE 2

ESTIMATED PRUDHOE BAY

FUEL GAS USAGE INCLUDING GAS CONDITIONING PLANT



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COMMUNICATIONS PLANT
—GASLINE

ARCO Oil and Gas Company
Prudhoe Facilities Project Group
Bin 89
Pasadena, California 91109
Telephone 213 440 3674

W. S. Dickinson
Manager

January 3, 1980

Mr. Arlon R. Tussing
Arlon R. Tussing & Associates, Inc.
2710 Rainier Bank Tower
Seattle, Washington 98101

Dear Mr. Tussing:

On December 3, 1979, you sent a series of questions concerning Prudhoe Bay gas handling to both myself and K. R. Dickerson at Dallas. The attachment to this letter will serve to answer those questions for both of us.

If, after reviewing the answers provided, you feel that further clarification is required, we would be happy to arrange a meeting at Pasadena for this purpose.

Very truly yours,

W S Dickinson
W. S. Dickinson

WSD:el

Attachment

cc Connie Barlow
K. R. Dickerson

RESPONSES TO ARLON R. TUSSING
LETTER OF DECEMBER 3, 1979

Question #1:

(a) What is the expected composition and daily production rate of total raw gas (less whatever is reinjected to maintain reservoir pressures)? How much uncertainty accompanies anybody's best guesses about this matter?

Answer: The expected gas composition for the initial sales gas conditioning facility inlet was shown in the September 1978 report prepared by The Ralph M. Parsons Company (Vol. I, p. 6-2). There is not a great deal of composition difference between the gas cap gas and the oil rim gas (free plus dissolved); therefore, the uncertainty accompanying the composition of the initial inlet gas stream is not large. The maximum annual average gas offtake rate for the field is 2.7 billion SCF per day pursuant to Rule 15 of Conservation Order Number 145. There is no uncertainty regarding this field rule at this time.

(b) How is the daily production rate and composition expected to change through time, particularly when oil production begins to drop off? (Specifically, is there a significant difference in composition of solution gas versus gas cap gas, and is the ratio of produced volumes of solution and gas cap gas affected by time?)

Answer: Once gas sales commences, the annual average gas offtake rate for the field will remain essentially constant with time at the maximum of 2.7 billion SCF per day to meet the planned nominal 2.0 billion SCF per day to the gas pipeline plus field fuel gas requirements and conditioning plant shrinkage. The composition of the gas production will become slightly leaner in NGL hydrocarbons as the field is depleted. The gas production will tend to approach the composition of the gas cap gas as the oil rim gas rate decreases and the proportion of gas cap gas increases. For instance, after about 25 years of gas sales the mole fraction of methane in the gas stream is expected to be about 3% higher than it was initially, and the mole fraction of NGL components is expected to be about 17% lower.

(c) Of these produced hydrocarbons, how much of each will be consumed to meet North Slope total fuel requirements? How will field fuel requirements change through time as oil production and transportation drop?

Answer: The Ralph M. Parsons Company September 1978 Report shows the initial composition of the central gathered field gas production and the disposition of the products that could be made initially from the sales gas conditioning facility (Vol. I, Section 6). The facility recovery is such that each component is shown in volume of avails and the BTU content of butane and lighter components. The field fuel requirements shown in the report are those to meet what was considered to be a maximum fuel requirement

(363 MMBTU/D) and a minimum fuel requirement (131 MMBTU/D) that might occur during some period of sales gas conditioning facility operation. Table 6-4 (Volume I) of The Ralph M. Parsons Company report shows the fuel stream compositions for the maximum field fuel requirement case. We currently believe that this is the more representative of the two cases for the majority of the gas conditioning facility life considering the most likely scenario of field development events and their timing. Under this scenario, field fuel requirements will hold relatively constant (± 10 - 13%) through time even though oil production and transportation requirements will drop.

(d) Hence, what volumes of methane, ethane, propane, butane, and pentanes+ will actually be available through time for delivery to some transportation system?

Answer: The volume of methane and NGL avails can be determined over time using the gas production and composition data referred to in Parts (a) and (b) above and the fuel requirements data referred to in Part (c) above. Products will be available in relation to recovery percentages that can be derived from Section 6 of Volume I of The Ralph M. Parsons Company reports (maximum field fuel case).

Question #2:

(a) For each pipeline pressure specification under consideration (1080 psig in Canada and 1260, 1440, and 1680 psig in Alaska), what temperature and pressure combinations constitute "upset" conditions, "normal minimum operating" conditions, or whatever terminology you use to signify the most troubling conditions (from the standpoint of condensation) that gas may encounter during the journey south?

Answer: The 1260 gas pipeline hydrocarbon dewpoint specification of -10°F at 1000 psig is assumed to be the troublesome and minimum condition that would be encountered as the gas travels through permafrost and discontinuous permafrost areas of the pipeline route. It is our understanding no condensing is expected in the Canadian 1080 psig line operating south of permafrost areas. The pipeline is intended to operate only in the single gas phase and would be operated to not encounter any condensation for possible two-phase operation. A pipeline operated at 1680 psig per Gas Arctic FERC filing had a gas hydrocarbon dewpoint of -10°F at 1200 psig. We do not know of an actual 1440 psig design but a -10°F at 1100 psig dewpoint was used in the Parsons report.

