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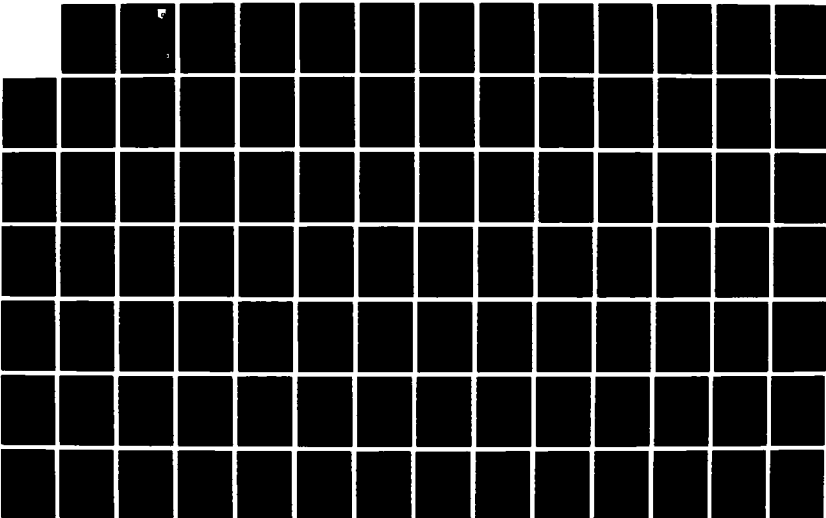
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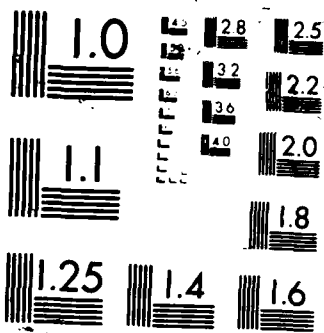
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VOLUME II



TURBINE FUELS FROM TAR SANDS BITUMEN  
AND HEAVY OIL

VOL II - Phase III. Process Design Specifications for a  
Turbine Fuel Refinery Charging San Ardo Heavy Crude Oil

A. F. Talbot, J. R. Swesey, and L. G. Magill

Applied Research and Development  
Sun Refining and Marketing Company  
P.O. Box 1135  
Marcus Hook, PA 19061-0835

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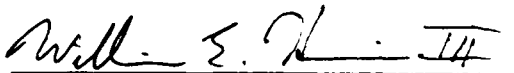
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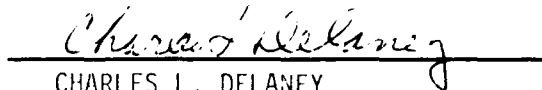
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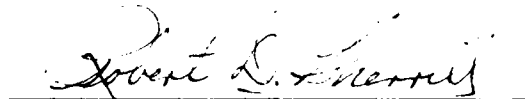


WILLIAM E. HARRISON III  
Fuels Branch  
Fuels and Lubrication Division  
Aero Propulsion Laboratory



CHARLES L. DELANEY  
Chief, Fuels Branch  
Fuels and Lubrication Division  
Aero Propulsion Laboratory

FOR THE COMMANDER



ROBERT D. SHERRILL, Chief  
Fuels and Lubrication Division  
Aero Propulsion Laboratory

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19. ABSTRACT (Continue on reverse if necessary and identify by block number) An engineering design was developed for a 50,000 BPSD grass-roots refinery to produce aviation turbine fuel grades JP-4 and JP-8 from San Ardo heavy crude oil. The design was based on the pilot plant studies described in Phase III - Volume I of this report. The detailed plant design described in this report was used to determine estimated production costs.					
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Special recognition is deserved by co-author Lloyd Magill, process design consultant, whose broad experience and many talents were essential in the successful completion of the entire refinery design effort from the preliminary stages to the final design package, including the development of final refinery process flow schemes, selection of auxiliary processes, and specification of major refinery equipment.

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## I. INTRODUCTION

The following process design specifications describe a new "grassroots" heavy oil upgrading refinery whose purpose is to produce principally JP-4 or JP-8 aviation turbine fuel from San Ardo (California) heavy crude oil at a crude charge rate of 50,000 barrels per operating day. The importance of the design is that it represents a strategic alternative for the production of U.S. military jet fuel using a domestic, rather than foreign, hydrocarbon resource. The feedstock is generally considered to be difficult to process due to its high viscosity and molecular weight, and high sulfur and nitrogen contents.

The purpose of the refinery design was to obtain an accurate assessment of construction and operating costs, which in turn would be used to determine the selling price of refinery fuel products to maintain an economically viable operation. The total installed capital cost for the refinery based on fourth quarter 1985 prices for a Salt Lake City, Utah, location was estimated to be \$1.126 billion. Components of the refinery operating costs also are listed later in this report. Results of the economic study to determine product fuel pricing appear in a separate report of the overall Phase III project <sup>1</sup>.

The distinguishing heavy oil upgrading process in the refinery design is the "hydrovisbreaking" of vacuum reduced crude oil in the presence of a coke-suppressing molybdenum-based additive. The operation converts reduced crude to lower molecular weight hydrocarbons which can be converted to high-quality fuels by conventional petroleum refinery hydrotreating and hydrocracking processes. The attractiveness of hydrovisbreaking as described here is its high conversion of the reduced crude with suppression of coke formation by a non-proprietary process using conventional processing equipment.

The design basis for each of the hydroprocessing units in the refinery including hydrovisbreaking, hydrotreating, and hydrocracking was developed

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<sup>1</sup> Talbot, A. F. et al, AFWAL-TR-87-2043 " Volume I - Phase III Pilot Plant Testing, Final Design, and Economics", August 1987.

from pilot plant experiments conducted for this project.

Most of the process design calculations were done with the "PROCESS" computer simulation program developed by Simulation Sciences, Inc. Comprehensive heat and material balances were computed for the six major refinery plants including the Crude Unit, Hydrovisbreaker, Naphtha Hydrotreater, Distillate Hydrotreater, Distillate Hydrocracker, and the Gas Plant. Consequently, their capital cost estimates are based upon detailed, major equipment specifications.

The other process units have very standardized "packaged" designs for which the literature and reputable vendors have provided detailed estimates of construction and operating costs. For these units it was, therefore, unnecessary to perform computer simulations and derive rigorous heat and material balances. However, the large compressor systems for the Hydrogen Plant and Hydrogen Purification Unit, were specified individually to permit a compressor manufacturer to determine their costs.

## II. OVERALL PROJECT DESCRIPTION

This design package was one element of a much larger project performed under U.S. Air Force contract No. F33615-83-C-2352, entitled "Turbine Fuels from Tar Sands Bitumen and Heavy Oil". The program of work was subdivided into the following three phases:

### Phase I - Preliminary Process Analysis

A series of case studies of potential upgrading and refining processes for the purpose of selecting a candidate conversion scheme.

### Phase II - Laboratory Sample Production

Preliminary bench-scale experiments with a variety of heavy crude oil and tar sand bitumen feedstocks to demonstrate the candidate conversion scheme, to determine overall processing requirements, and to supply small product samples for evaluation.

### Phase III - Pilot Plant Studies, Final Design and Economics

Pilot plant testing of a specific heavy crude oil, i.e., San Ardo heavy crude, final process design of a commercial-scale refinery, and an economic evaluation of the refinery capital construction and operating costs. Phase III entailed scale-up from preliminary work to demonstrate operability, to generate a design basis, to project fuel product costs, and to provide larger representative fuel samples.

The overall objective of the program was to assess the costs, yields, and physical and chemical characteristics of aviation turbine fuels made from U.S. domestic tar sands bitumen and heavy crude oil. The program required that conversion of these low grade feedstocks to high quality finished fuels be accomplished by commercially viable upgrading and refining processes to achieve product slates emphasizing either JP-4 or JP-8 turbine fuel, or a mixture of transportation fuels.

### III. REFINERY PROCESS DESCRIPTION

The basic refinery process flow scheme includes twelve onsite processing units and is illustrated in Figure 1. A more detailed working sketch of all interplant streams is provided in Figure 2. The major processing units are:

1. Crude Unit
2. Hydrovisbreaker Unit
3. Naphtha Hydrotreating Unit
4. Distillate Hydrotreating Unit
5. Distillate Hydrocracking Unit (with product fractionation)
6. Gas Plant

Additional auxiliary units include:

7. Hydrogen Plant (Steam-Hydrocarbon Reforming)
8. Hydrogen Purification Unit (Pressure Swing Adsorption)
9. Low Pressure Amine Unit
10. Sour Water Stripper and Ammonia Plant (Chevron WWT Process)
11. Sulfur Recovery Unit (Claus unit followed by BSR/MDEA tail gas unit)
12. Flue Gas Desulfurization Unit (Wellman-Lord/Davy Powergas Process)

The refinery process concept emerged from the Phase I case studies and was demonstrated in the Phase II bench-scale work. The processing objective was the exclusive production of wide-cut gasoline, or JP-4 type aviation turbine fuel. However, the refinery has been configured to allow the production of kerosene, or JP-8 type turbine fuel with a minimum of equipment or operational changes.

To simplify the change from one product slate to the other, the refinery has been designed so that all process operations upstream of the Distillate Hydrocracker and Main Fractionator are identical for both JP-4 and JP-8 production. Consequently, switching from one product slate to the other requires changing only one primary control parameter (the fractionation end-point of the turbine fuel product from the Main Fractionator Tower) and

Note: Dotted line show products flow during JP-8 operation.

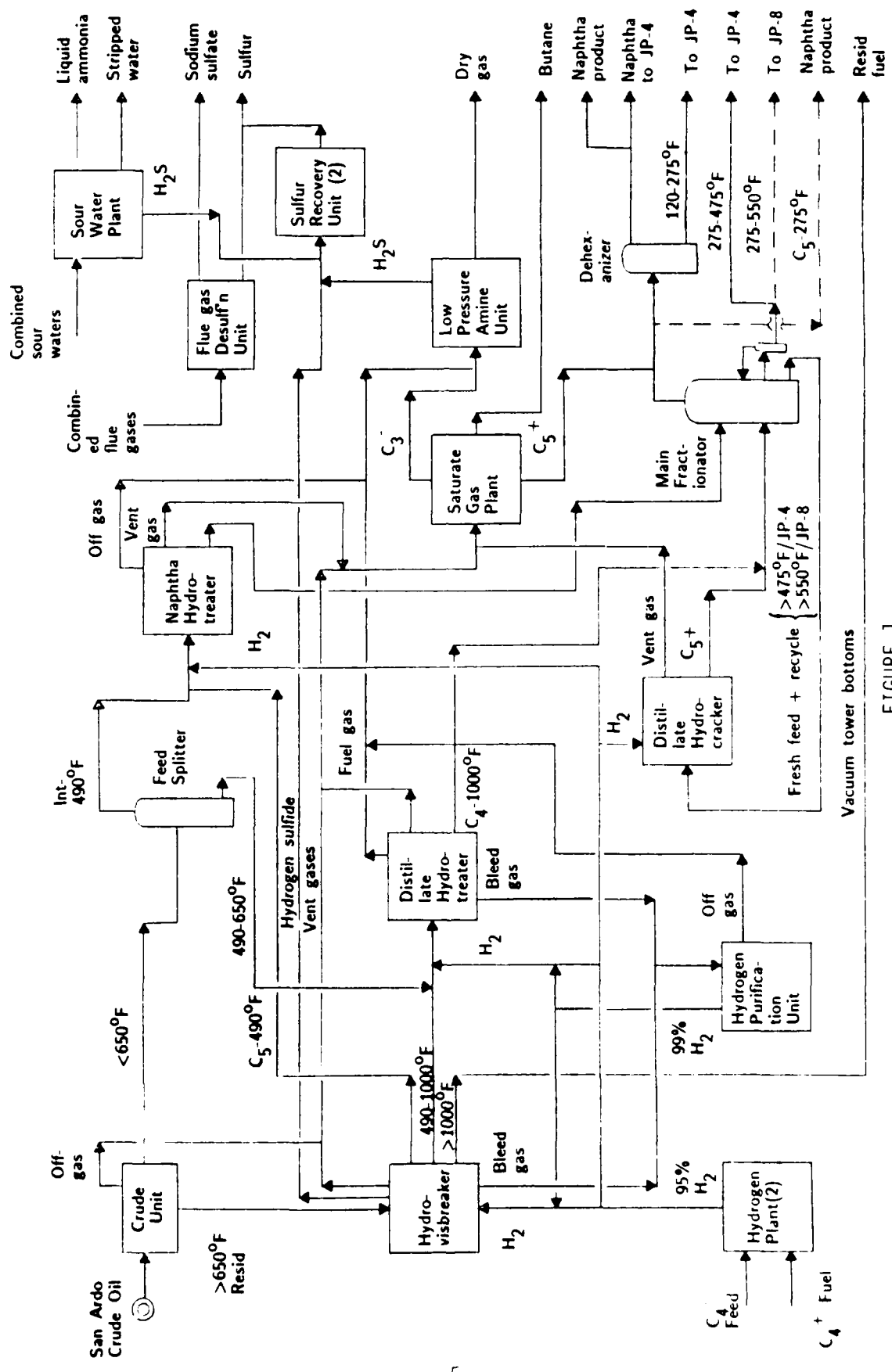


FIGURE 1  
REFINERY PROCESS FLOW SCHEME

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# HEAVY OIL UPGRADING REFINERY

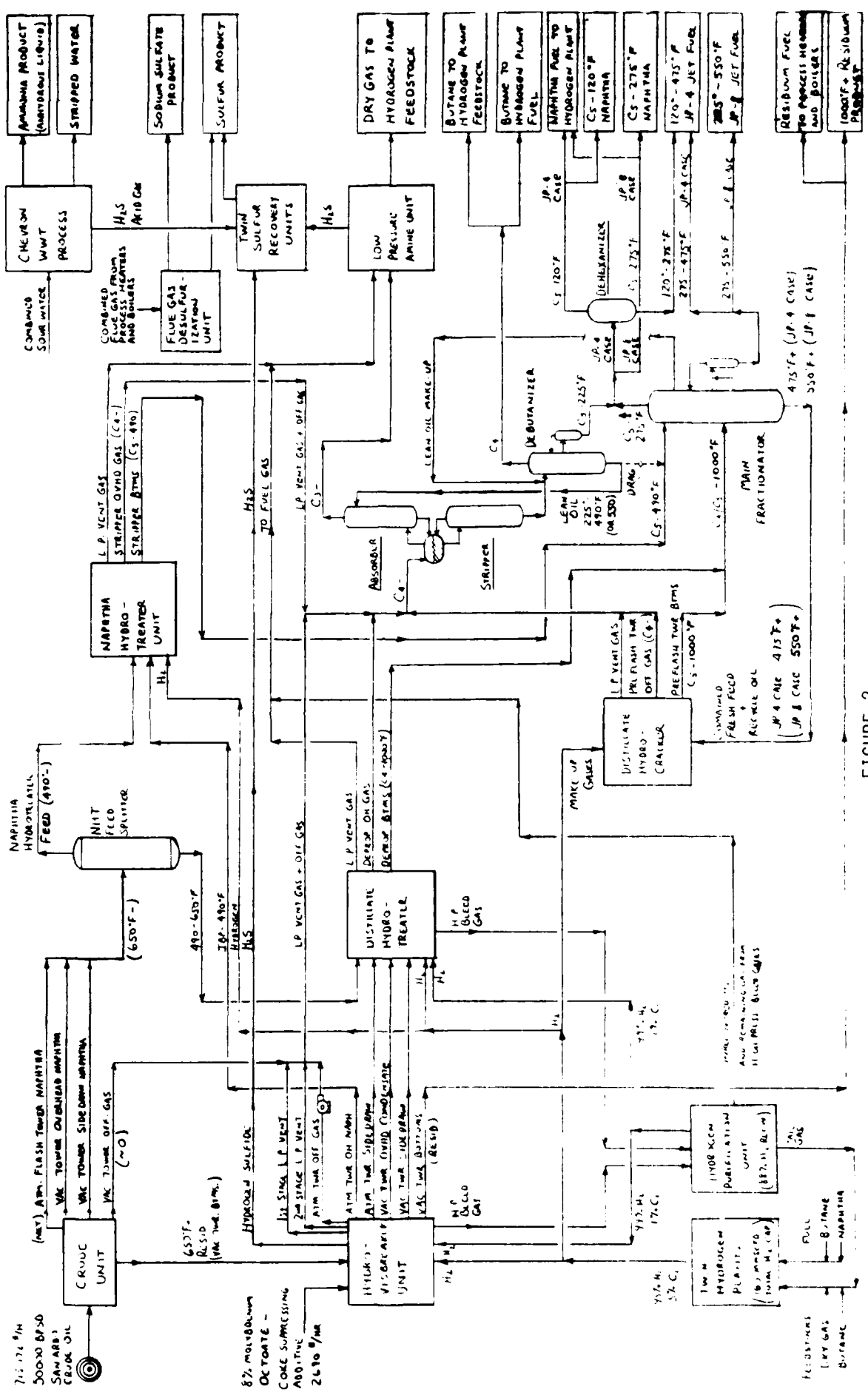


FIGURE 2  
SKETCH OF INTERPLANT STREAMS



placing the Dehexanizer Tower downstream of the Main Fractionator in or out of service. During JP-4 operations the Dehexanizer is in operation; during JP-8 production it is out of service.

Descriptions of individual refinery operating units follow.

### Crude Unit

The refining process begins with 50,000 BPSD of crude oil being charged to the Crude Unit, where straight-run distillate boiling below 650°F is separated from the >650°F vacuum reduced crude, which is to be upgraded at the Hydrovisbreaker. The incoming crude oil is first diluted with 16,000 BPSD recycled straight-run naphtha to reduce the crude viscosity to nearly 2 centipoise which is optimal for the subsequent 2-stage desalting operation. In the desalters, inorganic contaminants (sediment and water-soluble) are removed from the crude.

The desalted crude is flashed at atmospheric pressure to recover a large quantity of naphtha, most of which is recycled for crude dilution. The remainder is removed from the Crude Unit as a net straight-run naphtha stream. Sharp fractionation of the recycled naphtha is unnecessary, and the atmospheric flash is adequate. This reduces the load on the downstream vacuum distillation tower to which the atmospheric flash tower bottoms is fed.

The vacuum tower fractionates this stream at a true-boiling cut point of 650°F. The straight-run <650°F vacuum distillate is combined with the net straight-run naphtha from the atmospheric flash tower and routed to the Naphtha Hydrotreater Feed Splitter Tower. The vacuum reduced crude boiling above 650°F, representing 79 volume percent of the whole crude, becomes the liquid feed for the Hydrovisbreaker.