(b) Are the phase characteristics of probable gas mixtures such that the temperature and pressure combinations of most concern represent the lowest temperature and the lowest pressure likely to be encountered?

Answer: The area of concern is that which represents the lowest temperature and the lowest pressure that is likely to be encountered in the pipeline. Phase characteristic curves were developed for the probable gas that would be delivered to this pipeline and shown in the Parsons report.

(c) At least in the documents we have seen, there appears to be some difference of opinion on the answer to the above questions. For example, I believe Exxon has indicated that a 0 degree F upset temperature is appropriate in Alaska, and 30 degree F for the sections south of Whitehorse, while Sohio lists -10 degrees and +42 degrees respectively. With regard to pressure, Sohio appears to assume that a 1680 psig line will have a 1300 psia upset while ARCO assumes a 1200 psia upset. What might be the reasons behind these differences of opinion (e. g., compressor spacing, which compressor station is assumed out, duration of outage, immediacy of response in reducing throughput, normal station outlet temperature of the gas, ambient air or ground temperatures, etc.)?

Answer: There have been numerous comments made on the phase characteristics of the probable gas that would be delivered to a gas pipeline from Prudhoe. These comments were made by various parties with different assumptions and bases for calculations. The most consistent and one with a common base of assumption is that which is reported in the Parsons September 1978 report. The report makes no attempt at speaking to the conditions on the gas pipeline. It only reflects the gas specification as set out by the gas pipeline consortium, i. e., -10°F @ 1000 psig for 1260 psig, -10°F @ 1200 psig for 1680 psig, and the assumed -10°F @ 1100 psig for the Parsons study at 1440 psig. The Sales Gas Conditioning Facility was designed to process the produced gas to any of these controlled conditions. No attempts in the Sales Gas Conditioning Study were made at predicting the gas pipeline configuration such as compressor spacing, outages, ambient air, or ground temperatures, etc.

Question #3:

(a) What is the lowest TAPS inlet temperature that might safely be considered? What are the physical considerations that determine this (i. e., what worst case assumptions about ambient air temperature and duration do you make with regard to TAPS shutdown or pump station outage?). Are there absolute physical limitations on the minimum inlet temperature, or is it primarily an economic matter of the costs and aggravation of frequent pigging?

Answer: Limitation on TAPS inlet temperature would primarily be a function of effect on throughput and wax deposition. Lower temperatures significantly reduce TAPS capacity due to increased friction and static head losses. TAPS has been designed for restart after a three-week shutdown during the winter with -40°F and 20 MPH wind. The economic limitations on the minimum inlet temperature to TAPS deal with the approximate 100°F wax point of the crude.

(b) Approximately how much would it cost to add chilling facilities to bring the inlet temperature down to the level indicated above?

Answer: The Ralph M. Parsons Company report of September 1978 discusses the facilities and costs needed to chill the crude to about 112°F to allow blending of the butanes and gasolines for delivery to TAPS during high oil production. The cost is approximately $60MM based on 1978 dollars. Cooling the oil in TAPS to temperatures below waxpoint of about 100°F has been considered impractical due to the huge heat exchanges required and the problems of wax deposition in the exchangers and subsequent wax disposal problems.

(c) Are light hydrocarbons such as butane and pentane useful in reducing oil pipeline viscosity and friction, thereby increasing throughput capabilities with existing equipment? Would one part of additional butane, therefore, require leaving behind maybe less than one part of heavier hydrocarbons if the line were already functioning at capacity? Given the likely timing of gas production for sale, and given the likely oil production scenario, is it a realistic prospect that throughput limitations of the line and pump stations might make it impossible to ship butanes and pentanes separated from the raw gas unless oil shipments are cut back?

Answer: Addition of light hydrocarbons to the crude would reduce viscosity and friction losses. Drag reducing chemicals are being used in TAPS to reduce friction losses and provide for meeting stated shippers' needs for a sustained 1,515 MBD capacity in 1980. TAPS capacity can be further increased in the future, by increments, to the ultimate 2.0 MBD. To expand TAPS above current capacity levels requires construction of new pump stations and additional pumps at existing stations. The TAPS owners would have to decide upon future expansions based on shippers' requirements. If the oil pipeline were operating at full capacity, the addition of butanes and pentanes would require the cutback of approximately one to two barrels of oil for every two barrels of butanes and pentanes+ added.

(d) What volumes of oil production for each year of field operation should one assume when trying to ascertain TAPS throughput capabilities? To your knowledge, did the other parties commenting on the ability of TAPS to carry light hydrocarbons assume similar production schedules?

Answer: Table I shows a crude oil production forecast by year for the Prudhoe Bay Unit Main Area Sadlerochit reservoir. We submitted this forecast to the Division of Oil and Gas (May 17, 1977) to become part of the record of the State of Alaska Prudhoe Bay Unit Pool Rules Hearing that was held on May 5, 1977. The peak rate of 1500 MBOD was used as the basis for the Ralph M. Parsons report. For purposes of ascertaining TAPS NGL throughput capabilities, production (volumes and vapor pressures) from other reservoirs, both at the Prudhoe Bay Unit (Sag, Shublik, West End), and elsewhere (Kuparuk, and possible Beaufort Sea reservoirs) should be added to the forecast. This additional production will extend the period of plateau oil rate.

We have no knowledge of what production schedules may have been used by other parties in their studies to determine the ability of TAPS to carry light hydrocarbons.

(e) Given the oil production scenario noted above and the volume of butanes and pentanes implied in your answer to Question 1(d), how much butane and pentane could be transported through TAPS, (noting time as a variable) at the minimum inlet temperature you have specified in your answer to Question 3(a) at the existing TAPS inlet temperature?

Answer: Given the oil production forecast shown on Table I, all of the pentane+ production can be shipped with the crude with a modest chilling to meet vapor pressure limitations of about 5°F depending on crude volume. All of the butanes can be shipped with the crude during current high oil production from the field when subcooled to 112°F. Low oil throughput rates in future years will cause vapor pressure limitations and cutback in butane shipping to meet the 11.1 psia limit at destination storage tank temperatures.

(f) Currently, what (if anything) is being done at Valdez to ensure that oil shipped through TAPS (at 140 degrees F and 14.7 psia) meets tanker and lower 48 storage specifications for vapor pressure? How would those activities change if the TAPS inlet temperature were lowered to the point mentioned in your answer to Question 3(a)?

Answer: Nothing is done at Valdez to control temperature of the crude. At the 1.5 MBD current throughput rate, oil arrival temperature is about 85°F at Valdez. EPA rules for floating roof tanks limit vapor pressure to 11.1 psia at storage tank temperature. That limit is not approached by the current north slope crude vapor pressure (14.7 psia at 142°F) and receipt temperatures.

If TAPS inlet temperature were lowered there would be no need to change the current practice at Valdez. Oil below 40°F has been loaded on ships.

Question #4:

(a) Given the likely oil and gas production schedules you have indicated previously, how will the required amounts of field fuel change through time?

Answer: Field fuel requirements are expected to hold relatively constant ($\pm$10 - 13%) through time even though oil production and transportation requirements will drop (reference Answer 1(c)).

(b) What are the physical and economic principles that limit flexibility in choosing the BTU content, dewpoint, and overall composition of field fuel? (e.g., the need to ensure a certain level of precision in combustion characteristics, dewpoint limitations imposed by exposure to ambient air temperatures en route, BTU specifications of existing equipment, and the cost of changing or expanding that range.)

Answer: The existing gas turbines in the field were designed to operate with a fuel of approximately 875 BTU per cubic foot net. These turbines generally have a plus or minus 10% fuel BTU variation capability without modification of their combustion system. The Sales Gas Conditioning Study considered this when designing a minimum of 825 BTU net fuel gas for delivery from the facility during its operation. The hydrocarbon dewpoint limitations of the field fuel were considered when developing the fuel gas blends. They were maintained as low as possible while still utilizing the high CO_2 blends from the Sales Conditioning Facility. Only propane was used for blend stock. Blending butane with the high CO_2 gas resulted in a hydrocarbon dewpoint higher than that considered acceptable in the field fuel system.

Miscellaneous Questions . . .

Question: When you speak of BTU's, do you mean gross or net BTU's? When calculating the BTU values of various compositions of gases, what assumptions do you use with respect to the BTU content per mcf and per barrel for each hydrocarbon type (C_1 through C_5). What phase diagrams do you use for the following pure substances: C_1, C_2, C_3, C_4, C_5, CO_2, H_2O?

Answer: The BTU of various streams reported in the Sales Gas Conditioning Study by Ralph M. Parsons were based on net BTU calculations. The residue gas to the gas pipeline were reported on a gross BTU basis compatible with FERC specifications. Gas Processors Association standards for BTU contents of hydrocarbons were used in our calculations of the various stream BTU contents. Gas Processors Association K&H Mod. II Program which uses Soave Redlich-Kwong equations were used for all equilibrium and material heat balances made in the process study instead of individual component phase diagrams.

Question: How does the addition of CO_2 affect the dewpoint characteristics of any particular gas composition, assuming that the composition of the mixture remains unchanged?

Answer: The addition of CO_2 to a fixed hydrocarbon content gas tends to slightly lower (less than $5^{\circ}F$) the hydrocarbon dewpoint of the resultant mixture.

Question: On a very approximate basis, how much more money would it cost to isolate most of the produced ethane from the CO_2 rich stream during the conditioning process? Does this represent, say, the added cost of using a Sulfinol or similar process with required dehydration facilities? What other options are available?

Answer: To isolate most of the produced ethane would require a different process. The only process studied at this time that would do so would require a Sulfinol CO_2 removal system followed by an expander cryogenic type, ethane plus recovery facility. Based on studies made in mid-1974 for the Gas Arctic Group, this cost would add approximately 300 to 400 million dollars more to the cost of the Sales Gas Conditioning Facility.

Question: Has anybody done any work yet on the added cost of transporting a gas of 13% CO_2 compared to the cost of transporting a 1% or 3% mixture, holding the pipeline operating pressure constant?