### Hydrovisbreaker

The hydrovisbreaker feed rate is 39,366 BPSD. Approximately 60 volume percent of this boils above 975°F. Because this material cannot be converted directly to high-quality turbine fuels by conventional hydrotreating and hydrocracking alone, it is first upgraded to lighter products in the hydrovisbreaker. In the hydrovisbreaker reactors, the residuum is thermally cracked at 850°F and 2500 psig in the presence of hydrogen and a molybdenum-

based coke-suppressing additive. Molybdenum concentration is 367 ppm (wt.) of reactor liquid feed. About 70 percent of the >975°F fraction is converted to lower boiling products.

The hydrovisbreaker reactor effluent is separated into recycle gas, naphtha, middle and heavy distillates, and a vacuum residuum by a series of phase separators and atmospheric and vacuum fractionation towers.

Since feed sulfur conversion exceeds 50 percent, a high pressure amine scrubber is included to remove hydrogen sulfide from the recycled hydrogen gas stream. The recovered hydrogen sulfide is routed directly to the refinery Sulfur Recovery Unit.

#### Feed Splitter Tower (at the Naphtha Hydrotreater Unit)

The straight-run naphtha plus distillate mixture from the crude unit is charged to the feed splitter tower at the Naphtha Hydrotreating Unit, where it is fractionated more sharply into naphtha and middle distillate fractions at a nominal 490°F cut-point.

The feed splitter overhead stream needs only to be hydrotreated to be suitable for JP-4 blend stock. It is combined with hydrovisbroken naphtha of the same boiling range and processed in the Naphtha Hydrotreating Unit.

The feed splitter bottoms is straight-run distillate boiling nominally from 490-650°F. It needs to be more severely hydrotreated and hydrocracked to make acceptable turbine fuel blend stock. This refining takes place in the distillate hydrotreater/hydrocracker complex.

#### Naphtha Hydrotreater Unit

The straight-run and hydrovisbroken naphthas are combined and catalytically hydrotreated at the Naphtha Hydrotreating Unit to reduce olefin, aromatic, sulfur, and nitrogen contents to acceptable levels. Yields are based upon pilot plant results with a nickel-molybdenum-on-alumina catalyst. The reactor design temperature and pressure are 650°F and 1250 psig, respectively.

Before further processing, the hydrotreated liquid product is stripped of light ends to permit safe intermediate storage if that becomes necessary. It is then routed to the Main Fractionator Tower at the Distillate Hydrocracker Unit, where it is separated into naphtha and turbine fuel products.

#### Distillate Hydrotreater Unit

The feed to the Distillate Hydrotreater Unit consists of the straight run gas oil from the Naphtha Hydrotreater Feed Splitter Tower bottoms, and the middle and heavy distillate cuts from the Hydrovisbreaker atmospheric and vacuum fractionation towers.

The combined distillates are catalytically hydrotreated to reduce olefin, heteroatom, and aromatics contents, producing acceptable quality hydrocracker feed. Pilot plant experiments, using the same nickel-molybdenum-on-alumina catalyst as in the Naphtha Hydrotreater Unit, formed the yield basis for this design. Operating conditions were 750°F and 2500 psig.

Distillate Hydrotreater Unit processing conditions are relatively severe, producing considerable reduction in feed molecular weight and distillation range. Therefore, hydrotreater plant product, after preliminary stripping/stabilization, is directed to the main product fractionator, rather than to the hydrocracking reactor, to separate the lighter products from the heavier gas oil feed intended for the Hydrocracker. This step reduces charge rate to the Hydrocracker Unit and avoids excessive light ends production in the Hydrocracker Unit.

#### Main Fractionator Tower & Distillate Hydrocracker Unit

The Main Fractionator Tower is located at the Distillate Hydrocracker Unit. It serves to 1) fractionate final refinery products, 2) to separate light components from the the Distillate Hydrocracker fresh feed, and 3) to separate hydrocracker products from the unconverted heavy gas oil which is recycled to extinction in the hydrocracker.

Stabilized naphtha from the Naphtha Hydrotreater Unit is supplied to the upper part of the Man Fractionator Tower becoming overhead (light naphtha) and side stream (turbine fuel) products.

The Distillate Hydrotreater liquid product, which is the source of fresh feed for the Distillate Hydrocracker, contain a significant amount of light ends. It is fed to the lower part of the Main Fractionator Tower, where the naphtha and turbine fuel components are removed from the heavy gas oil, which is routed to the Hydrocracker reactor.

The same feed tray also receives liquid product from the Hydrocracker Unit. This stream as also been stabilized, by high pressure and low pressure flashes, to unload the Main Fractionator and to permit safe intermediate storage if required.

The Main Fractionator bottoms product is the total liquid feed for the Distillate Hydrocracker. Under design conditions 60 volume percent of it is fresh feed originating from the Distillate Hydrotreater and 40 percent is unconverted oil recycled from the Hydrocracker product stream.

Operating conditions within the hydrocracking reactor are established to maintain 60 percent conversion of total reactor liquid feed per pass through the reactor. By recycling all unconverted gas oil back to the reactor the Distillate Hydrocracker achieves overall 100% conversion, catalytically cracking all of the heavy gas oil to turbine fuel and lighter products.

Turbine fuel sidestream product is removed from the Main Fractionator and is steam-stripped. Stripper tower vapor is returned to the Main Fractionator above the sidedraw, and the stripped bottoms is sent to turbine fuel storage. During JP-4 operations this turbine fuel stream is combined with the Dehexanizer bottoms to form the entire JP-4 turbine fuel product. During JP-8 production, the sidestream is drawn from a lower tray of the Main Fractionator, and after being stripped, comprises the entire JP-8 product.

A lean oil purge stream from the Gas Plant Debutanizer bottoms is fed to the upper part of the Main Fractionator for removal of light naphtha. A lean oil makeup stream is drawn from an upper tray and returned to the Debutanizer.

### Switching Product Slates

Combining these various fractionation duties in the one Main Fractionator column reduces to three the number of process changes necessary to switch from JP-4 to JP-8 turbine fuel production. All of them occur at the Main Fractionator and Distillate Hydrocracker Unit. They include:

- (1) The fractionation cut-point between the Main Fractionator sidestream turbine fuel and the bottoms product (which is Distillate Hydrocracker feed) must be increased, nominally from the 475°-490°F range up to 550°F. This is done by drawing the sidestream from a lower tower tray and adjusting the reboiler duty upward.
- (2) In a response to the altered cut-point mentioned above, the Distillate Hydrocracker liquid charge rate is lowered and the initial boiling point of the feed is increased by about 60°F. This will require an adjustment to the hydrocracker reactor temperature in order to maintain the nominal 60 percent conversion per pass.
- (3) For both JP-4 and JP-8 operations the Main Fractionator overhead, which boils nominally below 275°F, is combined with the Gas Plant Debutanizer naphtha sidestream. During JP-4 operations this combined stream is separated into light naphtha, boiling below about 120°F, and a 120°-275°F turbine fuel component in the Dehexanizer Tower. The turbine fuel portion is then combined with the stripped sidestream from the Main Fractionator to form the entire JP-4 turbine fuel product.

During JP-8 production, the Main Fractionator overhead and Debutanizer sidestream are combined as naphtha product because they are not within the 275°-550°F boiling range of the JP-8 product. Consequently, the Dehexanizer tower may be taken out of service and the entire JP-8 product comes from from the Main Fractionator sidedraw.

## Gas Plant

The Gas Plant is comprised of a combination Absorber-Stripper Tower, a Debutanizer Tower and a small Debutanizer Sidestripper Tower. The plant processes the combined gaseous products and light condensibles which originate at certain low pressure separator vents, plus all of the tower overhead gases.

The Gas Plant produces a dry gas stream containing hydrogen and C<sub>1</sub>-C<sub>3</sub> hydrocarbons, a butane product stream, and a naphtha stream.

Gas Plant feed enters the top of the Stripper section of the Absorber-Stripper tower. Gas travels upward through the Absorber tower, while lean oil introduced at the top of the Absorber recovers condensibles from the gas. The Stripper removes dissolved gases from the rich oil, which leaves as Stripper bottoms enroute to the Debutanizer.

The rich oil is combined with makeup lean oil from the Main Fractionator, to form the total Debutanizer feed. The Debutanizer produces a butane product overhead plus a stabilized naphtha sidestream. Debutanizer bottoms is recovered as absorber lean oil. A Debutanizer Sidestripper tower strips light ends from the naphtha before it leaves the unit.

A small lean oil purge stream is drawn from the Debutanizer bottoms and routed back to the Main Fractionator Tower.

The Absorber dry gas is sent to the Low Pressure Amine unit with other light refinery gases for removal of hydrogen sulfide before it is used as hydrogen plant feed. The butane product is consumed partly as feedstock for the hydrogen plant and partly as fuel for the hydrogen plant furnaces. The naphtha from the gas plant is mixed with the overhead distillate from the Main Fractionator upstream of the Dehexanizer tower.

### Low Pressure Amine Unit

The dry gas from the Gas Plant is combined with low pressure vent gas from both hydrotreaters and the tail gas from the Hydrogen Purification Unit to form the feed to the Low Pressure Amine Unit. Hydrogen sulfide in these gases must be removed before the dry gas can be used as feedstock to the hydrogen plant.

The hydrogen sulfide is absorbed by a lean alkanolamine solution in an amine contactor tower. The  $H_2S$ -rich amine leaving the bottom of the contactor enters the top of an amine still, where the hydrogen sulfide is stripped overhead from the circulating amine and is directed to the Sulfur Recovery Unit. The lean amine is recycled to the top of the Absorber.

The  $H_2S$ -free dry gas from the Low Pressure Amine Unit is totally consumed as feedstock for the Hydrogen Plant (Steam-Hydrocarbon Reformer).

### Hydrogen Plant

Two parallel hydrogen plants convert light hydrocarbon gases to hydrogen by a steam reforming process. Two are required, since the total refinery hydrogen requirements are nearly twice the capacity of the largest packaged hydrogen plant normally constructed.

The primary feedstock for the Hydrogen Plant is the dry gas product from the Low Pressure Amine Unit. Butane product from the Gas Plant is supplemental feed. The remaining butane fuels the hydrogen plant reformer furnaces, which require a relatively clean, sulfur-free fuel. Since the reformer furnace duties exceed the available butane fuel, some naphtha product is also fired.

Hydrogen Plant feedstock is further desulfurized and combined with water (plant condensate). Tubes of catalyst in the reformer furnace convert the steam and hydrocarbons to hydrogen and to carbon oxides. Any carbon monoxide is converted to carbon dioxide in a high-temperature shift reactor at 750°F followed by a low-temperature shift reactor at 400°F.

Potassium carbonate solution removes carbon dioxide from the gas products at an absorber, then releases the  $\text{CO}_2$  to an atmospheric vent at a stripper tower while being regenerated for recycle to the absorber. A methanation reactor converts residual amounts of carbon monoxide in the gas to methane. The product is a makeup gas containing 95 percent hydrogen and 5 percent methane, which is then supplied to the hydrovisbreaking unit, both hydrotreaters, and the hydrocracker unit.

#### Hydrogen Purification Unit

A 95 percent purity for makeup hydrogen is adequate for the Naphtha Hydrotreater and Distillate Hydrocracker plants. Neither unit requires a high pressure purge gas be drawn from the recycle hydrogen stream to maintain high purity, because potential impurities are satisfactorily carried away in the liquid products.

However, if the Hydrogen plant were the only source of fresh hydrogen for the Hydrovisbreaker and Distillate Hydrotreater units, very large high-pressure purge rates, and correspondingly high makeup hydrogen rates, would be necessary to meet the hydrogen partial pressure requirements. The operations would approach expensive "once-through" hydrogen flow with little gas recycled.

To avoid this, a Hydrogen Purification Unit recovers, purifies, and recycles the hydrogen present in the high-pressure purge gases from the Hydrovisbreaker and Distillate Hydrotreater. The Pressure Swing Adsorption (PSA) process, which is a cyclic operation, is used. Its five distinct steps require installation of five identical vessels, which are filled with molecular sieve separation media.

Feed entering at high pressure is expanded through a turboexpander to recovery energy, and is charged to one of the five vessels. During the on-stream cycle, the hydrocarbons in the feed are selectively adsorbed on the molecular sieve. Hydrogen containing a small amount of hydrocarbon, principally methane, passes through as purified product. Meanwhile the other



four vessels are undergoing varying degrees of depressurization, desorption of hydrocarbons, and purging to a tail gas stream. Periodically, each of the vessels switches to the next process step in the series, and feed is routed to a different vessel.

Hydrogen recovery from the PSA feed gas is nearly 90 percent, and the resulting purity is 99 percent, with 1 percent methane. The purified hydrogen is recompressed and directed to the Hydrovisbreaker and Distillate Hydrotreater reactors. The purged tail gas is processed at the Low Pressure Amine Unit, eventually to be consumed as feedstock for the Hydrogen Plant.

#### Sour Water Stripper and Ammonia Plant

Sour process water is generated at several operating units and requires treatment before reuse or disposal. Inorganic contaminants are ammonia, hydrogen sulfide, or chemical combinations of the two in the form of ammonium sulfides, pentasulfides or hydrosulfide. Traditional two-stage water stripping would consume both caustic soda and acid to recover ammonia and hydrogen sulfide separately.

However, the proprietary Chevron WWT process for sour water clean-up does not require caustic and acid, and permits reuse of the stripped water for hydroprocessing injection, crude unit desalting water, and other process uses. The reduced refinery water usage and disposal is important in the arid environment in which the refinery would be constructed.

The plant first strips incoming sour water of residual dissolved gases, then sequentially steam strips the water of hydrogen sulfide, then ammonia. Plant products are the cleaned-up water available for re-use, anhydrous liquid ammonia for sale, and hydrogen sulfide, which is supplied to the sulfur recovery unit.

#### Sulfur Recovery Unit

Accepted environmental practice requires that hydrogen sulfide be removed from vented gases throughout the refinery. Typically, the hydrogen sulfide is converted at a Sulfur Recovery Unit to elemental sulfur, which is sold.

In this refinery design, hydrogen sulfide is recovered from three sources: 1) the High Pressure Amine Unit that processes the Hydrovisbreaker recycle gas, 2) the Low Pressure Amine Unit which cleans the refinery dry gas and vent gas, and 3) the Chevron WWT Unit (Sour Water Stripper and Ammonia Plant) which recovers hydrogen sulfide from refinery sour water.

The Sulfur Recovery Unit consists a Claus unit and a tail gas clean-up unit. The Claus unit partially reacts hydrogen sulfide and oxygen (air) to produce molten elemental sulfur. Because the resulting combustion gases include low concentrations of sulfur oxides, they cannot be vented directly to the atmosphere. The tail gas unit reduces the sulfur content to acceptable levels by catalytic reduction to hydrogen sulfide, followed by extraction of the hydrogen sulfide with alkanolamine. The hydrogen sulfide is returned to the front end of the Claus unit, and the scrubbed gases are vented. The primary plant product is saleable molten sulfur.

#### Flue Gas Desulfurization Unit

With the exception of the Hydrogen Plant Reformer Furnace, all refinery process heaters plus the main refinery boiler are fueled by vacuum residuum produced at the Hydrovisbreaker Vacuum Distillation Tower. Flue gas from these furnaces contains sulfur dioxide which cannot be vented to the atmosphere. In addition, the resid fuel contains appreciable trace metals (nickel, vanadium, and molybdenum from the Hydrovisbreaker coke-suppressing additive), which become fly ash during fuel combustion.

Therefore all refinery flue gas (except from the Hydrogen Plant reformer furnace) is processed in the Flue Gas Desulfurization Unit. Flue gas first passes through an electrostatic precipitator to capture fly ash, after which it is desulfurized.

The regenerative Wellman Lord/Davy Powergas process was selected to desulfurize the flue gas. Because it is regenerative, it produces a molten sulfur product and thus avoids a significant disposal problem associated with spent treating agent, which non-regenerative processes have (e.g., the limestone slurry throw-away processes).

## Furnace Fuels

The vacuum residuum from the Hydrovisbreaker Vacuum Distillation Tower becomes fuel for the refinery main boiler and all process heaters with the exception of the Hydrogen Plant reformer furnace. While its sulfur content of 1.5 weight percent is tolerable in the process heaters, the residual fuel is not suitable for the high temperature Hydrogen plant reformer furnace, where it would tend to hasten corrosion of the furnace tubing and risk catastrophic hydrogen leaks.

The primary clean fuel used in the Hydrogen Plant reformer furnace is that portion of the butane product from the Gas Plant which has not been used as feedstock for the hydrogen plant. The butane fuel is supplemented by naphtha product to meet the firing requirements.

Commercial operations using vacuum residuum as furnace fuel indicate that there should be no problem in firing undiluted Hydrovisbreaker residuum, provided it can be stored and circulated at temperatures to 450°F. However, it would be necessary to establish the appropriate burner design to obtain adequate atomization. Separate burners for firing a lighter oil to control the heater duty more accurately may be helpful.

Because the residuum is an undesirable net product, its use as furnace fuel helps to solve the problem of its sale or disposal.

## IV. IMPACT OF SWITCHING TO JP-8 PRODUCTION

Rigorous design calculations were completed only for the case of JP-4 turbine fuel production. Separate calculations for the JP-8 case would have been redundant. However, JP-8 production does impact operations and economics of the refinery, specifically at the Main Fractionator, the Distillate Hydrocracker, and the Dehexanizer tower.

Raising the cut-point between the Main Fractionator turbine fuel sidedraw and the tower bottoms from 475°F to 550°F will produce little, if any, change in the total combined production of naphtha and turbine fuel distillates at the tower. The net effect is that slightly more of the fresh gas oil feed from the Distillate Hydrotreater will pass directly through the Main Fractionator to the liquid product pools without first having to be hydrocracked.

Obviously, the change in initial boiling point of the turbine fuel product, from about 120°F for JP-4 to about 275°F for JP-8, means that more of the Main Fractionator overhead distillate is going to exit the refinery as naphtha, and less as turbine fuel. However, the economic bases employed in this project assigned equal market values to the naphtha and turbine fuel. Consequently, although the ratio of naphtha to turbine fuel is greater for the JP-8 case than the JP-4 case, it does not change the economics; the combined sales revenue from these two products remains the same. Of course, if separate values were assigned to naphtha, JP-4, and JP-8, an economic impact could be calculated.