Answer: Northwest Alaskan Pipeline prepared a report on the added cost for transporting a 3% CO_2 gas and a 13% CO_2 gas versus the base filing of 1% gas. No doubt others have made such studies, but to our knowledge none have been published.

Question: General questions on the effect of CO_2 content on the hazards of carbonic acid and hydrate corrosion and erosion at the pipeline operating pressures and temperatures under consideration.

Answer: Numerous comments have been made on the subject of pipeline corrosion. Sohio, Exxon, and ARCO prepared statements and filed such statements with FERC. Northwest Alaska Pipeline did likewise. These responses fairly well speak to the problems of corrosion or erosion in the pipeline when operating with CO_2. In essence, no corrosion is possible as long as the gas is above water dewpoint temperature, so the gas must be maintained with a low water dewpoint.

WSD:ARC:el
1-3-80

TABLE 1

OIL PRODUCTION FORECAST
MAIN AREA SADLEROCHIT RESERVOIR
PRUDHOE BAY UNIT

(From the record of the State of Alaska Pool Rules Hearing of 5-5-77)

YEAR OF PRODUCTION	OIL RATE MBOD
1	1200
2	1350
3	1500
4	1500
5	1500
6	1500
7	1500
8	1500
9	1360
10	1020
11	910
12	850
13	860
14	840
15	790
16	710
17	570
18	450
19	410
20	330
21	300
22	250
23	260
24	220
25	190
26	160
27	140
28	130
29	110
30	100

PLEASE NOTE: THE PRECEDING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT.

PLEASE NOTE: THE FOLLOWING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT

CONDITIONING - GASLINE ①

Meeting w Exxon — Houston. Jan 15, 1980

Tussing & Barlow

Judd Miller (mgr under Ray Booth?)
Ed Travis (lawyer)
Charlie Parr (engineer)
Jimmy Shanks (engineer)
Harold Galloway (head of gas engineering production dept.)

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overall vibes:

① pc - very interested in getting the word out on pc. Exxon can't say because looks like trying to stifle competition.
Why doesn't AK realize gas conditioning / pc is capital intensive? Not long-term jobs.

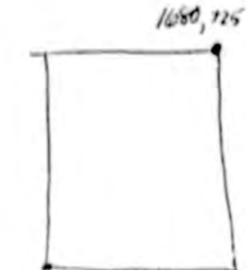
② liquids sale - emphasized Exxon has sold liquids. State has to deal w. purchasers.

③ condit pl. ownership - interested in no subtle hints or references made that Exxon owns or has anything to do w conditioning

---

pressure -

1680, 125 "normal operating window" "maximum allowable operating P"



1350 +10°

1680psi at +25° = "discharge P & T"
1350 at +10° = "normal operating suction P & T"
(1680 ÷ 1.25) (1.2→1.3 is "compression ratio" common to most pipelines)

upset = 1250 at -10°

~~1260 at 25° = discharge~~
~~1000 at 10° = normal operating suction P&T~~

For "operating windows" of 1260 & 1440 use same temps as above and divide discharge pressure by 1.25.

FERC's role is to be concerned about operating pressures; let pipeline deal with safety specs for upset.

upset conditions: assume 1 compressor down & as a last resort reduce thruput when that happens; but try not to.
What if some sections above ground & shut-in in winter. Appeared hadn't thought about it, but pipeline could handle. Not insurmountable. Sure, liquid forms, but figure it happens once every 3 years so don't worry. Have to pig periodically anyway & have scrubbers & drains.

Pipe diam & pressure - (most gas pipeline companies were for 1680 not 1260)

Exxon's position - advocated 2160 in 42". The problem is when have 48" at high pressures. No line like that before. Exxon believes over 1260 at 48" need crack arrestors. Cheaper to insert crack arrestors rather than super engineer pipe (cost pipe hi quality). The problem is not that the pipe bursts, but somebody runs into it w a tractor & the crack runs 2-3 miles to the next valves. The concern in AK is it happens in a remote area at a time of year when hard to fix quick. Canada specs not so great — figure more cost effective to let crack once in a while & fix quick than do super hi quality.

48" pipe hasn't been built before.

"Great Lakes pipeline" (Mich-Wisc) is 1440 at 36" (has had probs with long shear fracture (ductile fracture)

Exxon & Shell have a 2000° 36" somewhere in North Sea?

Exxon gave us a chart showing Inlet Gas, Fuels, Sales Gas all the way thru in barrels. Only diff fr. Parsons is excludes ~~recently~~ TF&U. uses gross BTU's. Based on Parsons max field fuel case (which includes fuel for artificial lift, waterflood, etc.)

Fuel - TAPS fuel gas line goes to first 4 pump stations. Buried. Dewpt comp = 800 psi at −40° (so very light HC comp.)

IMP - 1 producers have right to take whatever volume of gas need for field fuel & TAPS before goes to conduit. This is why ~~prod~~ Exxon conference sold all gas liquids. So producers take care of early TAPS & field fuel. Includes 307 billion BTU/d more field fuel on chart because that's what gas production causes.

2-phase flow — is in gathering lines now. Sized really large tho so won't be a problem. Also, very large separators at the end serve same function as slug catchers.