Switching from JP-4 to JP-8 production will reduce refinery operating costs modestly. Shutting down the Dehexanizer tower obviously eliminates the consumption of reboiler steam, condenser cooling water, and electrical demand for pumps, etc. The Distillate Hydrocracker will operate in a less severe mode. With the slower charge rate and increased reactor residence time, a lower reaction temperature can be used to achieve the same 60% conversion per pass, and less charge preheating will be needed. However, this may be somewhat off-set by a potentially greater reboiler temperature at the Main Fractionator to achieve the higher cut-point.

For equipment sizing, JP-4 production is the size-determining Hydrocracker operation; less throughput is required for the JP-8 mode of operation. Total fresh liquid feed to the Main Fractionator from the two hydrotreaters remains the same for both cases. Thus, switching to JP-8 would not bottleneck the refinery in any way. JP-8 operation did impact the size of the naphtha storage tank. It is sized for the larger naphtha flow rate that results from the shutdown of the Dehexanizer tower throughout JP-8 operations.

Concern was raised that the processing change from JP-4 to JP-8 could alter the quality, primarily sulfur content, of the naphtha being fired in the Hydrogen Plant reformer furnace. Review of the respective stream flows and qualities indicated the naphtha sulfur content would remain at extremely low levels.

Overall, the impact of switching to JP-8 production is negligible.

Table 1

TURBINE FUEL REFINERY CAPITAL COST SUMMARY

ONSITES	<u>Installed Cost</u> <sup>1</sup>
Crude Unit .....	\$ 19,968,000
Hydrovisbreaker Unit .....	168,119,000
Naphtha Hydrotreater Unit .....	24,154,000
Distillate Hydrotreater Unit .....	140,183,000
Distillate Hydrocracker Unit .....	94,724,000
Gas Plant .....	9,380,000
Hydrogen Plant .....	99,075,000
Hydrogen Purification Unit .....	60,915,000
Low Pressure Amine Unit .....	3,079,000
Sour Water Stripper and Ammonia Plant .....	33,091,000
Sulfur Recovery Unit .....	37,119,000
Flue Gas Desulfurization Unit .....	<u>54,328,000</u>
 Total Onsites .....	 \$ 744,135,000
<hr/>	
OFFSITES	
Tankage .....	\$ 45,061,000
Other: Specified by U.S. Air Force as 45% of onsite costs .....	334,861,000
SPARE PARTS .....	1,498,000
ROUND UP TO NEAREST MILLION DOLLARS .....	445,000
<hr/>	
TOTAL REFINERY INSTALLED COST .....	\$ 1,126,000,000

<sup>1</sup> Based on 4th Quarter 1985 prices, Salt Lake City, Utah location

Table 2

SCOPE OF ONSITE AND OFFSITE FACILITIES

ONSITES

Crude Unit  
Hydrovisbreaker Unit  
Naphtha Hydrotreater Unit  
Distillate Hydrotreater Unit  
Distillate Hydrocracker Unit  
Gas Plant  
Hydrogen Plant  
Hydrogen Purification Unit  
Low Pressure Amine Unit  
Sour Water Stripper and Ammonia Plant  
Sulfur Recovery Unit  
Flue Gas Desulfurization Unit

OFFSITES

TANKAGE:

Crude tankage  
Intermediate tankage  
Product tankage  
Fuel tankage

OTHERS:

Site preparation, grading, dyking and piling  
Paved roads and railroad spur  
Office building, cafeteria, change rooms  
Maintenance buildings, warehouse, and spare parts  
Medical facilities  
Powerhouse  
Electrical power substation and grid  
Boiler feed water treating and water storage  
Steam boilers, distribution piping, and condensate tank  
Air systems - plant and instrument  
Firehouse and trucks  
Fire water pond, fire water pumps, and distribution system  
Foam fire system on tanks  
Cooling towers and water supply  
Sanitary drinking water  
Hydrogen and gas flare system  
Butane treating  
Product loading for sulfur, ammonia  
Crude and product receiving and pumping station  
Communication systems  
Offsite piping and pipeways  
Nitrogen purge system  
Sewer systems (3): contaminated, sanitary, run-off  
Waste water treatment plant, API separators, sloop tanks  
Duct work from fired heaters  
Receiving truck rack and chemical storage  
Sludge disposal storage  
Spent catalyst disposal site

Table 3

SUMMARY OF MAJOR REFINERY STREAM RATES

The stream rates below were calculated by computer simulations of individual operating plants and were used to specify major equipment sizes. Generally, the individual plant weight balances achieved 100 percent recoveries. However, the simulations dealt principally with the distribution of hydrocarbons and hydrogen gas, not the precise flow of sulfur and nitrogen throughout the refinery. Sulfur and nitrogen flows were determined separately and were superimposed on the overall refinery design. The resulting overall refinery material balance slightly exceeded 100 percent weight recovery, and therefore some reduction of refinery hydrocarbon product rates was necessary to establish 100 percent closure and determine product yields. Consequently, the data below, used for equipment sizing, will differ somewhat from the final product yields, intended for refinery economic evaluation, but in either case the values are conservative for their respective purposes.

	<u>BPSD</u>	<u>MSCFH</u>	<u>M LB/HR</u>
<u>CRUDE UNIT</u>			
Crude Oil Charge	50,000	-	715,195
Naphtha Recycled as Crude Oil Diluent	16,008	-	200,947
<u>HYDROVISBREAKER UNIT</u>			
Liquid Charge	39,366	-	577,565
Additive Feed	-	-	2,690
Naphtha Recycled as Additive Diluent	600	-	6,724
Fresh Hydrogen Feeds			
from Hydrogen Plant	-	1,670	11,960
from Hydrogen Purification Unit	-	1,546	8,785
Recycled Gas	-	6,355	90,381
Total Reactor Gas Feed	-	9,571	111,126
<u>NAPHTHA HYDROTREATER UNIT</u>			
Feed Splitter Charge	10,542	-	136,443
Naphtha Hydrotreater Charge:			
Feed Splitter Overhead Naphtha	4,267	-	53,271
Hydrovisbroken Naphtha	11,875	-	135,003
Total Liquid Charge	16,142	-	188,274



Table 3

SUMMARY OF MAJOR REFINERY STREAM RATES

(continued)

	<u>BPSD</u>	<u>MSCFH</u>	<u>M LB/HR</u>
<u>NAPHTHA HYDROTREATER UNIT (continued)</u>			
Makeup Hydrogen	-	374	2,677
Recycle Gas	-	2,885	34,680
Total Reactor Gas Feed	-	3,259	37,357
<u>DISTILLATE HYDROTREATER UNIT</u>			
Charge:			
Feed Splitter Bottoms	6,276	-	83,172
Hydrovisbroken Distillates	24,376	-	337,644
Total Liquid Charge	30,652	-	420,816
Fresh Hydrogen Feeds			
from Hydrogen Plant	-	2,290	16,398
from Hydrogen Purification Unit	-	1,663	9,450
Recycled Gas	-	2,430	21,286
Total Reactor Gas Feed	-	6,383	47,134
<u>GAS PLANT</u>			
Total Gas Feed (to Absorber-Stripper)	-	1,413	111,806
Lean Oil Makeup (to Debut. Feed)	975	-	10,745
Absorber Overhead Dry Gas	-	944	37,062
Debutanizer Overhead Butane	7,165	-	59,512
Debutanizer Sidedraw Naphtha	1,798	-	16,585
Lean Oil Purge (from Debut. Bottoms)	843	-	9,396
<u>DISTILLATE HYDROCRACKER UNIT</u>			
Total Liquid Charge			
(Main Fractionator Bottoms)	44,184	-	563,618
Makeup Hydrogen	-	2,416	17,301
Recycled Gas	-	3,107	41,312
Total Reactor Gas Feed	-	5,524	58,613

Table 3

SUMMARY OF MAJOR REFINERY STREAM RATES

(continued)

	<u>BPSD</u>	<u>MSCFH</u>	<u>M LB/HR</u>
<u>DISTILLATE HYDROCRACKER UNIT (continued)</u>			
MAIN FRACTIONATOR			
Feeds:			
Naph. Hydrotreater Liquid Prod.	15,770	-	187,498
Dist. Hydrotreater Liquid Prod.	33,354	-	422,847
Dist. Hydrocracker Liquid Prod.	41,983	-	487,938
Lean Oil Purge from Gas Plant	843	-	9,396
Lean Oil Makeup to Gas Plant	975	-	10,745
Sidestripped Turbine Fuel	26,052	-	301,799
DEHEXANIZER TOWER			
Feeds:			
Main Fractionator Overhead	23,180	-	246,028
Gas Plant Naphtha Product	1,975	-	18,155
Overhead Naphtha Product	5,505	-	50,986
Bottoms Turbine Fuel Product	19,650	-	213,197
<u>HYDROGEN PLANT</u>			
Total Dry Gas Feedstock	-	2,015	69,889
Butane Feedstock	1,425	-	11,833
Reaction Water Feed	422	-	147,728
Makeup Gas Product	-	6,752	48,336
Carbon Dioxide Product	-	1,562	181,114
<u>HYDROGEN PURIFICATION UNIT</u>			
Feedstocks:			
Hydrovisbreaker High Pres. Bleed Gas	-	2,035	28,951
Dist. Hydrotreater High Pres. Bleed	-	1,982	17,358
Purified Hydrogen Product	-	3,209	18,235
Tail Gas (used as Hydrogen Plant feed)	-	808	28,074

M = thousands, e.g. 1 MSCFH = 1 thousand standard cubic feet per hour

Table 4

REFINERY PRODUCT YIELDS

Basis: JP-4 Turbine Fuel Production

<u>Net Salable Products</u>	<u>BPSD</u>	<u>Vol% of Crude Oil Charge</u>	<u>lb/hour</u>	<u>Wt% of Crude Oil charge</u>
JP-4 Turbine Fuel	44,298	88.60	499,180	69.80
Naphtha	3,290	6.58	30,468	4.26
Residuum	681	1.36	10,770	1.50
Ammonia (101.4 Tons/Day)	-	-	8,367	1.17
Sulfur (172.8 Tons/Day)	-	-	14,398	2.01
Sodium Sulfate	-	-	272	0.04
<u>Net Unsold Product</u>				
Carbon Dioxide (1.514 MMSCFH)	-	-	175,552	24.55
<u>Totals</u>			739,007	103.33

Table 5

REFINERY WEIGHT BALANCE

Basis: JP-4 Turbine Fuel Production

Boundary Limits:

Boundary limits for this balance include the oil and gas processing operations at all the major processing units and auxiliary units with these exceptions: combustion of refinery products used as fuels for all fired heaters, plus the desulfurization of resulting flue gases are considered outside the boundary.

Refinery Inputs:

	<u>lb/hr</u>
Crude Oil .....	715,195
Coke-suppressing Additive .....	2,690
Natural Gas to Sulfur Recovery Unit .....	1,012
Reaction water for Hydrogen Plant steam reformer .....	147,728
Total Input .....	866,625

Refinery Outputs:

To Fuels:	Butane .....	46,215
	Naphtha .....	18,952
	Residum .....	63,365
To Sales:	Naphtha .....	30,468
	JP-4 Turbine Fuel .....	499,180
	Residuum .....	10,770
	Sulfur (from Sulfur Recovery Unit) .....	13,757
	Ammonia .....	8,367
Vented:	Carbon Dioxide from Hydrogen Plant .....	175,552
Total Output .....		866,625

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Streams outside Boundary Limits defined above:

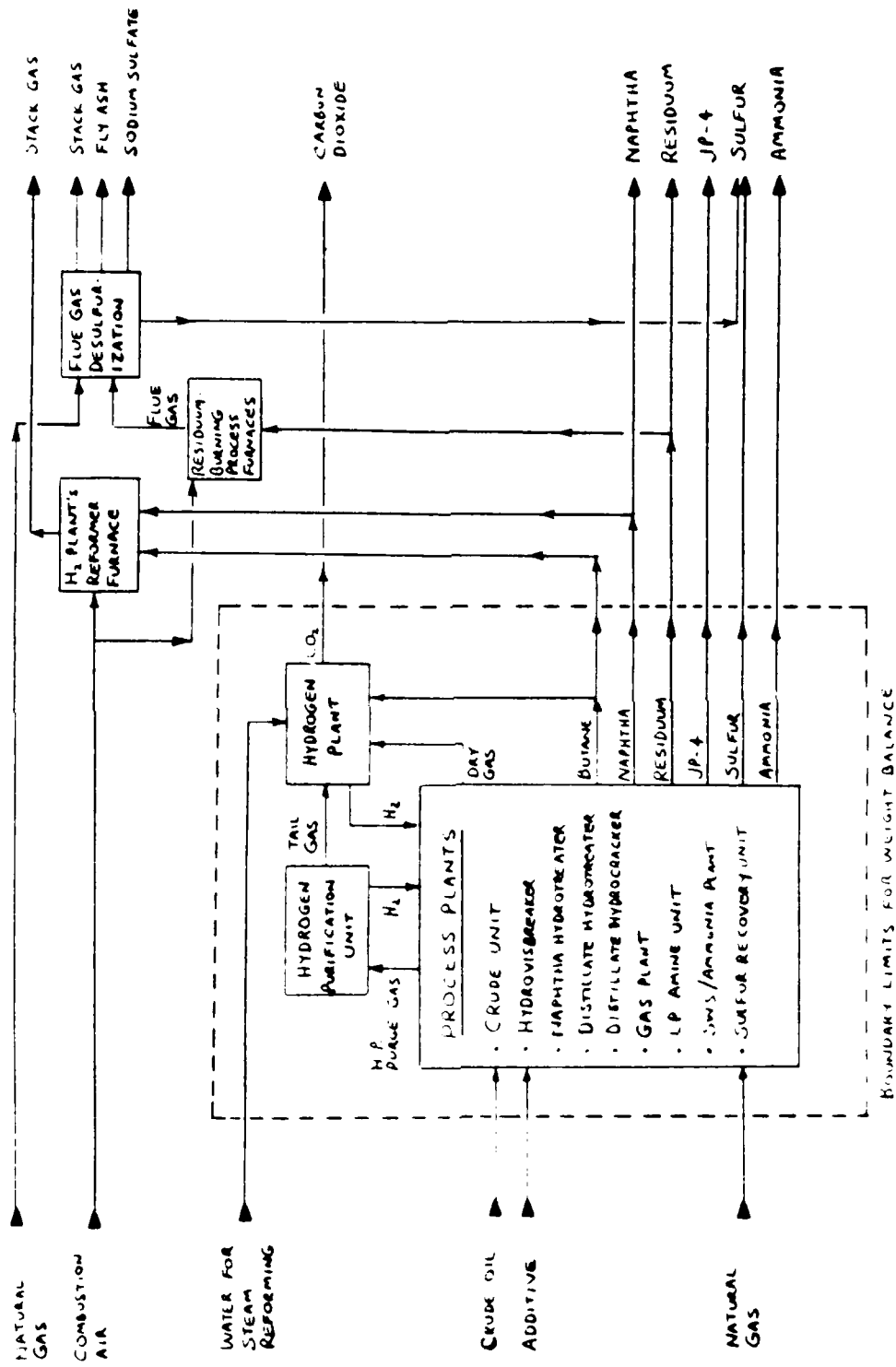
Additional Refinery Input:

Natural Gas to the Flue Gas Desulfurization Unit .....	823
--	-----

Additional Refinery Byproducts:

From the Flue Gas Desulfurization Unit

Sulfur .....	641
Sodium Sulfate .....	272



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Figure 3 - Refinery Weight Balance Boundary Limits

Table 6

REFINERY HYDROGEN BALANCE

Basis: JP-4 Turbine Fuel Production

<u>HYDROGEN USERS</u>	<u>MAKE-UP HYDROGEN</u> (95% hydrogen) from <u>Hydrogen Plant</u>		<u>PURIFIED HYDROGEN</u> (99% Hydrogen) from <u>H2 Purification Unit</u>	
	<u>lb/hour</u>	<u>MMSCFD</u>	<u>lb/hour</u>	<u>MMSCFD</u>
Hydrovisbreaker	11,960	40.08	8,785	37.11
Naphtha Hydrotreater	2,677	8.97	-	-
Distillate Hydrotreater	16,398	54.96	9,450	39.92
Distillate Hydrocracker	17,301	57.99	-	-
	<hr/>	<hr/>	<hr/>	<hr/>
Totals	48,336	162.00	18,235	77.03
 <u>HYDROGEN SUPPLIERS</u>				
<u>Hydrogen Plant</u> (Steam Reforming)				
From Dry Gas	42,085	141.1	-	-
From Butane	6,251	21.0	-	-
 <u>Hydrogen Purification</u> (Pressure Swing Adsorption)				
From Recovered High-Pressure Purge Gases from the Hydro- Visbreaker and Distillate Hydrotreater Units	-	-	18,235	77.03
	<hr/>	<hr/>	<hr/>	<hr/>
Totals	48,336	162.0	18,235	77.03

Table 7

REFINERY FUEL BALANCE

Basis: JP-4 Turbine Fuel Production

<u>RESIDUUM FUEL REQUIREMENTS</u>	<u>Fired Duty MMBTU/HR</u>	<u>Residuum Fuel FOEB/DAY</u>
MAIN REFINERY BOILER	244.0	976
PROCESS HEATERS		
<u>Crude Unit</u>		
H-1 Flash Tower Feed Heater	115.4	462
H-2 Vacuum Tower Feed Heater	53.7	215
<u>Hydrovisbreaker</u>		
H-1 Recycle Gas Heater	110.1	440
H-2 Atmospheric Tower Feed Heater	101.1	404
H-3 Vacuum Tower Feed Heater	60.7	243
<u>Naphtha Hydrotreater</u>		
H-1 Feed Splitter Reboiler	40.23	161
H-2 Recycle Gas Heater	28.34	113
H-3 Stripper Reboiler	14.23	57
<u>Distillate Hydrotreater</u>		
H-1 Recycle Gas Heater	48.55	194
H-2 Feed Heater	19.35	77
<u>Distillate Hydrocracker</u>		
H-1 Recycle Gas Heater	15.44	62
H-2 Main Fractionator Reboiler	182.87	732
TOTAL RESIDUUM FUEL REQUIREMENT	1034.0	4136
TOTAL RESIDUUM AVAILABLE		4839
REMAINING RESIDUUM NET PRODUCT		703
<u>BUTANE &amp; NAPHTHA FUEL REQUIREMENTS</u>	<u>Fired Duty MMBTU/HR</u>	<u>Fuel Consumed BBL/DAY @ 60°F</u>
HYDROGEN PLANT REFORMER FURNACE		
Butane Product used as Fuel	917.5	5,743
Naphtha Product used as Fuel	371.5	2,111
Total Fired Duty	1289.0	

Table 8

REFINERY THERMAL EFFICIENCY

Basis: JP-4 Turbine Fuel Production

DEFINITION

$$\text{Thermal Efficiency} = \frac{\text{Heat of Combustion of Hydrocarbon Products} + \text{Energy Outputs}}{\text{Heat of Combustion of Hydrocarbon Feeds} + \text{Energy Inputs}} \times 100\%$$

		<u>THERMAL VALUE BTU/HR</u>
FEEDS:	Crude Oil	13,352,000,000
	Natural Gas	38,800,000
ENERGY INPUT:	Electrical Power	181,890,000
TOTAL INPUT:		<u>13,572,690,000</u>
PRODUCTS:	Naphtha to Sales	765,930,000
	JP-4 Turbine Fuel to Sales	10,238,770,000
	Residuum to Sales	190,420,000
TOTAL OUTPUT:		<u>11,195,120,000</u>

REFINERY THERMAL EFFICIENCY = 100 % × ( 11,195,120,000 / 13,572,690,000 )

= 82.5 %



Table 9

REFINERY ELECTRICAL REQUIREMENTS

Basis: JP-4 Turbine Fuel Production

	Brake Horsepower <u>Operating</u>	Brake Horsepower <u>Installed</u>
Crude Unit	2,000	3,500
Hydrovisbreaker Unit	5,716	10,132
Naphtha Hydrotreater Unit	2,262	3,525
Distillate Hydrotreater Unit	3,522	6,021
Distillate Hydrocracker Unit	6,745	12,290
Gas Plant	395	790
Hydrogen Plant	29,377	35,448
Hydrogen Purification Unit	4,000	4,000
Low Pressure Amine Unit	31	62
Sour Water Stripper/Ammonia Plant	1,314	2,628
Sulfur Recovery Unit	257	514
Flue Gas Desulfurization Unit	2,950	2,950
Boiler House and Water Treating	1,200	2,400
Cooling Water System	7,200	14,400
Crude Oil and Product Transfer	1,600	3,200
Plant Air Compression	2,400	2,400
Waste Water Treating & Misc.	500	1,000
 Totals      BHP	 71,469	 105,260
 Equivalent Kilowatts	 53,294	 78,492
MMBTU/HR	181.89	267.89

Table 10

REFINERY STEAM BALANCE

Basis: JP-4 Turbine Fuel Production

## STEAM CONSUMED, lb/hr

	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>
Crude Unit	20,000	11,000	10,000	41,000
Hydrovisbreaker	19,200	31,460	41,200	91,860
Naphtha Hydrotreater	1,000	5,000	5,000	11,000
Distillate Hydrotreater	25,080	5,000	10,000	40,080
Distillate Hydrocracker	91,500	34,500	-	126,000
Gas Plant	33,294	39,600	14,376	87,270
Hydrogen Plant	-	-	-	-
Hydrogen Purification	-	19,200	-	19,200
Low Pressure Amine Unit	1,310	-	12,500	13,800
Sour Water Strip/Ammonia Plant	-	90,000	118,000	208,000
Sulfur Recovery Unit	-	-	-	-
Flue Gas Desulfurization	-	45,007	-	45,007
Boiler House and Water Treating	-	-	116,131	116,131
Totals	191,384	280,767	327,207	799,358

## STEAM PRODUCED, lb/hr

	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>
Crude Unit	-	36,000	-	36,000
Hydrovisbreaker	32,450	126,025	90,625	249,100
Naphtha Hydrotreater	-	18,900	10,000	28,900
Distillate Hydrotreater	44,970	10,600	46,700	102,270
Distillate Hydrocracker	51,000	43,400	-	94,400
Gas Plant	-	-	-	-
Hydrogen Plant	-	-	-	-
Hydrogen Purification	-	-	-	-
Low Pressure Amine Unit	-	-	-	-
Sour Water Strip/Ammonia Plant	-	-	-	-
Sulfur Recovery Unit	9,000	-	28,900	37,900
Flue Gas Desulfurization	-	-	-	-
Boiler House and Water Treating	53,964	45,842	150,982	250,788
Totals	191,384	280,767	327,207	799,358

Table 11

REFINERY BOILER FEED WATER BALANCE

Basis: JP-4 Turbine Fuel Production

Steam Condensate Recycled as Boiler Feed Water

<u>Condensate Recycled from:</u>	<u>Condensate Recycled lb/hr</u>
Crude Unit	-
Hydrovisbreaker Unit	31,510
Naphtha Hydrotreater Unit	-
Distillate Hydrotreater Unit	-
Distillate Hydrocracker Unit	94,500
Gas Plant	77,270
Hydrogen Plant	-
Hydrogen Purification Unit	19,200
Low Pressure Amine Unit	13,810
Sour Water Stripper and Ammonia Plant	182,000
Sulfur Recovery Unit	-
Flue Gas Desulfurization Unit	45,007
Boiler House and Water Treating	<u>116,131</u>
Total	579,428

---

Boiler feed water to produced steam	799,358 lb/hr
Boiler feed water to blowdown loss	79,939 lb/hr
Total boiler feedwater supply required	879,297 lb/hr
Condensate recycled as boiler feed water	579,428 lb/hr
Makeup boiler feedwater required	299,869 lb/hr

Table 12  
REFINERY COOLING WATER REQUIREMENTS

Basis: JP-4 Turbine Fuel Production

	Cooling Water <u>gal/min</u>
Crude Unit	3,000
Hydrovisbreaker Unit	11,228
Naphtha Hydrotreater Unit (air-cooled)	-
Distillate Hydrotreater Unit	2,973
Distillate Hydrocracker Unit	12,650
Gas Plant	7,460
Hydrogen Plant	49,232
Hydrogen Purification Unit	3,821
Low Pressure Amine Unit	480
Sour Water Stripper and Ammonia Plant	1,720
Sulfur Recovery Unit	3,390
Flue Gas Desulfurization Unit	2,403
 Total	 98,357

Table 13

REFINERY OPERATING REQUIREMENTS

CATALYST AND CHEMICALS

COKE-SUPPRESSING ADDITIVE FOR HYDROVISBREAKER

Trade name:	8% Molybdenum Octoate
Chemical name:	Molybdenum 2-ethyl hexoate
Chemical family:	Metal carboxylate
Molecular formula:	$\text{Mo}(\text{C}_8\text{H}_{15}\text{O}_2)_2$
Normal physical state:	Liquid

Design dosage:

Based on pilot plant experiments which used an additive concentration of 1.63 wt% of Hydrovisbreaker fresh feed (i.e. Crude Unit vacuum residuum). This is equivalent to 367 wt. ppm molybdenum in the total reactor liquid feed (including combined fresh feed, recycled naphtha and additive). Elemental molybdenum is 8 wt% of the additive before dilution in recycled naphtha.

Consumption:

2690 lb/hr of 8% Molybdenum Octoate (undiluted)  
21,207,960 lb/yr for 90% on-stream operation.

Cost:

Manufacturer's estimate, if additive were produced on a large scale, is \$2/lb. Total annual cost is \$42,415,920.

Table 13

REFINERY OPERATING REQUIREMENTS

CATALYST AND CHEMICALS  
(continued)

HYDROPROCESSING CATALYSTS

Naphtha Hydrotreating Unit

Design basis: Ketjen KF-840 hydrotreating catalyst (Nickel-Moly on alumina). Single reactor with 1272 cu.ft. volume, operated at 2.5 liquid hourly space velocity. Initial fill is 59,784 lb. costing \$209,244.

Expect 24 month run-life and 5-year total use with in-situ regeneration after second and fourth year. Replace after fifth year of use. Each regeneration would cost \$89,676.

Distillate Hydrotreating Unit

Design basis: Ketjen KF-840 hydrotreating catalyst (Nickel-Moly on alumina). Three identical reactors with a combined volume of 14,844 cu.ft. operated in series with overall 0.49 liquid hourly space velocity. Initial total fill is 679,668 lb. costing \$2,378,838.

Expect 24-month run-life and 4-year total use with in-situ regeneration once after second year. Replace after fourth year. Each regeneration would cost \$1,019,502.

Distillate Hydrocracking Unit

Design basis: a high-activity hydrocracking catalyst (Nickel-Moly on zeolite). Single reactor with 2477 cu.ft. volume operated at a 2.5 liquid hourly space velocity. Initial fill is 116,420 lb. costing \$1,629,880.

Expect 24 month run-life and 4-year total use with ex-situ regeneration after second years of use. Replace after fourth. Each regeneration would cost \$465,680.

Table 13

REFINERY OPERATING REQUIREMENTSCATALYST AND CHEMICALS  
(continued)HYDROGEN PLANT REQUIREMENTS

Consumption estimated from Stanford Research Institute data.

	<u>Annual consumption cu. ft./year</u>	<u>Annual Cost \$/year</u>
Hydrodesulfurization catalyst	449	47,052
Zinc oxide	1,373	128,111
Reformer catalyst	634	73,838
High-temperature shift catalyst	1,848	107,613
Low-temperature shift catalyst	2,112	368,959
Methanation catalyst	243	28,417
	<u>lb/year</u>	<u>\$/year</u>
Potassium carbonate	369,600	77,565
98% Hydrosulfuric acid	1,845,360	64,521
50% Sodium Hydroxide	546,480	51,011

HYDROGEN PURIFICATION UNIT

Cost of molecular sieves is non-recurring and included in capital cost.

SULFUR RECOVERY UNIT

	<u>\$/year</u>
Claus catalyst (per vendor estimate)	32,430

FLUE GAS DESULFURIZATION UNIT

		<u>\$/year</u>
Estimated from Stanford Research Institute data.		
Soda ash	871 tons/year	99,679
Catalyst		23,319

Table 14

REFINERY OPERATING REQUIREMENTS

## FEEDSTOCK, UTILITIES, AND LABOR

FEEDSTOCKS

Crude oil	50,000 barrels per operating day at \$20 per barrel, \$328,500,000 per year for 90% on-stream operations
Natural gas	Used as reducing gas for Claus Units at the Sulfur Recovery Unit and the Flue Gas Desulfurization Unit.  38.8 MMBTU/hour at \$20 per FOE Barrel (6 MMBTU), \$1,019,664 per year for 90% on-stream operations

UTILITIES

Cooling Water	98,357 gal/min at \$.07/1000 gal \$3,256,876 per year for 90% on-stream operations
Boiler Feed Water	880,000 lb/hr circulation at \$.40/1000 lb \$2,775,000 per year for 90% on-stream operations
Electrical Power	53,294 KW at \$.05/KW•HR \$21,008,495 per year for 90% on-stream operations
Steam	450 psig: 191,384 lb/hr produced 150 psig: 280,767 lb/hr produced 50 psig: 327,207 lb/hr produced

The cost of steam is essentially the cost of boiler feedwater (above), because no imported fuel is used to produce steam. Nearly 69% of all the steam is made using recovered heat, and only 31% is made at the main refinery boiler. The main boiler is fueled by 976 FOE BBL/DAY of the Residuum product from the Hydrovisbreaker, whose cost essentially was already included in the cost of crude oil. Theoretically the 976 FOEB/DAY is equivalent to \$19,520 day, or \$6,412,320 per year (90% on-stream) of unrealized resid sales. Realistically, the resid is difficult to sell.

LABOR AND RELATED EXPENSES

Labor	16 Operator positions per shift:	\$ 2,470,000 per year
	8 Helper positions each shift:	1,080,000 per year
	Total labor:	3,550,000 per year
Supervision @ 25 % of labor:		890,000 per year
Overhead @ 100 % of labor:		<u>3,550,000</u> per year
Total:		\$ 7,990,000 per year



Table 15

ROYALTY PAYMENTS

SOUR WATER STRIPPER AND AMMONIA RECOVERY PLANT

Licensor: Chevron

Process: Chevron WWT Process

Fixed Royalty: \$400,000 paid over 2 years

Running Royalty: \$ 87,500 per year, based on projected recovery of hydrogen sulfide and ammonia. This will vary with actual recovery rates.

SULFUR RECOVERY UNIT

A royalty is required for the BSR/MDEA Tail Gas Unit, which follows the twin Claus sulfur recovery units. There is no royalty on the Claus units.

Licensor: Ralph M. Parsons Company

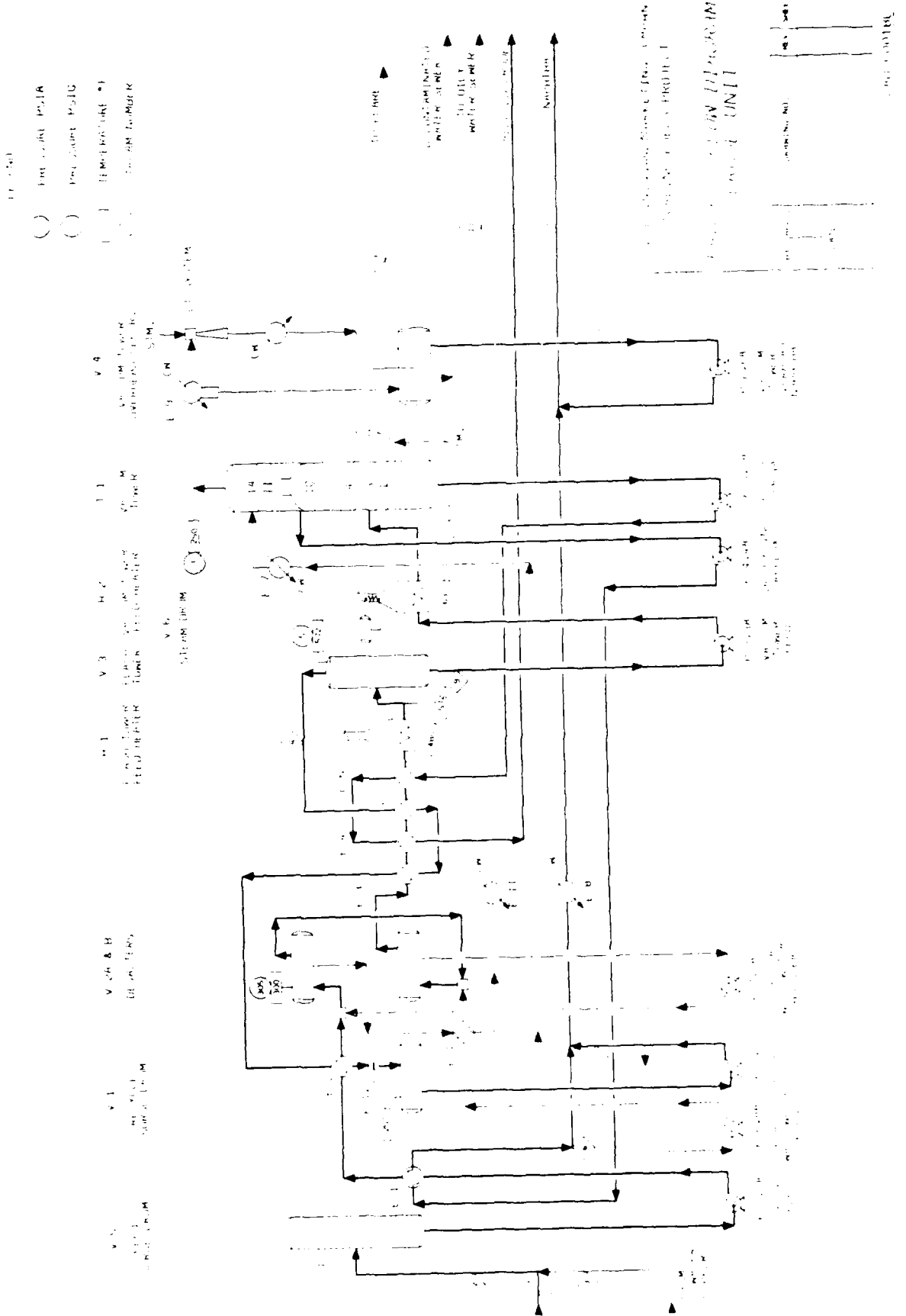
Process: BSR/MDEA Tail Gas Unit

Total royalty: \$172,000 (based on 165 tons/day sulfur production)

25% payable upon signing the licensing agreement  
25% payable upon the start of construction  
25% payable upon completion of mechanical construction  
25% payable after performance guarantee is met

V. PROCESS DESIGN SPECIFICATIONS AND  
PROCESS FLOW DIAGRAMS

**PROCESS DESIGN SPECIFICATIONS**  
for the  
**CRUDE DISTILLATION UNIT**



CRUDE DISTILLATION UNIT  
MATERIAL BALANCE AND STREAM PROPERTIES

Stream Number	1	2	3	4	5	6	7	8	9
Stream Label	Crude Oil Feed	Diluted Crude	Diluent	Flash Vapor	Flash Product	Vacuum Naphtha	Vacuum Tower Overhead	Naphtha Product	Flashed Crude
Stream Conditions									
Temperature, F	150	175	250	572	250	368	250	100	572
Physical state	Liquid	Liquid	Liquid	Vapor	Liquid	Liquid	Vapor	Liquid	Liquid
API Gravity	12.7	17.1	32.8	32.8	32.8	25.4	30.7	28.0	11.7
Sp.Gr. @ 60 F	0.981	0.962	0.861	-	0.961	0.902	0.842	0.887	0.988
Sp.Gr. @ Temp.	0.950	0.910	0.800	-	0.800	0.800	-	-	0.830
Pressure, psia	50	50	50	21	105	7.1	7.0	100	21
BBLs/DAY @ 60 F	50,000	60,000	16,000	18,568	2,560	7,209	778	10,547	47,440
lb/hr	714,000	914,948	200,948	233,089	32,142	94,760	9,543	138,446	683,054
GPM @ 60 F	1,450	1,925	467	-	32,175	210	-	1,384	1,384
Vis., cSt @ Temp.	450	22	0.60	-	0.60	-	-	1.7	2.0

Stream Number	10	11	12	13
Stream Label	Reduced Crude	Desalter Water	Steam	Vacuum Tower Vent
Stream Conditions				
Temperature, F	572	110	300	100
Physical state	Liquid	Liquid	Vapor	Vapor
API Gravity	11.7	10.0	-	(29 mole wt.)
Sp.Gr. @ 60 F	0.988	1.00	-	-
Sp.Gr. @ Temp.	0.830	-	-	-
Pressure, psia	21	100	55	20
BBLs/DAY @ 60 F	47,440	3,960	-	-
lb/hr	683,054	57,750	10,000	400 (a)
GPM @ 60 F	1,384	116	-	-
Vis., cSt @ Temp.	2.0	1	-	-

(a) Non-condensables

## CRUDE UNIT DESIGN BASIS

### Crudes and Rates

The unit is designed to process 50,000 BPSD of 12.7 °API San Ardo California crude. This unit will be capable of maintaining a 95% on stream factor.