Efficiency of flow (cost of pumping & compression) —
① most efficient = liquids line like TAPS
② much less efficient = gas line
③ far less efficient = 2 phase flow.

Use of CCP compressors for gas sales boost vs. new "stand alone" facilities: Exxon can't answer when & how decide. Working interest owners will have to decide whether to sell CCP to gas purchasers. Seems to think gas purchasers would be wise to install new machinery that would be designed for gas density because so much more fuel efficient. Didn't indicate any reason why producers might not want to sell CCP.

Conduit re-design
B&M report throwing in a "de-ethanizer" (B&M doesn't list costs) & using all propane doesn't make sense. Must re-design whole process, if different goals in order to optimize. Can't just add facilities. Plus if goal is to separate liquids for liquids line it becomes a "processing plant" not conditioning.
If State wants all propane can't use Selexol because must then design a process that leaves a ~~ ~~ totally lean CO_2 flash — not something that must be enriched & used as fuel.

Carbonic acid / corrosion probs

absolutely no concerns of carbonic acid in AK sections because if any water it's found as ice; and acid can only form if H_2O is in water phase in presence of CO_2.

Hydrates - is stupid to worry about carbonic acid corrosion because if any problem at those low temp will be a hydrate prob. And unlike corrosion which eats away slowly, hydrates cause instantaneous probs. (Hydrates = H_2O ice crystal with HC molecules inside which gunk things up. CO_2 has nothing to do w hydrates).

carbonic acid may be problem in Canada in the 40° Temp line there if Canada adds gas with hi H_2O. (Exxon appeared worried that Canada might do that.) ← Appears H_2O specs for line in Canada are higher. Pipeline in Canada can be "looser" about acid & crack probs because can fix easily, unlike remote AK.

Production - even when gas production starts will not drill any specifically gas cap wells. Figure gas cap will expand down & inevitably enter oil well production. Depending on how soon that happens, may need to perforate oil well higher up if need to produce more gas.

"gas cap gas", "oil", "solution gas" & "condensate" are artificial terms. Really only account for "oil" & "gas".

For terms of ownership assume diff amts of gas cap gas vs. solution, but all comes out of same wells together.

Arco & Exxon say solution & gas cap NGL content essentially the same (because own more gas cap gas), while Sohio probably says solution is richer because owns more.

EXXON

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Figure 2

Comments on Data Used in Parsons Process Design*

The physical and thermodynamic data used in the design of the SGCF is generally Parsons' normal design data. Where possible, vapor-liquid equilibruim data were verified with test data from the Prudhoe Bay facilities. Allied Chemical Corporation's proprietary data were used for the SELEXOL CO_2 removal system. These SELEXOL data were incorporated in Parsons' process simulator which was used to model plant configurations.

Parsons' Soave-Redlich-Kwong vapor-liquid equilibrium data and associated computer-generated enthalpy data were used for all hydrocarbon and CO_2-hydrocarbon multicomponent systems. Correlations of the available data with tests of operating Prudhoe Bay facilities were found to be in good agreement; however, portions of the process are at temperatures, pressures, and compositions where no current operations exist and field verification of data in these areas could not be made.

The Elliott Company pressure-enthalpy diagram for Propane was used for the refrigeration system design. Heating values and general physical data are from or in agreement with the Engineering Data Book of the Gas Processors Suppliers Association and, where applicable, with the API Data Book.

* These comments are also made in Volume II, Section 1.1.6, Page 1-7 of the Parsons Study.