### Plant Processing Steps

The crude is first mixed with 16,000 BPSD of recycled distillate and is desalted under pressure at 300°F in the presence of 6 volume percent water containing a demulsifying chemical. This water extraction step is necessary to remove compounds such as sodium chloride which would cause system corrosion and pluggage. The naphtha recycle is necessary to reduce viscosity for effective desalting.

For maximizing thermal efficiency the desalted crude is heated to 480°F by heat exchange against distillation products. The crude is then heated to 570°F and the pressure reduced in a low pressure chamber (flash tower). A fired heater is required to raise the temperature from 480° to 570°F. Light naphtha is vaporized off the flash tower, cooled and liquified to produce the 16,000 BPSD of naphtha recycle required for desalting.

The flashed crude temperature is then further heated to 630°F with fired heat. The heater effluent is fed to a vacuum tower to vaporize an overhead distillate product having a nominal boiling range of 160-650°F. This overhead is fed to the feed splitter tower at the Naphtha Hydrotreater Unit.

## Distillation Products

A small amount of net naphtha product is recovered from the flash tower vapors and is mixed with the light distillate overhead product from the vacuum tower. The crude oil residuum boiling above 650°F is recovered as feedstock for the hydrovisbreaker unit.

<u>Product</u>	<u>BPSD</u>	<u>Volume%</u> <u>of Crude</u>	<u>Nominal</u> <u>Boiling range</u>	<u>Gravity</u> <u>°API</u>
Flashed Naphtha to recycle	16,008	32.0%	160-650°F	32.8
Flashed Naphtha to Product	2,560	5.1%	160-650	32.8
Vac. Twr. Distillate Prod.	8,074	16.1%	160-650	32.8
Total Distillate Products	10,634	21.3%	160-650	32.8
Vac. Twr. Btms. Product	39,366	78.7%	650-1250+	9.1

CRUDE DISTILLATION UNIT

Utilities and Chemical Requirements

Saturated Steam Produced, lb/hr

Steam Generators:	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>
H-1	-	24,160	-	24,160
V-6 at Heater H-2	-	<u>11,370</u>	-	<u>11,370</u>
Totals	-	35,530	-	35,530

Steam Used, lb/hr

	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Condensate Recovered</u>
Heater H1	13,650	-	-	-
Heater H2	6,350	-	-	-
Vacuum jet system	-	11,000	-	-
Miscellaneous	-	-	10,000	-
Totals	<u>20,000</u>	<u>11,000</u>	<u>10,000</u>	-

Net Steam Export                    -20,000            24,530            -10,000

Boiler Feed Water, lb/hr

Steam production  
(rounded)                                    36,000  
For 10% blowdown                        3,600  
Total BFW needed                        39,600

Condensate recovered                    -  
Net BFW make-up                        39,600            =79 GPM

Cooling Water Circulated

<u>Exchanger</u>	<u>Rate</u>	<u>Supply</u>	<u>Return</u>	<u>Duty</u>
E-7	510	85	105	5.1 MMBTU/hr
E-8	700	85	105	7.0
E-9 Vacuum System	<u>1733</u>	85	100	<u>13.0</u>
Total	2943			25.1 MMBTU/hr

Make-up, at 3% circulation rate            88 GPM

CRUDE DISTILLATION UNIT

Utilities and Chemical Requirements  
(continued)

Heater Fuel Fired (Hydrovisbreaker vacuum residuum)

Heater H-1	115.44 MMBTU/HR	
Heater H-2	53.7	
	<u>169.1</u> MMBTU/HR	= 28.1 BBL/hr
		= 9,840 lb/hr Residuum Fuel

Electrical Power

<u>Pumps:</u>	<u>Brake Horsepower Operating</u>	<u>Brake Horsepower Connected</u>
P-1,1A Crude feed	700	1,400
P-2,2A Diluent recycle	50	100
P-3,3A Vacuum Tower feed	325	650
P-4,4A Circulating reflux	125	250
P-5,5A Reduced crude	200	200
P-6,6A Vacuum tower overhead	6	8
P-7,7A Wash water booster	20	40
P-8,8A Desalter wash water	60	120
P-9,9A Boiler feed water	20	40
P-10,10A Desalting chemical	1	2
 <u>Other:</u>		
V-2A,B Desalters	<u>350</u>	<u>350</u>
Total Brake Horsepower	1855 BHP	3360 BHP
Kilowatts (220 & 440V)	1383 KW	2505 KW

Air Requirements

Instrument air	50 psig	100 SCFM
Plant air	120 psig	

Chemicals

Demulsifier for desalters	33 gal/day
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## CRUDE DISTILLATION UNIT

### List of Major Equipment

#### Heat Exchangers

E-1	Circulating reflux - Desalter feed exchanger
E-2	Flash tower overhead - Desalter feed exchanger
E-3	Flash tower preheater - No. 1
E-4	Flash tower preheater - No. 2
E-5	Flash tower preheater - No. 3
E-6	Flash tower preheater - No. 4
E-7	Circulating reflux trim cooler
E-8	Naphtha product cooler
E-9	Vacuum tower overhead condenser
E-10	Desalter water feed - effluent exchanger
E-11	Desalter effluent water cooler
E-12	Vacuum tower after-condenser
E-13	Recycle air cooler

#### Fired Heaters

H-1	Flash tower feed heater
H-2	Vacuum tower feed heater

#### Pumps

P-1,1A	Crude feed pumps
P-2,2A	Recycle pumps
P-3,3A	Vacuum tower feed pumps
P-4,4A	Circulating reflux pumps
P-5,5A	Reduced crude pumps
P-6,6A	Vacuum tower overhead naphtha product pumps
P-7,7A	Desalter wash water booster pumps
P-8,8A	Desalter wash water feed pumps
P-9,9A	Boiler feed water pumps
P-10,10A	Desalting chemical pumps

#### Distillation towers

T-1	Vacuum distillation tower
-----	---------------------------

#### Vessels

V-1	Recycle surge drum
V-2A	Desalter No. 1
V-2B	Desalter No. 2
V-3	Flash tower
V-4	Vacuum tower overhead receiver
V-5	Crude feed surge drum
V-6	Steam drum

CRUDE DISTILLATION UNIT

San Ardo Crude Oil Assay Data

<u>Whole Crude Properties</u>	<u>1984 Analyses</u>	<u>1985 Analyses</u>
Gravity, °API	12.8	12.68
Sp.Gr. @ 60°F	0.9806	0.9814
Density, g/cc		0.9523
Kinematic Visc. @ 77°F	12,698.2	-
@ 100°F	3,293.6	3,935.55
@ 210°F	-	77.37
Pour Point, °F	35	-
Flash Point, °F	194	-
Carbon, Wt%	81.62	-
Hydrogen, Wt%	10.51	-
Oxygen, Wt%	1.82	-
Sulfur, Wt%	1.89	1.9
Nitrogen, Basic, ppm	2,287 (.2287 wt%)	3009
Total, ppm	13,200	10,544
Metals, ppm		
Fe	42	59
Ni	78	70
Cu	0.9	<0.1
V	96	68
Ash, Wt%	0.12	-
Ramscarbon, Wt%	8.68	-
Salt content, lb/1000 BBL	34	6.27
BS&W, Vol%	-	0.34

True Boiling Point Fractions:

	<u>Spec. Grav. 60/60°F</u>		<u>Kinematic Viscosity, cSt</u>			<u>Sulfur</u>
	<u>Hydrometer</u>	<u>Densitometer</u>	<u>100°F</u>	<u>160°F</u>	<u>210°F</u>	<u>Wt%</u>
IBP-400°F	0.8272	0.8224	1.18	0.80	-	0.328
400-475	0.8655	0.8617	1.94	-	0.76	0.446
475-550	0.8874	0.8877	3.33	-	2.08	0.822
550-650	0.9124	0.9105	8.49	-	2.10	1.119
650-725	0.9377	0.9372	40.15	-	4.61	1.365
725-800	0.9486	0.9479	88.02	-	6.73	1.376
800-900	0.9673	0.9674	829.40	-	18.81	1.372
900-932	0.9716	0.9761		189.61	-	1.504
932+	1.0329	-	-	-	-	-
Whole Crude	0.9806	0.9814	3935.55	-	77.37	1.9

CRUDE DISTILLATION UNIT  
San Ardo Crude Oil Assay Data  
(continued)

True Boiling Point Distillation

<u>Accumulated Volume % Distilled</u>	<u>Temperature</u>
0	179
2	335
5	445
7	465
10	504
15	560
20	612
25	655
30	708
35	752
40	795
45	840
50	883
55	925
60	968
65	1010
70	1050
75	1088
80	1126
85	1165
90	1205
95	1300
100	1300+

NOTES  
SR - STRESS RELIEVED  
IR - X RAYED

ITEM NO	SERVICE	UNIT PERFORMANCE												CONSTRUCTION			NO	DATE	REVISION	BY																																												
		DESIGN				OPERATION				ANALYSIS				MATERIAL	THICKNESS	MATERIAL					THICKNESS																																											
1	2	3	4	5	6	7	8	9	10	11	12	13	14				15	16	17	18		19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54	55	56	57	58	59	60	61
1	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER	CONDENSER

Sun Refining and Marketing Company

HEAT EXCHANGER SCHEDULE

PLANT NO: UNIT NO:

LOCATION: BY APPT: DRAWING NO: REV SHEET:

DATE:

ITEM NO	DESCRIPTION	QTY	UNIT	DATE	BY	REVISION	WT
1	MANUFACTURER - NO SHELLS						
2	SIZE AND TYPE						
3	SURFACE/UNIT - SHELL						
4	CONNECTED IN						
5	FLUID CIRCULATED						
6	QUANTITY						
7	WEIGHT						
8	STEAM						
9	TOTAL						
10	LIQUID VAPORIZED OR CONDENSED						
11	STEAM CONDENSED						
12	GRAVITY LIQUID (SG) AT 60°F						
13	GRAVITY LIQUID (SG) AT 90°F						
14	VISCOSITY LIQUID (C.P.) AT 60°F						
15	VISCOSITY LIQUID (C.P.) AT 90°F						
16	SPEC HEAT LIQUID - BTU/LB						
17	SPEC HEAT VAPOR - BTU/LB						
18	OPER TEMP °F IN						
19	OPER TEMP °F OUT						
20	OPER PRESSURE - PSIG						
21	DESIGN PRESSURE - PSIG						
22	DESIGN TEMPERATURE - °F						
23	CORROSION ALLOWANCE						
24	CONNECTIONS - IN						
25	SIZE STD TYPE TUBES						
26	SIZE STD TYPE TUBES						
27	TUBES - O.D.						
28	TUBES - I.D.						
29	LENGTH						
30	PITCH						
31	SHELL I.D.						
32	SHELL O.D.						
33	STATIONARY TUBE SHEET						
34	FLOATING TUBE SHEET						
35	CROSS BAFFLES						
36	LONGITUDINAL BAFFLES						
37	TUBE SUPPORTS						
38	SHELL						
39	SHELL COVERS						
40	FLOATING HEAD COVER						
41	CHANNEL COVER						
42	CROSS BAFFLES - TYPE						
43	LONGITUDINAL BAFFLE TYPE						
44	GASKETS						
45	WELDING FLANGES AND NOZZLES						
46	STAYS						
47	HEAT EXCHANGER - UNIT						
48	WEIGHT - LB						
49	WEIGHT - KG						
50	W.M.W. DRAWING NO						
51	REVISION NO						

NOTES  
SR STRESSES RELIEVED  
XR X RAYED

Sun Refining and  
Marketing Company

HEAT EXCHANGER SCHEDULE

PLANT NO. \_\_\_\_\_ UNIT NO. \_\_\_\_\_  
 LOCATION \_\_\_\_\_  
 BT APPD. \_\_\_\_\_ DRAWING NO. \_\_\_\_\_ REV SHEET \_\_\_\_\_  
 DATE \_\_\_\_\_  
 4-15/87

ITEM NO	DESCRIPTION	UNIT	QTY	UNIT PRICE	TOTAL	REMARKS
1	DESIGNER EFFLUENT WATER	1600				E-11
2	MANUFACTURER - NO SHELLS					E-12
3	SIZE AND TYPE					E-13
4	SURFACE/UNIT - /SHELL					
5	CONNECTED IN					
6	FLUID CIRCULATED					
7	QUANTITY					
8	PIPED GASES					
9	STEAM					
10	TOTAL					
11	FLUID VAPORIZED OR CONDENSED					
12	STEAM CONDENSED					
13	GRAVITY LIQUID (SG @ 60°F)					
14	VISCOSITY LIQUID (CENTIPOISE @ 60°F)					
15	SPEC HEAT LIQUID - BTU/LB °F					
16	LATENT HEAT VAPOR - BTU/LB					
17	OPER TEMP °F IN					
18	OPER TEMP °F OUT					
19	OPER PRESSURE - PSIG					
20	MINIMUM FOULING FACTOR					
21	NUMBER OF PASSES					
22	VELOCITY - FT/SEC					
23	PRESSURE DROP - PSIG					
24	HEAT EXCHANGED - BTU/HR					
25	WTD CORRECTED WEIGHTED TRANSFER RATE SERVICE - CLEAN					
26	DESIGN PRESSURE - PSIG					
27	TEST PRESSURE - PSIG					
28	DESIGN TEMPERATURE - °F					
29	CORROSION ALLOWANCE					
30	CONNECTIONS - IN					
31	SIZE STD TYPE					
32	TUBES - OD					
33	LENGTH					
34	PITCH					
35	SHELL - ID					
36	STATIONARY TUBE SHEET					
37	FLOATING TUBE SHEET					
38	CROSS BAFFLES					
39	LONGITUDINAL BAFFLES					
40	TUBE SUPPORTS					
41	SHELL COVERS					
42	FLOATING HEAD COVER					
43	CHANNEL COVER					
44	CROSS BAFFLES TYPE - SPACING					
45	LONGITUDINAL BAFFLE TYPE					
46	GASKETS					
47	WELDING FLANGES AND NOZZLES					
48	STUDS					
49	WEIGHT EXCHANGER					
50	WEIGHT BUNDLE					
51	CODE REQUIREMENTS					
52	NUM OIL CO DRAWING NO					
53	ILLUSTRATION NO					
54	PURCHASE ORDER NO					

**NOTES**  
 SR - STRESS RELIEVED  
 IR - X-RAYED

Sun Refining and Marketing Company  
 HEAT EXCHANGER SCHEDULE

PLANT NO: \_\_\_\_\_ UNIT NO: \_\_\_\_\_  
 LOCATION: \_\_\_\_\_  
 BY: APPD \_\_\_\_\_ DRAWING NO: \_\_\_\_\_ REV SHEET: \_\_\_\_\_  
 DATE: 4/25/87

SERVICE ITEM NO. 4-1  
 SERVICE FURNACE FRED HEATER

DESIGN CONDITIONS  
 TOTAL OUT OF HEATER, MM STU PER HR ASSIGNED/FIRED 79.2

HEATER SECTION  
 SERVICE RAVANT  
 HEAT ABSORPTION MM STU/HR 692  
 FLOW RATE 1800 (SID)  
 FLOW RATE SPEC (SID) 66,000  
 PRESSURE DROP PAI (CALCULATED) 125  
 PRESSURE DROP PAI (ALLOWABLE) 900

FOULING FACTOR

HEAT CONDITIONS  
 TEMPERATURE DEG F 480  
 PRESSURE (PSIA) 185  
 LIQUID FLOW LB/HR 116,000  
 VAPOR FLOW LB/HR 10,000  
 LIQUID IDER APPLICABILITY  
 VAPOR IDER APPLICABILITY  
 LIQUID IDER APPLICABILITY  
 VAPOR IDER APPLICABILITY  
 LIQUID IDER APPLICABILITY  
 VAPOR IDER APPLICABILITY  
 LIQUID IDER APPLICABILITY  
 VAPOR IDER APPLICABILITY

HEATER CHARACTERISTICS:  
 TYPE OF FUEL  
 HEATING VALUE LHV  
 SPECIFIC GRAVITY  
 SULFUR/NITROGEN WEIGHT %

HEATER SECTION  
 OVERALL PRESSURE DROP  
 SURFACE FLOW TEMPERATURE DEG F  
 CIRCULATION FLOW RATE  
 NUMBER OF PASSES  
 HEATING VALUE TEMPERATURE DEG F  
 SPECIFIC GRAVITY  
 SULFUR/NITROGEN WEIGHT %

HEATER SECTION  
 OVERALL PRESSURE DROP  
 SURFACE FLOW TEMPERATURE DEG F  
 CIRCULATION FLOW RATE  
 NUMBER OF PASSES  
 HEATING VALUE TEMPERATURE DEG F  
 SPECIFIC GRAVITY  
 SULFUR/NITROGEN WEIGHT %

HEATER SECTION  
 OVERALL PRESSURE DROP  
 SURFACE FLOW TEMPERATURE DEG F  
 CIRCULATION FLOW RATE  
 NUMBER OF PASSES  
 HEATING VALUE TEMPERATURE DEG F  
 SPECIFIC GRAVITY  
 SULFUR/NITROGEN WEIGHT %

REFERENCES: SURFACE TUBE NUMBER  
 TUBE SPACING (CENTER TO CENTER IN ISLANDS/SECTION LINE)  
 TUBE CENTER TO SURFACE WALL IN MM  
 HEAT TREATMENT  
 WELD INSPECTION REQUIREMENTS (RAY OR OTHER)  
 TUBES  
 MATERIALS OR NON-DESIGN  
 TUBE MATERIALS (HEAT TREATMENT AND STRESS)  
 OUTSIDE DIAMETER IN

NOTES

Sun Refining and Marketing Company  
 FIRED HEATER DESIGN

PLANT NO. UNIT NO.  
 LOCATION  
 BY APPD. DRAWING NO. REV SHEET  
 DATE 4.25.77 4

SERVICE ITEM NO.

HEATER

41-2

MANUFACTURER

HEATER

DATE

4/12/53

TOTAL DUTY OF HEATER IN BTU PER HR. ALLOWED/FIRED

PROCESS DESIGN CONDITIONS

RADIANT SECTION

PROCESS

2

HEATING VALUE

11,440

HEATER SECTION

HEAT ABSORPTION IN BTU/HR

32

HEATING VALUE

11,440

HEATING VALUE

11,440

FLUID

FLOW RATE, LB/HR

683,100

HEATING VALUE

11,440

HEATING VALUE

11,440

FLUID

FLOW RATE, LB/HR

87,600

HEATING VALUE

11,440

HEATING VALUE

11,440

PRESSURE DROP, PSI (CALCULATED)

1.00

HEATING VALUE

11,440

HEATING VALUE

11,440

AVERAGE FLOW VELOCITY, FT/HR

1,800

HEATING VALUE

11,440

HEATING VALUE

11,440

FOULING FACTOR

1.00

HEATING VALUE

11,440

HEATING VALUE

11,440

INLET CONDITIONS

TEMPERATURE DEG F

574

HEATING VALUE

11,440

HEATING VALUE

11,440

TEMPERATURE DEG F

120

HEATING VALUE

11,440

HEATING VALUE

11,440

TEMPERATURE DEG F

165

HEATING VALUE

11,440

HEATING VALUE

11,440

LIQUID FLOW, LB/HR

683,100

HEATING VALUE

11,440

HEATING VALUE

11,440

LIQUID FLOW, LB/HR

87,600

HEATING VALUE

11,440

HEATING VALUE

11,440

WATER FLOW, LB/HR

18

HEATING VALUE

11,440

HEATING VALUE

11,440

WATER FLOW, LB/HR

18

HEATING VALUE

11,440

HEATING VALUE

11,440

WATER FLOW, LB/HR

18

HEATING VALUE

11,440

HEATING VALUE

11,440

WATER FLOW, LB/HR

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HEATING VALUE

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HEATING VALUE

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WATER FLOW, LB/HR

18

HEATING VALUE

11,440

HEATING VALUE

11,440

WATER FLOW, LB/HR

18

HEATING VALUE

11,440

HEATING VALUE

11,440

WATER FLOW, LB/HR

18

HEATING VALUE

11,440

HEATING VALUE

11,440

WATER FLOW, LB/HR

18

HEATING VALUE

11,440

HEATING VALUE

11,440

NOTES

PLANT: HEATER UNIT NO: 1

DESIGNER: HEATER

DATE: 4/12/53

REVISION: BY

Sun Refining and Marketing Company

FIRED HEATER DESIGN

PLANT: HEATER UNIT NO: 1

DESIGNER: HEATER

DATE: 4/12/53

REVISION: BY

Sun Refining and Marketing Company

FIRED HEATER DESIGN

PLANT: HEATER UNIT NO: 1

DESIGNER: HEATER

DATE: 4/12/53

REVISION: BY

Sun Refining and Marketing Company

FIRED HEATER DESIGN

PLANT: HEATER UNIT NO: 1

DESIGNER: HEATER

DATE: 4/12/53

REVISION: BY

Sun Refining and Marketing Company

FIRED HEATER DESIGN



PUMP ITEM NO	SERVICE	PUMP DATA										PUMP NO	REVISED	DATE	BY	
		1	2	3	4	5	6	7	8	9	10					
1	LIQUID	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
2	PUMPING TEMPERATURE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
3	SPECIFIC GRAVITY AT P T	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
4	VISC. SSU. CP. CORR. AT P T	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
5	GPM AT P T	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
6	SUCTION PRESSURE (PSIG)	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
7	DISCHARGE PRESSURE (PSIG)	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
8	DIFFERENTIAL HEAD (FT)	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
9	NPSH AVAILABLE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
10	RPM	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
11	BHP AT RATED GPM	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
12	MAX BHP FOR INSTALLED IMP	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
13	MANUFACTURER	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
14	TYPE AND SIZE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
15	STAGES	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
16	SUCT. FLG. SIZE, RATING, FACE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
17	DISCH. FLG. SIZE, RATING, FACE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
18	NOZZLE ARGGT. SUCTION	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
19	NOZZLE ARGGT. DISCHARGE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
20	IMPELLER DIA. REQUIRED	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
21	MAX. IMP. DIA. AND HEAD	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
22	MP. EYE AREA	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
23	NPSH REQUIRED	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
24	HYDRASTATIC TEST PRESSURE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
25	THRUST BEARING	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
26	RACIAL BEARING	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
27	COUPLING	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
28	IMPELLER GUARD	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
29	ROTATION FROM COUPLING END	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
30	MECHANICAL SEALS TYPE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
31	MECHANICAL SEALS TYPE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
32	BOX	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
33	BOX	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
34	WATER COOLED BEARING	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
35	PEDESTAL	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
36	WEIGHT OF PUMP AND BASE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
37	CASTING	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
38	CAST BEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
39	SPARE PIECES	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
40	THROAT	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
41	DISCHARGE INTERSEAL	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
42	IMPELLER	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
43	IMPELLER WEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
44	SHAFT SLEEVES	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
45	SHAFT	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
46	SHAFT	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
47	ANTHRAX PUMP	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
48	BEARING HOUSING	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
49	BASE PLATE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
50	IMPELLER	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
51	IMPELLER WEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
52	IMPELLER WEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
53	IMPELLER WEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
54	IMPELLER WEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
55	IMPELLER WEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
56	IMPELLER WEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE
57	IMPELLER WEARING RINGS	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE	CLONE

**NOTES**

NO. DATE REVISION BY

Sun Refining and Marketing Company

CENTRIFUGAL PUMP SCHEDULE

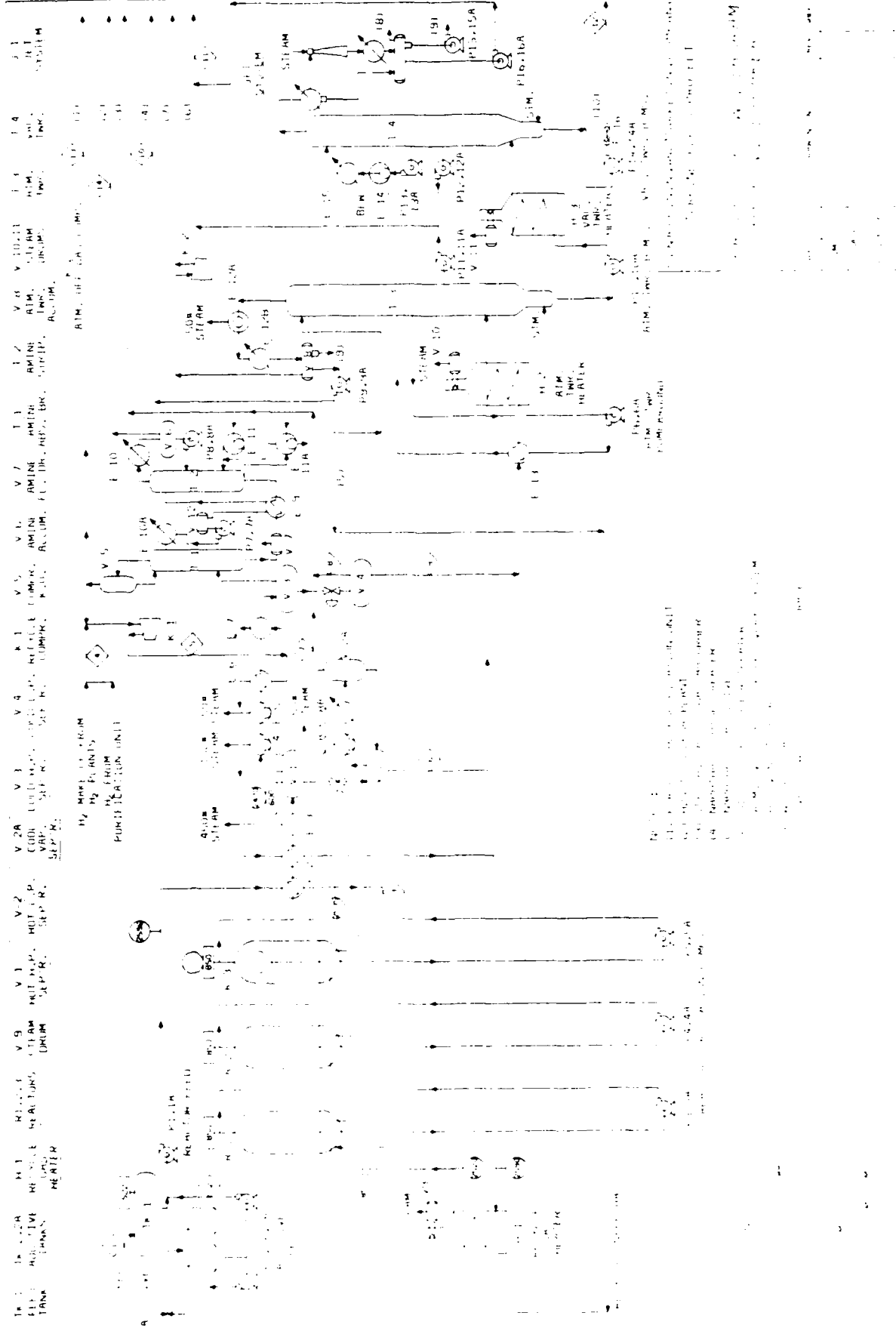
PLANT	UNIT NO.
CLIENT	UNIT NO.
DATE	REV. SHEET
DRAWING NO.	REV. SHEET



**PROCESS DESIGN SPECIFICATIONS**

for the

**HYDROVISBREAKER UNIT**



V.1	AMINE TANK	V.2	HOT VAP. SEPR.	V.3	HOT VAP. SEPR.	V.4	AMINE FLASH TANK	V.5	AMINE FL. SEPR.	V.6	AMINE FL. SEPR.	V.7	AMINE FL. SEPR.	V.8	AMINE FL. SEPR.	V.9	STEAM DRUM	V.10	STEAM DRUM	V.11	STEAM SEPR.	V.12	NITROGEN TANK	V.13	NITROGEN SEPR.	V.14	NITROGEN HEATER
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HYDROVISBREAKER UNIT  
MATERIAL BALANCE AND STREAM PROPERTIES

Stream Number	1	2	3	4	5	6	7	8	9
Stream Label	Fresh Feed	Additive	Reactor Liquid Feed	Hydrogen Make-up	Total Gas Feed to Reactor 1	1st Low Pressure Separator Liquid	1st Low Pressure Sep. Vapor (from V-2A)	2nd Low Pressure Vapor (from V-4)	2nd Low Pressure Liquid (from V-4)
Overhead Stream Conditions									
Temperature, F	480	60	750	125	131	565	110	107	107
Physical state	Liquid	Liquid	Liquid	Vapor	Vapor	Liquid	Vapor	Vapor	Liquid
API Gravity	9.1	47.1	9.7	-	-	15.3	-	-	50.7
Sp.Gr. @ 60 F	0.87	0.79	0.83	(2.7)	(4.4)	0.78	(10.2)	(16.2)	0.76
Sp.Gr. @ Temp. Pressure, psia	20	20	2550	2450	2635	165	155	165	165
BBLS/DAY @ 60 F	39,366	815	40,181	-	-	30,326	-	-	12,245
MMSCFD, (60 F, 1 atm)	-	-	-	40.0	229.5	-	5.2	2.86	-
M lb/hr	577.6	9.4	587.0	11.96	111.6	426.0	5.85	5.1	138.6
GPM @ 60 F	2.2	2.6	0.9	-	-	0.8	-	-	1.0
Vis., cSt @ Temp.									

Stream Number	10	11	12	13	14
Stream Label	Naphtha Product	490-1000 F Distillate	Vacuum Tower Bottoms	High-Pressure Hydrogen Vent	H <sub>2</sub> S from Amine
Stream Conditions					
Temperature, F	100	495	380	100	120
Physical state	Liquid	Liquid	Liquid	Vapor	Vapor
API Gravity	52.5	17.4	-1.01	-	-
Sp.Gr. @ 60 F	0.77	0.80	1.0	(5.2 Mol Wt)	(32.4 Mol Wt)
Sp.Gr. @ Temp. Pressure, psia	90	80	145	2440	23
BRLS/DAY @ 60 F	12,045	24,377	4,839	-	-
MMSCFD, (60 F, 1 atm)	-	-	-	29.96	1.72
M lb/hr	135.0	337.6	78.5	17.1	6.126
GPM @ 60 F	0.6	0.9	15	-	-
Vis., cSt @ Temp.					

## HYDROVISBREAKER DESIGN BASIS

### Unit Feed Rate

The unit is designed to hydroprocess 39,366 BPSD of San Ardo reduced crude boiling above a true-boiling-point 5% point of 650°F. The unit will convert 71 vol% of the feed boiling above 975°F to lighter hydrocarbons. It is anticipated that a 90% on-stream factor can be maintained.

### Plant Processing Steps

The 650°F crude unit bottoms at a temperature of 450° to 480°F is mixed with an additive that reduces coke formation in the reactor system. This mixed stream is preheated by heat exchange with the reactor effluent and charged to the reactor system. Three reactors in series with internal circulation are used.

The reactor system operates at 2500 psig and 850°F and a one liquid hourly space velocity. External circulating pumps are used with block valving for maintenance access. Preheated circulating hydrogen is injected into the first reactor at 750°F. A total of 5700 SCF of 85% hydrogen is circulated per barrel of reactor feed.

The effluent mix of hydrogen gas and liquid product from the reactors is cooled to 600°F by heat exchange and separated. The hydrogen gas is further cooled to 100°F, separated, compressed, and recirculated along with make-up hydrogen.

Fresh hydrogen makeup gas is obtained from two sources. The hydrogen plant supplies hydrogen of 95% purity. This source alone is insufficient to maintain the desired hydrogen purity and partial pressure in the reactor gas

feed, so a hydrogen purification unit is used to recover hydrogen at 99% purity from the high pressure bleed gas of the Hydrovisbreaker and Distillate Hydrotreater Units.

The hydrocarbon liquids separated in the 600°F and 100°F separators (hot and cold high-pressure separators, V-1 and V-3) are reduced in pressure and vented in V-2 and V-4 prior to reheating for fractionation. A naphtha cut, light and heavy distillate cuts and resid fuel cuts are recovered in the fractionation. This requires an atmospheric and vacuum tower.

HYDROVISBREAKING UNIT  
Utility and Chemical Requirements

Steam Balance

<u>Steam Production, lb/hr</u>	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Totals</u>
Steam generators:				
E-3	32,450	-	-	32,450
E-4	-	66,570	-	66,570
E-5	-	-	25,100	25,100
E-8A	-	-	8,035	8,035
E-12A	-	-	18,840	18,840
E-14	-	-	38,650	38,650
V-10 at Heater H-1	-	24,040	-	24,040
V-11 at Heater H-2	-	22,170	-	22,170
V-12 at Heater H-3	-	13,245	-	13,245
Totals	32,450	126,025	90,625	249,100

<u>Steam consumed, lb/hr</u>	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Recovered Condensate</u>
E-11 Reboiler	-	-	31,200	31,200
E-11A Reclaimer	-	310	-	310
T-3 stripping	-	11,150	-	-
T-4 stripping	-	10,000	-	-
H-1 heater	1,300	-	-	-
H-2 heater	9,200	-	-	-
H-3 heater	8,700	-	-	-
Vacuum Jet Motive Steam	-	10,000	-	-
Miscellaneous	-	-	10,000	-
Totals	19,200	31,460	41,200	31,510
<u>Net steam export, lb/hr</u>	13,000	94,000	49,000	

Boiler Feed Water Required

Steam production (rounded)	250,000 lb/hr
10% condensate loss to blowdown	25,000
Total gross BFW needed	275,000 lb/hr
Total condensate recovered for use as BFW	- 31,500
Net Boiler Feed Water make-up required	243,500 lb/hr = 487 GPM



HYDROVISBREAKING UNIT  
Utility and Chemical Requirements  
(continued)

Fuel fired (Hydrovisbreaker Vacuum Residuum)

H-1 heater	110.1 MMBTU/HR
H-2 heater	101.1
H-3 heater	<u>60.7</u>
Total	271.9 MMBTU/HR = 45.2 FOE Bbl/hr = 15,822 #/hr Resid fuel

Air Requirements

Instrument Air	50 psig	300 SCFM
Plant Air	120 psig	

<u>Cooling Water Circulation</u>	<u>GPM</u>	<u>Supply</u>	<u>Return</u>	<u>MMBTU/HR</u>
E-7 cooler	2480	85°F	110°F	31.00
E-8B	510	85	110	6.37
E-10 Amine condenser		85		
E-10A Amine cooler		85		
E-12B Condenser	3738	85	110	46.73
Vac. Jet System	<u>4500</u>	<u>85</u>	<u>100</u>	<u>33.75</u>
Totals	11,228	85	106	117.85
Assume 3% make-up	337			

Coke-suppressing Additive

8% Molybdenum Octoate dissolved in Naphtha to establish 367 wppm Molybdenum in the Total Reactor Liquid Feed.