gjm#2/B-28
ae

PHYSICAL CONSTANTS FOR CALCULATIONS

Data from NGPSA - Ninth Edition - 1972

	METHANE	ETHANE	PROPANE	ISO BUTANE	N BUTANE	ISO PENTANE	N PENTANE	N HEXANE	N HEPTANE	N OCTANE	N NONANE	N DECANE	CARBON DIOXIDE	HYDROGEN SULFIDE	NITROGEN
Atomic Formula	CH_4	C_2H_6	C_3H_8	C_4H_{10}	C_4H_{10}	C_5H_{12}	C_5H_{12}	C_6H_{14}	C_7H_{16}	C_8H_{18}	C_9H_{20}	$C_{10}H_{22}$	CO_2	H_2S	N_2
Molecular Weight	16.043	30.070	44.097	58.124	58.124	72.151	72.151	86.178	100.205	114.232	128.259	142.286	44.010	34.076	28.013
Boiling Point °F, 14.696 psia	-258.69	-127.48	-43.67	10.90	31.10	82.12	96.92	155.72	209.17	258.22	303.47	345.48	-109.3	-75.6	-320.4
Critical Temperature °F	-116.63	90.09	206.01	274.98	305.65	369.10	385.7	453.7	512.8	564.22	610.68	652.1	87.9	212.7	-232.4
Critical Pressure lbs. abs	667.8	707.8	616.3	529.1	550.7	490.4	488.6	436.9	396.8	360.6	332	304	1071	1306	493.0
Critical Volume cu.ft / lb mol	1.590	2.370	3.250	4.208	4.080	4.899	4.870	5.929	6.924	7.882	8.773	9.661	1.505	1.564	1.440
Critical Density lbs./cu ft	10.09	12.69	13.57	13.81	14.25	14.73	14.81	14.53	14.47	14.49	14.62	14.73	29.24	21.79	19.46
LIQUID DENSITY DATA															
Specific Gravity at 60°F	0.3	0.3564	0.5077	0.5631	0.5844	0.6247	0.6310	0.6640	0.6882	0.7068	0.7217	0.7342	0.827	0.79	0.808
API Gravity at 60°F	340	265.5	147.2	119.8	110.6	95.0	92.7	81.6	74.1	68.7	64.6	61.2	39.6	47.6	43.8
lbs./gal. at 60°F in air	2.5	2.962	4.223	4.686	4.865	5.199	5.251	5.526	5.728	5.883	6.008	6.112	6.89	6.58	6.73
Fraction of mol/lb	.06233	.03326	.02268	.01720	.01720	.01386	.01386	.01160	.00998	.00875	.00780	.00703	.02272	.02935	.03570
Gallons of liquid/lb mol	6.4	10.12	10.42	12.38	11.93	13.85	13.71	15.57	17.46	19.39	21.32	23.24	6.38	5.17	4.16
lb mols/gal. of Liquid	0.156	.0988	.0960	.0808	.0838	.0722	.0729	.0642	.0573	.0516	.0469	.0430	.1567	.1934	.2404
VAPOR DENSITY DATA STP (Assuming perfect gases)															
Specific Gravity of Air = 1	0.5539	1.0382	1.5225	2.0068	2.0068	2.4911	2.4911	2.9753	3.4596	3.9439	4.4282	4.9125	1.5195	1.1765	0.9672
lbs/1000 cu ft	42.28	79.24	116.20	153.16	153.16	190.13	190.13	227.09	264.05	301.01	337.98	374.94	115.97	89.79	73.82
Ratio Gas to Liquid Volume	443	280.5	272.51	229.30	237.98	204.93	207.00	182.37	162.56	146.45	133.18	122.13	444.8	548.7	682.7
cu. ft. of Gas/gal. Liquid	59	37.5	36.43	30.65	31.81	27.39	27.67	24.38	21.73	19.58	17.80	16.33	59.5	73.5	91.3
Combustion Data MBTU/gal.	59.829	65.998	91.065	99.022	102.989	108.790	110.102	115.060	118.668	121.419	123.613	125.444	–	–	–
Gross Heat BTU/cu ft	1009.7	1768.8	2517.5	3252.7	3262.1	4000.3	4009.6	4756.2	5502.8	6249.7	6996.5	7742.1	–	637	–
Abs. V.P at 100°F psia	(5000)	(800)	190	72.2	51.6	20.44	15.570	4.956	1.620	0.537	0.179	0.0597	(1200)	394.0	–

NOTES:

1 lb. mol. = 379.495 cu. ft. at STP
GPM 150 M.W. Oil = mols./hr. x 0.393
GPM 200 M.W. Oil = mols./hr. x 0.4795
GPM 180 M.W. Oil = mols./hr. x 0.437

7.480 gal. = 1 cu. ft.
Mol. Wt. of Air = 28.964
1 gal. of Water Weighs 8.328 lbs.
MSCFD = Mols./hr. x 9.10788