8% Molybdenum Octoate	2,690 lb/hr
(Molybdenum 2-ethyl hexoate)	
Naphtha	6,724 lb/hr
Total additive mixture	9,414 lb/hr

HYDROVISBREAKING UNIT  
Utility and Chemical Requirements  
 (continued)

Electical Power Requirement

<u>Equipment</u>	<u>Operating Horsepower</u>	<u>Connected Horsepower</u>
Pumps:		
P-1,1A Reactor Feed	3800	7600
P-2,2A Additive Feed	5	10
P-3,3A Reactor R-1 Circulation	100	200
P-4,4A Reactor R-2 Circulation	100	200
P-5,5A Reactor R-2 Circulation	100	200
P-6,6A Atmospheric Tower Pumpharound	25	50
P-7,7A Lean Amine	1100	2200
P-8,8A Amine Reflux	4	8
P-9,9A Atmospheric Tower Reflux & Product	40	80
P-10,10A Atmospheric Tower Bottoms	75	150
P-11,11A Atmospheric Tower Gas Oil Sidedraw	15	30
P-12,12A Vacuum Tower Gas Oil Sidedraw	50	100
P-13,13A Vacuum Tower Pumpharound	60	120
P-14,14A Vacuum Tower Residuum	35	70
P-15,15A Vacuum Tower Sour Water	3	6
P-16,16A Vacuum Tower Bottoms Residuum	8	16
Compressors:		
K-1 Recycle Gas	1500	1500
K-2 Atmospheric Tower Off-Gas	<u>50</u>	<u>100</u>
TOTALS	7020 BHP	12,540 BHP
KILOWATTS = (.7457 x HORSEPOWER)	5235 KW	9,351 KW

## HYDROVISBREAKING UNIT

### Major Equipment List

#### Heat Exchangers

E-1	Reactor Feed - Effluent Exchanger
E-2	Reactor Effluent - Recycle Gas Exchanger
E-3	Reactor Effluent - 450 psig Steam Generator
E-4	Hot Separator Vapor - 150 psig Steam Generator
E-5	Hot Separator Vapor - 50 psig Steam Generator
E-6	Hot Separator Vapor - Recycle Gas Exchanger
E-7	Hot Separator Vapor Cooler
E-8A	Hot Low-Pressure Separator Gas - Steam Generator
E-8B	Hot Low-Pressure Separator Gas Cooler
E-9	Rich Amine - Lean Amine Exchanger
E-10	Amine Stripper Overhead Condenser
E-10A	Lean Amine Cooler
E-11	Amine Stripper Reboiler
E-11A	Amine Reclaimer
E-12A	Atmospheric Tower Overhead - 50 psig Steam Generator
E-12B	Atmospheric Tower Overhead Condenser
E-13	Atmospheric Tower Feed - Pumpharound Exchanger
E-14	Vacuum Tower Pumpharound - 50 psig Steam Generator
E-15	Vacuum Tower Pumpharound - Boiler Feed Water Exchanger
E-16	Vacuum Tower Bottoms Aircooler

#### Fired Heaters:

H-1	Recycle Gas Heater
H-2	Atmospheric Tower Feed Heater
H-3	Vacuum Tower Feed Heater

#### Towers, Tanks, and Reactors:

T-1	Amine Absorber
T-2	Amine Stripper (Regenerator)
T-3	Atmospheric Tower
T-4	Vacuum Tower
TK-1	Hydrovisbreaker Feed Tank
TK-2,2A	Coke-Suppressing Additive Feed Tanks
R-1,2,3	Hydrovisbreaker Reactors

Vacuum Jet System - Consisting of Vacuum Jets, Condensers, and Knockout Drum

## HYDROVISBREAKING UNIT

### Major Equipment List (continued)

#### Vessels:

V-1	Hot High-Pressure Separator
V-2	Hot Low -Pressure Separator
V-3	Cold High-Pressure Separator
V-4	Cold Low -Pressure Separator
V-5	Recycle Compressor Knockout Drum
V-6	Amine Stripper Overhead Accumulator
V-7	Amine Flash Drum
V-8	Atmospheric Tower Overhead Accumulator
V-9	Steam Drum at H-1 Recycle Gas Heater
V-10	Steam Drum at H-2 Atmospheric Tower Feed Heater
V-11	Steam Drum at H-3 Vacuum Tower Feed Heater
V-12	Lean Amine Surge Drum

#### Compressors and Pumps:

K-1	Recycle Gas Compressor
K-2	Atmospheric Tower Off-Gas Compressor
P-1, 1A	Reactor Feed
P-2, 2A	Coke-Suppressing Additive Feed
P-3, 3A	Reactor R-1 Circulation
P-4, 4A	Reactor R-2 Circulation
P-5, 5A	Reactor R-3 Circulation
P-6, 6A	Atmospheric Tower Pumparound
P-7, 7A	Lean Amine
P-8, 8A	Amine Stripper Reflux
P-9, 9A	Atmospheric Tower Overhead Reflux and Product
P-10, 10A	Atmospheric Tower Bottoms (Vacuum Tower Feed)
P-11, 11A	Atmospheric Tower Sidestream Gas Oil
P-12, 12A	Vacuum Tower Sidestream Gas Oil
P-13, 13A	Vacuum Tower Pumparound
P-14, 14A	Vacuum Tower Bottoms (Refinery Vacuum Residuum Product)
P-15, 15A	Vacuum Overhead Accumulator Sour Water
P-16, 16A	Vacuum Tower Overhead Gas Oil Product

NOTES  
 50 - STRESS RELIEVED  
 10 - STAYED

ITEM NO	DESCRIPTION	E-1	E-2	E-3	E-4	E-5
1	REACTOR FEED EFFLUENT	32,000	32,000	40,000	40,000	40,000
2	REACTOR FEED EFFLUENT	32,000	32,000	40,000	40,000	40,000
3	EFFLUENT	32,000	32,000	40,000	40,000	40,000
4	EFFLUENT	32,000	32,000	40,000	40,000	40,000
5	EFFLUENT	32,000	32,000	40,000	40,000	40,000
6	EFFLUENT	32,000	32,000	40,000	40,000	40,000
7	EFFLUENT	32,000	32,000	40,000	40,000	40,000
8	EFFLUENT	32,000	32,000	40,000	40,000	40,000
9	EFFLUENT	32,000	32,000	40,000	40,000	40,000
10	EFFLUENT	32,000	32,000	40,000	40,000	40,000
11	EFFLUENT	32,000	32,000	40,000	40,000	40,000
12	EFFLUENT	32,000	32,000	40,000	40,000	40,000
13	EFFLUENT	32,000	32,000	40,000	40,000	40,000
14	EFFLUENT	32,000	32,000	40,000	40,000	40,000
15	EFFLUENT	32,000	32,000	40,000	40,000	40,000
16	EFFLUENT	32,000	32,000	40,000	40,000	40,000
17	EFFLUENT	32,000	32,000	40,000	40,000	40,000
18	EFFLUENT	32,000	32,000	40,000	40,000	40,000
19	EFFLUENT	32,000	32,000	40,000	40,000	40,000
20	EFFLUENT	32,000	32,000	40,000	40,000	40,000
21	EFFLUENT	32,000	32,000	40,000	40,000	40,000
22	EFFLUENT	32,000	32,000	40,000	40,000	40,000
23	EFFLUENT	32,000	32,000	40,000	40,000	40,000
24	EFFLUENT	32,000	32,000	40,000	40,000	40,000
25	EFFLUENT	32,000	32,000	40,000	40,000	40,000
26	EFFLUENT	32,000	32,000	40,000	40,000	40,000
27	EFFLUENT	32,000	32,000	40,000	40,000	40,000
28	EFFLUENT	32,000	32,000	40,000	40,000	40,000
29	EFFLUENT	32,000	32,000	40,000	40,000	40,000
30	EFFLUENT	32,000	32,000	40,000	40,000	40,000
31	EFFLUENT	32,000	32,000	40,000	40,000	40,000
32	EFFLUENT	32,000	32,000	40,000	40,000	40,000
33	EFFLUENT	32,000	32,000	40,000	40,000	40,000
34	EFFLUENT	32,000	32,000	40,000	40,000	40,000
35	EFFLUENT	32,000	32,000	40,000	40,000	40,000
36	EFFLUENT	32,000	32,000	40,000	40,000	40,000
37	EFFLUENT	32,000	32,000	40,000	40,000	40,000
38	EFFLUENT	32,000	32,000	40,000	40,000	40,000
39	EFFLUENT	32,000	32,000	40,000	40,000	40,000
40	EFFLUENT	32,000	32,000	40,000	40,000	40,000
41	EFFLUENT	32,000	32,000	40,000	40,000	40,000
42	EFFLUENT	32,000	32,000	40,000	40,000	40,000
43	EFFLUENT	32,000	32,000	40,000	40,000	40,000
44	EFFLUENT	32,000	32,000	40,000	40,000	40,000
45	EFFLUENT	32,000	32,000	40,000	40,000	40,000
46	EFFLUENT	32,000	32,000	40,000	40,000	40,000
47	EFFLUENT	32,000	32,000	40,000	40,000	40,000
48	EFFLUENT	32,000	32,000	40,000	40,000	40,000
49	EFFLUENT	32,000	32,000	40,000	40,000	40,000
50	EFFLUENT	32,000	32,000	40,000	40,000	40,000
51	EFFLUENT	32,000	32,000	40,000	40,000	40,000
52	EFFLUENT	32,000	32,000	40,000	40,000	40,000
53	EFFLUENT	32,000	32,000	40,000	40,000	40,000
54	EFFLUENT	32,000	32,000	40,000	40,000	40,000
55	EFFLUENT	32,000	32,000	40,000	40,000	40,000
56	EFFLUENT	32,000	32,000	40,000	40,000	40,000
57	EFFLUENT	32,000	32,000	40,000	40,000	40,000
58	EFFLUENT	32,000	32,000	40,000	40,000	40,000
59	EFFLUENT	32,000	32,000	40,000	40,000	40,000
60	EFFLUENT	32,000	32,000	40,000	40,000	40,000
61	EFFLUENT	32,000	32,000	40,000	40,000	40,000
62	EFFLUENT	32,000	32,000	40,000	40,000	40,000

ITEM NO	DESCRIPTION	E-1	E-2	E-3	E-4	E-5
63	STATIONARY TUBE SHEET					
64	FLOATING TUBE SHEET					
65	CROSS BAFFLES					
66	LONGITUDINAL BAFFLES					
67	TUBE SUPPORTS					
68	SHELL COVERS					
69	FLOATING HEAD COVER					
70	CHANNEL COVER					
71	CROSS BAFFLES TYPE - SPACING					
72	LONGITUDINAL BAFFLE TYPE					
73	GASKETS					
74	WELDING FLANGES AND NOZZLES					
75	STUDS					
76	WEIGHT EXCHANGER					
77	WEIGHT BUNDLE					
78	CODE REQUIREMENTS					
79	SR&MC DRAWING NO					
80	REQUISITION NO					
81	PURCHASE ORDER NO					

UNIT PERFORMANCE	
1	FLUID CIRCULATED
2	LIQUID
3	VAPOR
4	QUANTITY
5	FIXED GASES
6	STEAM
7	TOTAL
8	FLUID VAPORIZED OR CONDENSED
9	GRAVITY LIQUID (ISSUING) AT °F
10	VISCOUSITY LIQUID (ISSUING) AT °F
11	SPEC HEAT LIQUID BTU/LB
12	LATENT HEAT VAPOR BTU/LB
13	OPER TEMP °F IN
14	OPER TEMP °F OUT
15	OPER PRESSURE PSIG
16	MINIMUM FOULING FACTOR
17	NUMBER OF PASSES
18	VELOCITY FT/SEC
19	PRESSURE DROP PSIG
20	HEAT EXCHANGED BTU/HR
21	WT CORRECTED WEIGHT
22	TRANSFER RATE SERVICE CLEAN
23	DESIGN PRESSURE PSIG
24	DESIGN TEMPERATURE °F
25	CONNECTION ALLOWANCE
26	CONNECTIONS IN
27	SIZE STD TYPE OUT
28	MAT'L
29	TUBES O.D.
30	LENGTH
31	SHELL I.D.
32	STATIONARY TUBE SHEET
33	FLOATING TUBE SHEET
34	CROSS BAFFLES
35	LONGITUDINAL BAFFLES
36	TUBE SUPPORTS
37	SHELL COVERS
38	FLOATING HEAD COVER
39	CHANNEL COVER
40	CROSS BAFFLES TYPE - SPACING
41	LONGITUDINAL BAFFLE TYPE
42	GASKETS
43	WELDING FLANGES AND NOZZLES
44	STUDS
45	WEIGHT EXCHANGER
46	WEIGHT BUNDLE
47	CODE REQUIREMENTS
48	SR&MC DRAWING NO
49	REQUISITION NO
50	PURCHASE ORDER NO

CONSTRUCTION

**NOTES**  
 SA - STRESS RELIEVED  
 SA - X RAYED

**REVISION**

NO	DATE	REVISION	BY

**Sun Refining and Marketing Company**

**HEAT EXCHANGER SCHEDULE**

PLANT AIR FORCE  
 HYDROLYSE/STAKE

LOCATION

BY	APPRO	DRAWING NO	REV SHEET

ITEM NO	DESCRIPTION	E-G	E-7	E-B-A	E-B-B	E-9
1	MANUFACTURER					
2	SIZE AND TYPE					
3	SURFACE/UNIT					
4	CONNECTED IN					
5	FLUID CIRCULATED					
6	QUANTITY					
7	FIRE GASES					
8	STEAM					
9	TOTAL					
10	FLUID VAPORIZED OR CONDENSED					
11	GRAVITY					
12	VISCOUSITY					
13	SPEC HEAT					
14	LATENT HEAT					
15	OPER TEMP					
16	OPER PRESSURE					
17	MINIMUM FOULING FACTOR					
18	NUMBER OF PASSES					
19	VELOCITY					
20	PRESSURE DROP					
21	HEAT EXCHANGED					
22	MTD CORRECTED					
23	TRANSFER RATE					
24	DESIGN PRESSURE					
25	TEST PRESSURE					
26	DESIGN TEMPERATURE					
27	CORROSION ALLOWANCE					
28	CONNECTIONS					
29	SIZE - STD TYPE					
30	MAT'L					
31	TUBES					
32	LENGTH					
33	SHELL I.D.					
34	PITCH					
35	STATIONARY TUBE SHEET					
36	FLOATING TUBE SHEET					
37	CROSS BAFFLES					
38	LONGITUDINAL BAFFLES					
39	TUBE SUPPORTS					
40	SHELL COVERS					
41	FLOATING HEAD COVER					
42	CHANNEL COVER					
43	CROSS BAFFLES TYPE					
44	LONGITUDINAL BAFFLE TYPE					
45	GASKETS					
46	WELDING FLANGES AND NOZZLES					
47	STUDS					
48	WEIGHT EXCHANGER					
49	WEIGHT BUNDLE					
50	CODE REQUIREMENTS					
51	SA & BC DRAWING NO					
52	REQUISITION NO					
53	PURCHASE ORDER NO					

ITEM NO	DESCRIPTION	E-G	E-7	E-B-A	E-B-B	E-9
34	STATIONARY TUBE SHEET					
35	FLOATING TUBE SHEET					
36	CROSS BAFFLES					
37	LONGITUDINAL BAFFLES					
38	TUBE SUPPORTS					
39	SHELL COVERS					
40	FLOATING HEAD COVER					
41	CHANNEL COVER					
42	CROSS BAFFLES TYPE					
43	LONGITUDINAL BAFFLE TYPE					
44	GASKETS					
45	WELDING FLANGES AND NOZZLES					
46	STUDS					
47	WEIGHT EXCHANGER					
48	WEIGHT BUNDLE					
49	CODE REQUIREMENTS					
50	SA & BC DRAWING NO					
51	REQUISITION NO					
52	PURCHASE ORDER NO					

ITEM NO	DESCRIPTION	UNIT	QTY	UNIT PRICE	TOTAL	REMARKS
1	AMINE STILL	TUBE	1	1700	1700	STEAM
2	AMINE STILL	TUBE	1	1700	1700	STEAM
3	AMINE STILL	TUBE	1	1700	1700	STEAM
4	AMINE STILL	TUBE	1	1700	1700	STEAM
5	AMINE STILL	TUBE	1	1700	1700	STEAM
6	AMINE STILL	TUBE	1	1700	1700	STEAM
7	AMINE STILL	TUBE	1	1700	1700	STEAM
8	AMINE STILL	TUBE	1	1700	1700	STEAM
9	AMINE STILL	TUBE	1	1700	1700	STEAM
10	AMINE STILL	TUBE	1	1700	1700	STEAM
11	AMINE STILL	TUBE	1	1700	1700	STEAM
12	AMINE STILL	TUBE	1	1700	1700	STEAM
13	AMINE STILL	TUBE	1	1700	1700	STEAM
14	AMINE STILL	TUBE	1	1700	1700	STEAM
15	AMINE STILL	TUBE	1	1700	1700	STEAM
16	AMINE STILL	TUBE	1	1700	1700	STEAM
17	AMINE STILL	TUBE	1	1700	1700	STEAM
18	AMINE STILL	TUBE	1	1700	1700	STEAM
19	AMINE STILL	TUBE	1	1700	1700	STEAM
20	AMINE STILL	TUBE	1	1700	1700	STEAM
21	AMINE STILL	TUBE	1	1700	1700	STEAM
22	AMINE STILL	TUBE	1	1700	1700	STEAM
23	AMINE STILL	TUBE	1	1700	1700	STEAM
24	AMINE STILL	TUBE	1	1700	1700	STEAM
25	AMINE STILL	TUBE	1	1700	1700	STEAM
26	AMINE STILL	TUBE	1	1700	1700	STEAM
27	AMINE STILL	TUBE	1	1700	1700	STEAM
28	AMINE STILL	TUBE	1	1700	1700	STEAM
29	AMINE STILL	TUBE	1	1700	1700	STEAM
30	AMINE STILL	TUBE	1	1700	1700	STEAM
31	AMINE STILL	TUBE	1	1700	1700	STEAM
32	AMINE STILL	TUBE	1	1700	1700	STEAM
33	AMINE STILL	TUBE	1	1700	1700	STEAM
34	AMINE STILL	TUBE	1	1700	1700	STEAM
35	AMINE STILL	TUBE	1	1700	1700	STEAM
36	AMINE STILL	TUBE	1	1700	1700	STEAM
37	AMINE STILL	TUBE	1	1700	1700	STEAM
38	AMINE STILL	TUBE	1	1700	1700	STEAM
39	AMINE STILL	TUBE	1	1700	1700	STEAM
40	AMINE STILL	TUBE	1	1700	1700	STEAM
41	AMINE STILL	TUBE	1	1700	1700	STEAM
42	AMINE STILL	TUBE	1	1700	1700	STEAM
43	AMINE STILL	TUBE	1	1700	1700	STEAM
44	AMINE STILL	TUBE	1	1700	1700	STEAM
45	AMINE STILL	TUBE	1	1700	1700	STEAM
46	AMINE STILL	TUBE	1	1700	1700	STEAM
47	AMINE STILL	TUBE	1	1700	1700	STEAM
48	AMINE STILL	TUBE	1	1700	1700	STEAM
49	AMINE STILL	TUBE	1	1700	1700	STEAM
50	AMINE STILL	TUBE	1	1700	1700	STEAM
51	AMINE STILL	TUBE	1	1700	1700	STEAM
52	AMINE STILL	TUBE	1	1700	1700	STEAM
53	AMINE STILL	TUBE	1	1700	1700	STEAM
54	AMINE STILL	TUBE	1	1700	1700	STEAM
55	AMINE STILL	TUBE	1	1700	1700	STEAM
56	AMINE STILL	TUBE	1	1700	1700	STEAM
57	AMINE STILL	TUBE	1	1700	1700	STEAM
58	AMINE STILL	TUBE	1	1700	1700	STEAM
59	AMINE STILL	TUBE	1	1700	1700	STEAM
60	AMINE STILL	TUBE	1	1700	1700	STEAM
61	AMINE STILL	TUBE	1	1700	1700	STEAM
62	AMINE STILL	TUBE	1	1700	1700	STEAM

NOTES  
 SH - STRESS RELIEVED  
 SR - STAYED

Sun Refining and Marketing Company

HEAT EXCHANGER SCHEDULE

PLANT AIR FORCE HYDROVISBREAKER  
 PLANT NO. UNIT NO.