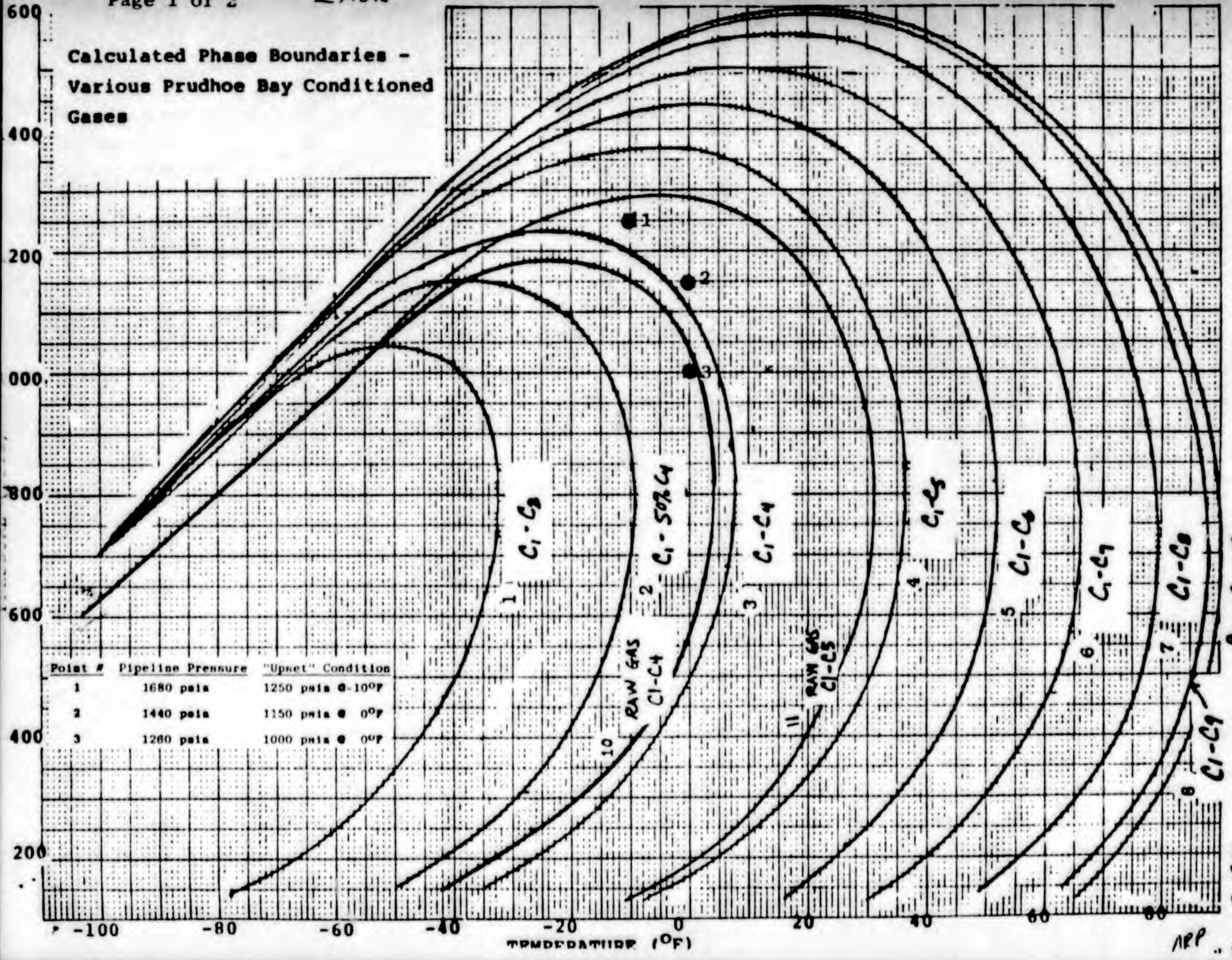
1000 cu. ft. of Air Weighs 76.32 lbs.
1000 cu. ft. of Gas = 2.63508 mols. Perfect Gas, STP
Values in Parentheses - Approximate

DICK BAXTER 9-16-72

PARSON'S used this for higher heating value BTU's. Note: Exxon will confirm w Parsons & send me lower heating value chart too.

FIGURE 1
Page 1 of 2

EXXON



Point #	Pipeline Pressure	"Upset" Condition
1	1680 psia	1250 psia @ -10°F
2	1440 psia	1150 psia @ 0°F
3	1260 psia	1000 psia @ 0°F

EXXON

Same HC mixture more CO_2

VARIOUS PRUDHOE BAY CONDITIONED GAS COMPOSITIONS
(Mole Percent)

Component	Unconditioned Separator Off-Gas	1 C_1-C_3	2 C_1-50% C_4	3 C_1-C_4	4 C_1-C_5	5 C_1-C_6	6 C_1-C_7	7 C_1-C_8	8 C_1-C_8	9 C_1-C_9	10 Off-Gas C_1-C_4
N_2	0.484	0.564	0.559	0.554	0.551	0.550	0.549	0.549	0.549	0.549	0.488
CO_2	12.659	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000	12.773
C_1	74.706	87.053	86.296	85.554	84.964	84.818	84.742	84.695	84.679	84.676	75.382
C_2	6.428	7.491	7.426	7.362	7.311	7.299	7.292	7.288	7.287	7.287	6.486
C_3	3.340	3.892	3.859	3.826	3.799	3.793	3.789	3.787	3.786	3.786	3.370
i-C_4	0.450	--	0.260	0.515	0.512	0.511	0.511	0.510	0.510	0.510	0.454
n-C_4	1.038	—	0.600	1.189	1.181	1.179	1.178	1.177	1.177	1.177	1.047
i-C_5	0.217	—	--	—	0.247	0.247	0.246	0.246	0.246	0.246	--
n-C_5	0.383	—	--	--	0.435	0.435	0.434	0.434	0.434	0.434	--
C_6	0.148	--	--	--	--	0.168	0.168	0.168	0.168	0.168	--
C_7	0.081	—	—	--	--	--	0.091	0.092	0.092	0.092	--
C_8	0.047	—	--	--	--	--	--	0.054	0.054	0.054	--
C_9	0.016	--	—	--	--	--	—	--	0.018	0.018	--
C_{10}^+	0.003	--	—	--	--	--	—	--	—	0.003	--
Molecular Wt.	22.7	18.5	18.8	19.2	19.5	19.7	19.8	19.8	19.8	19.9	22.2
Heating Value (Btu/cf*)	1027	1095	1113	1131	1150	1156	1160	1163	1164	1164	996

*Gross, Wet, Actual @ 60°F., 14.73 psia

3-7-78

note: Tables here assume no traces of heavier HC's for each column, but in reality will be traces.

Exxon translation of Parsons into barrels

gross BTU's used here.

uses Parsons Max Field Fuel Case

MATERIAL BALANCE

PRUDHOE BAY CONDITIONING PLANT

Parsons 1440 psi P/L -10°F at 1100#

1260 psi P/L -10°F at 950#

	Inlet Gas		Outlet Streams: Field Fuel		Plant Fuel		Del. to TAPS		Sales Gas			Total Out
	Mol %	MB/D	Mol %	MB/D	Mol %	MB/D	Mol %	MB/D	Mol %	MB/D		MB/D
C_1	74.00	-	30.12	-	2.42	-	-	-	90.72	-		-
C_2	6.53	111.4	18.25	39.2	9.18	14.5	-	-	4.41	58.1	58.1	111.8
C_3	3.52	61.8	10.14	22.5	10.31	16.7	-	-	1.72	23.3	23.3	62.5
iC_4	0.50	10.3	0.30	0.8	0.51	0.3	0.17	-	0.59	9.4	8.8	10.5
nC_4	1.18	23.7	0.35	0.9	0.08	0.1	5.84	1.4	1.41	21.9	20.3	24.3
iC_5	0.27	6.2	0.02	0.1	-	-	19.89	5.5	0.05	0.8	0.8	6.4
nC_5	0.49	11.4	0.02	0.1	-	-	42.13	11.5	0.01	0.2	0.2	11.8
C_6^+	0.36	9.7	-	-	-	-	31.98	10.20	-	-	T	10.2
CO_2	12.68	-	40.74	-	77.85	-	-	-	0.49	-		-
N_2	0.47	-	0.06	-	-	-	-	-	0.60	-		-
Total	100.00	234.5	100.00	63.6	100.00	31.6	100.00	28.6	100.00	113.8	111.5	237.5
											Δ=2.3*	
MMcf/D	2686		339		248		32		2071			2690
Btu/cu.ft.	1055		905		454		4350		1106			1059
Bil Btu/D	2834		307		113		139		2291	(80.3%)		2850

Notes: (1) Based on Parsons' September 1978 Study, maximum field fuel case.
(2) Existing Field Fuel Gas Unit volumes are excluded.
(3) Btu balance is off less than 1%.

*Gas Vol will decrease to 2027 + 1.8% = 2063 MMcf/D or 0.4% less than 2071 MMcf/D

if add FFGU to this get Parsons figure for Field Fuel less TAPS

HRG:cg
9/28/79

SOURCE OF FIELD FUEL (already in use = 800-900 BTU gas)

PRUDHOE BAY CONDITIONING PLANT SO design 825 BTU gas to field

	LTS Vapor		DeC_2 Overhead		Flash Vapors		Propane		Total Field Fuel	
	Mol %	MB/D	Mol %	MB/D	Mol %	MB/D	Mol %	MB/D	Mol %	MB/D
C_1	76.88	-	42.33	-	19.66	-	-	-	30.12	-
C_2	5.59	0.9	28.13	20.0	15.32	17.6	6.41	0.9	18.25	39.2
C_3	1.95	0.3	0.41	0.3	7.58	9.0	91.58	12.9	10.14	22.5
iC_4	0.14	-	-	-	0.37	0.5	1.44	0.2	0.30	0.8
nC_4	0.24	-	-	-	0.56	0.8	0.56	0.1	0.35	0.9
iC_5	0.02	-	-	-	0.04	0.1	-	-	0.02	0.1
nC_5	0.03	-	-	-	0.05	0.1	-	-	0.02	0.1
C_6^+	-	-	-	-	-	-	-	-	-	-
CO_2	14.65	-	29.09	-	56.41	-	-	-	40.74	-
N_2	0.50	-	.04	-	0.01	-	-	-	0.06	-
Total	100.00	1.2	100.00	20.3	100.00	28.0	100.0	14.1	100.00	63.6
MMcf/D	25		112		181		22		339	
Btu/cu.ft.	939		935		694		2484		905	
Bil Btu/D	23		105		126		55		307	

HRG:cg
9/28/79

note: turbines to be designed for 400 BTU gas and heaters too (even the Parson's shows only 200 BTU gas for heaters) [straight out of LFF)

SOURCE OF PLANT FUEL

PRUDHOE BAY CONDITIONING PLANT

	Local Fuel Fractionator O.H.		Flash Vapors		Propane		Total Plant Fuel	
	Mol %	MB/D	Mol %	MB/D	Mol %	MB/D	Mol %	MB/D
C_1	1.46	-	19.65	-	-	-	2.42	-
C_2	9.04	12.1	15.30	1.4	6.41	0.9	9.18	14.5
C_3	2.29	3.2	7.59	0.7	91.58	12.8	10.31	16.7
iC_4	-	-	0.37	-	1.44	0.2	0.51	0.3
nC_4	-	-	0.55	0.1	0.56	0.1	0.08	0.1
iC_5	-	-	0.06	-	-	-	-	-
nC_5	-	-	0.06	-	-	-	-	-
C_6^+	-	-	-	-	-	-	-	-
CO_2	87.21	-	56.43	-	-	-	77.85	-
N_2	-	-	-	-	-	-	-	-
Total	100.00	15.3	100.00	2.3	100.00	14.0	100.00	31.6
MMcf/D	212		15		21		248	
Btu/cu.ft.	232		695		2484		454	
Bil Btu/D	49		10		52		113	

HRG:cg
9/28/79

PLEASE NOTE: THE PRECEDING PAGES WERE TREATED AS A UNIT IN THE ORIGINAL DOCUMENT.