LOCATION

BY	APPRO	DRAWING NO.	REV.	SHEET
				3
DATE				

ITEM NO	SERVICE	MANUFACTURER	NO SHELLS	SIZE AND TYPE	SURFACE / UNIT	SHELL	TUBE	E-15 VAC TRK PA BEN DTC	E-16 VAC BTHS AIR COOLER	E-10A AMINE LOOLER	E-11A AMINE RECLAIMER	NOTE
1		E-14 VAC TRK PA FOR STM MAN		AKL 502								
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Sun Refining and Marketing Company

HEAT EXCHANGER SCHEDULE

PLANT NO. 40156  
UNIT NO. 4

LOCATION: BY APPLD DRAWING NO. REV SHEET  
DATE: 4/25/67



NOTES

HEATER SECTION	HEATER TYPE	HEATER DESIGN	HEATER DESIGN
<p>HEATER SECTION</p> <p>SERVICE</p> <p>HEAT ABSORPTION MM BTU/HR</p> <p>HEAT FLOW MM BTU/HR</p> <p>HEAT FLOW PER PASS (1.48/HR)</p> <p>HEAT FLOW PER TUBE (1.48/HR)</p> <p>HEAT FLOW PER TUBE AT 60°F</p> <p>HEAT FLOW PER TUBE AT 200°F</p> <p>HEAT FLOW PER TUBE AT 300°F</p> <p>HEAT FLOW PER TUBE AT 400°F</p> <p>HEAT FLOW PER TUBE AT 500°F</p> <p>HEAT FLOW PER TUBE AT 600°F</p> <p>HEAT FLOW PER TUBE AT 700°F</p> <p>HEAT FLOW PER TUBE AT 800°F</p> <p>HEAT FLOW PER TUBE AT 900°F</p> <p>HEAT FLOW PER TUBE AT 1000°F</p>	<p>H-1 RECYCLE GAS HEATER</p> <p>H-2 ATM. TOWER FEED HEATER</p>	<p>RADIANT CONVECTION PROCESS</p> <p>RADIANT CONVECTION PROCESS</p> <p>RADIANT CONVECTION PROCESS</p>	<p>86.9/BL</p> <p>86.9/BL</p> <p>86.9/BL</p>
<p>HEATER SECTION</p> <p>SERVICE</p> <p>HEAT ABSORPTION MM BTU/HR</p> <p>HEAT FLOW MM BTU/HR</p> <p>HEAT FLOW PER PASS (1.48/HR)</p> <p>HEAT FLOW PER TUBE (1.48/HR)</p> <p>HEAT FLOW PER TUBE AT 60°F</p> <p>HEAT FLOW PER TUBE AT 200°F</p> <p>HEAT FLOW PER TUBE AT 300°F</p> <p>HEAT FLOW PER TUBE AT 400°F</p> <p>HEAT FLOW PER TUBE AT 500°F</p> <p>HEAT FLOW PER TUBE AT 600°F</p> <p>HEAT FLOW PER TUBE AT 700°F</p> <p>HEAT FLOW PER TUBE AT 800°F</p> <p>HEAT FLOW PER TUBE AT 900°F</p> <p>HEAT FLOW PER TUBE AT 1000°F</p>	<p>H-1 RECYCLE GAS HEATER</p> <p>H-2 ATM. TOWER FEED HEATER</p>	<p>RADIANT CONVECTION PROCESS</p> <p>RADIANT CONVECTION PROCESS</p> <p>RADIANT CONVECTION PROCESS</p>	<p>86.9/BL</p> <p>86.9/BL</p> <p>86.9/BL</p>
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NO. DATE REVISION BY

**Sun Refining and Marketing Company**

**FIRING HEATER DESIGN**

PLANT AIR FORCE HYDROVISBREAKER

PLANT NO. UNIT NO.

LOCATION

BY APP. DATE

LCR

4/25/57

DRAWING NO. REV. SHEET

5

NOTES

SERVICE		PROCESS DESIGN CONDITIONS		MATERIAL		MANUFACTURE		HEATER DESIGN	
VAC TWR FEED HEATER		52.2, 160.7		RADIANT CONVECTION		36.38		36.38	
TOTAL DUTY OF HEATER MM BTU PER HR ABSORBED/FIRED		36.38		36.38		36.38		36.38	
HEATER SECTION		36.38		36.38		36.38		36.38	
SERVICE		36.38		36.38		36.38		36.38	
HEAT ABSORPTION MM BTU/HR		36.38		36.38		36.38		36.38	
FLOW RATE LB/HR		36.38		36.38		36.38		36.38	
FLOW RATE GPM TO COIL		36.38		36.38		36.38		36.38	
PRESSURE DROP PSI (ALLOWABLE)		36.38		36.38		36.38		36.38	
PRESSURE DROP PSI (CALCULATED)		36.38		36.38		36.38		36.38	
AVERAGE FLUX DENSITY BTU/HR SQ FT		36.38		36.38		36.38		36.38	
FOULING FACTOR		36.38		36.38		36.38		36.38	
INLET CONDITIONS		36.38		36.38		36.38		36.38	
TEMPERATURE DEG F		36.38		36.38		36.38		36.38	
PRESSURE (PSIA) (PSIG)		36.38		36.38		36.38		36.38	
LIQUID FLOW LB/HR		36.38		36.38		36.38		36.38	
VAPOR FLOW LB/HR		36.38		36.38		36.38		36.38	
LIQUID (DEG API) (SP GR AT 60 F)		36.38		36.38		36.38		36.38	
VAPOR MOLECULAR WEIGHT		36.38		36.38		36.38		36.38	
LIQUID VISCOSITY CP		36.38		36.38		36.38		36.38	
OUTLET CONDITIONS		36.38		36.38		36.38		36.38	
TEMPERATURE DEG F		36.38		36.38		36.38		36.38	
PRESSURE (PSIA) (PSIG)		36.38		36.38		36.38		36.38	
LIQUID FLOW LB/HR		36.38		36.38		36.38		36.38	
VAPOR FLOW LB/HR		36.38		36.38		36.38		36.38	
LIQUID (DEG API) (SP GR AT 60 F)		36.38		36.38		36.38		36.38	
VAPOR MOLECULAR WEIGHT		36.38		36.38		36.38		36.38	
LIQUID VISCOSITY CP		36.38		36.38		36.38		36.38	
FUEL CHARACTERISTICS		36.38		36.38		36.38		36.38	
TYPE OF FUEL		36.38		36.38		36.38		36.38	
HEATING VALUE LHV		36.38		36.38		36.38		36.38	
SPECIFIC GRAVITY		36.38		36.38		36.38		36.38	
SULFUR/NITROGEN WEIGHT		36.38		36.38		36.38		36.38	
COIL DESIGN		36.38		36.38		36.38		36.38	
HEATER SECTION		36.38		36.38		36.38		36.38	
DESIGN PRESSURE PSIG		36.38		36.38		36.38		36.38	
DESIGN FLUID TEMPERATURE DEG F		36.38		36.38		36.38		36.38	
CORROSION ALLOWANCE TUBES		36.38		36.38		36.38		36.38	
FITTINGS		36.38		36.38		36.38		36.38	
HYDROSTATIC TEST PRESSURE PSIG		36.38		36.38		36.38		36.38	
NUMBER OF PASSES		36.38		36.38		36.38		36.38	
OVERALL TUBE LENGTH FT		36.38		36.38		36.38		36.38	
EFFECTIVE TUBE LENGTH FT		36.38		36.38		36.38		36.38	
BARE TUBES NUMBER		36.38		36.38		36.38		36.38	
TOTAL EXPOSED SURFACE SQ FT		36.38		36.38		36.38		36.38	
EXTENDED SURFACE TUBES NUMBER		36.38		36.38		36.38		36.38	
TOTAL EXPOSED SURFACE SQ FT		36.38		36.38		36.38		36.38	
TUBE SPACING CENTER TO CENTER IN (STAGGERED) (IN LINE)		36.38		36.38		36.38		36.38	
TUBE CENTER TO FURNACE WALL IN MIN		36.38		36.38		36.38		36.38	
HEAT TREATMENT		36.38		36.38		36.38		36.38	
WELD INSPECTION REQUIREMENTS - X RAY OR OTHER		36.38		36.38		36.38		36.38	
TUBES		36.38		36.38		36.38		36.38	
GENERAL TUBES SPECIFICATION		36.38		36.38		36.38		36.38	
TUBE MATERIAL, ASTM SPECIFICATION AND GRADE		36.38		36.38		36.38		36.38	
TUBE SIZE, DIMENSIONS, IN		36.38		36.38		36.38		36.38	
WALL THICKNESS, MINIMUM, AVAILABLE IN		36.38		36.38		36.38		36.38	

NO. DATE REVISION BY

Sun Refining and Marketing Company  
FIRED HEATER DESIGN

PLANT AIR FORCE  
PLANT NO HYDRO VISBREAKER  
UNIT NO

LOCATION

BY (APP'D) DRAWING NO REV. SHEET  
DATE 4/15/17 6



SERVICE	ITEM NO.	TITLE	DESIGN				DIMENSIONAL				PURCHG DATA	SPEC	MATERIAL	WEIGHT	PRICE	DATE	BY	DRAWING NO.	REV SHEET		
			V-1	V-2	V-3	V-4	V-5	V-6	V-7	V-8										V-9	
	V-1	HOT HP SEPR	275 650 600 600 400	200 650 200	1680 650 4214.4 190	200 650 200	3670 650 195	75 650 125	204 650 178	AMINE ATHTWK FL DRUM	3.5 3.5 3.5	12.000	1750	15000	12000	12000	11/15/77	CS	8		
	V-2	HOT LP SEPR	200 650 300	220 650 1735	1735 300	200 650 200	2670 650 195	75 650 125	204 650 178	AMINE ATHTWK FL DRUM	3.5 3.5 3.5	12.000	15000	12000	12000	11/15/77	CS	8			
	V-3	COOL HP SEPR	1680 650 4214.4 190	200 650 200	3670 650 195	75 650 125	204 650 178	AMINE ATHTWK FL DRUM	3.5 3.5 3.5	12.000	1750	15000	12000	12000	11/15/77	CS	8				
	V-4	COLD LP SEPR	200 650 200	220 650 1735	300	200 650 200	2670 650 195	75 650 125	204 650 178	AMINE ATHTWK FL DRUM	3.5 3.5 3.5	12.000	15000	12000	12000	11/15/77	CS	8			
	V-5	COLD HP SEPR	1680 650 4214.4 190	200 650 200	3670 650 195	75 650 125	204 650 178	AMINE ATHTWK FL DRUM	3.5 3.5 3.5	12.000	1750	15000	12000	12000	11/15/77	CS	8				
	V-6	AMINE ATHTWK FL DRUM	3.5 3.5 3.5	12.000	1750	15000	12000	12000	11/15/77	CS	8										
	V-7	AMINE ATHTWK FL DRUM	3.5 3.5 3.5	12.000	1750	15000	12000	12000	11/15/77	CS	8										
	V-8	AMINE ATHTWK FL DRUM	3.5 3.5 3.5	12.000	1750	15000	12000	12000	11/15/77	CS	8										
	V-9	STEAM DRUM	3.5 3.5 3.5	12.000	1750	15000	12000	12000	11/15/77	CS	8										
	NOTES																				
Sun Refining and Marketing Company TOWER & VESSEL SCHEDULE																					
PLANT AIR FORCE HYDROBREAKER																			PLANT NO	8	
LOCATION																			BY APPD	DRAWING NO.	REV SHEET
																			DATE	11/15/77	

PUMP ITEM NO	H.L.A	P.S. 3A	P.S. 4A	P.S. 5A	P.6, 6A	P.7, 7A	P.8, 8A	P.9, 9A	P.10, 10A	NOTES
1	REACTOR FEED	REACTOR FEED	REACTOR CIRCULATION PUMPS	ATM TWR P. A	ATM TWR P. A	LEWY AMINE	AMINE REFLUX	ATM TWR REFLUX	ATM TWR BOTTOMS	
2	RESID. WASH	850								
3	514	70								
4	894	20								
5	45P	20P								
6	144	100								
7	2700	30								
8	2737	88								
9	1/2	1/2								
10										
11	3495	3								
12	3400	5								
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Sun Refining and Marketing Company

CENTRIFUGAL PUMP SCHEDULE

PLANT ATM FORCE HYDROLYSIS BREAKER UNIT NO

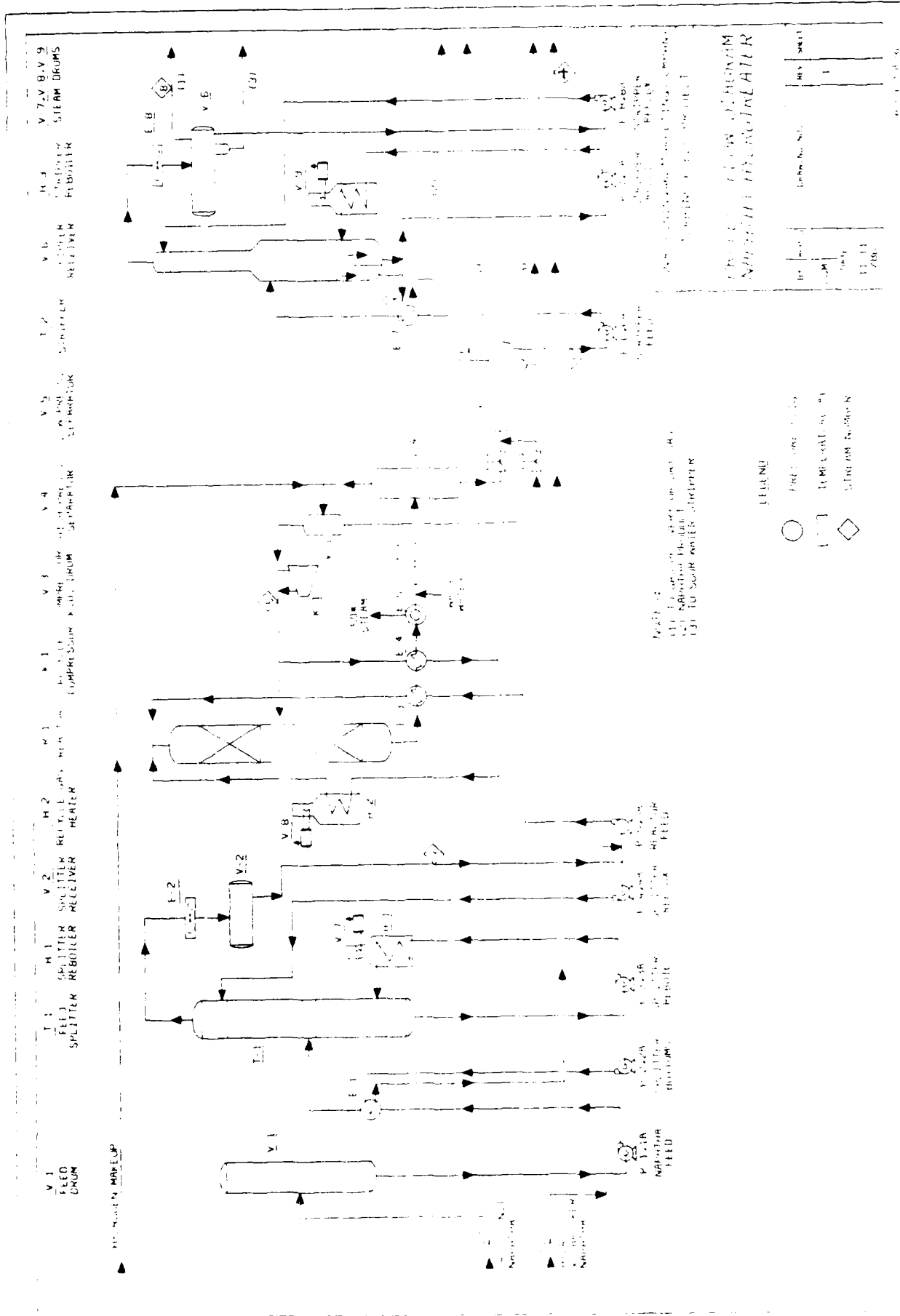
BY	APPRO	DRAWING NO	REV SHEET
LCV			9
DATE			
4/15/87			



**PROCESS DESIGN SPECIFICATIONS**

**for the**

**NAPHTHA HYDROTREATING UNIT**





NAPHTHA HYDROTREATING UNIT  
MATERIAL BALANCE AND STREAM PROPERTIES

Stream Number	1	2	3	4	5	6	7
Stream Label	Naphtha from Crude Unit	Naphtha from Hydrovisbreaker	Feed Splitter Overhead	Feed Splitter Bottoms	Recycle Gas	Low-Pressure Separator Gas	Low-Pressure Separator Liquid
Stream Conditions							
Temperature, °F	332	100	259	595	126	116	116
Physical state	Liquid	Liquid	Liquid	Liquid	Vapor	Vapor	Liquid
API Gravity	27.8	49.9	33.7	24.1	-	-	42.9
Sp.Gr. @ 60°F	0.78	0.77	0.78	0.70	-	-	0.79
Sp.Gr. @ Temp. Pressure, psia	70	35	15	23	(4.35 mol wt) 1325	(13.5 mol wt) 165	165
BBLs/DAY @ 60°F	10,542	11,875	4,267	6,276	-	-	15,966
M lb/hr	136.44	135.00	53.27	83.17	37.36	1.92	188.76
GPM @ 60°F							
Vis., cSt @ Temp.	0.3	0.6	0.3	0.2	0.01	0.01	0.5

Stream Number	8	9
Stream Label	Stripper Overhead Gas	Stripper Bottoms
Stream Conditions		
Temperature, °F	100	506
Physical state	Vapor	Liquid
API Gravity	-	42.0
Sp.Gr. @ 60°F	-	0.61
Sp.Gr. @ Temp. Pressure, psia	(29.9 mol wt) 165	175
BBLs/DAY @ 60°F	-	15,790
M lb/hr	1.26	187.50
GPM @ 60°F	-	-
Vis., cSt @ Temp	0.01	0.02

## NAPHTHA HYDROTREATING UNIT DESIGN BASIS

### Naphtha Feed Rate

The unit is designed to hydrotreat 16,327 BPSD of 47.3°API San Ardo naphtha with an ASTM end-point of 490°F. The sulfur and nitrogen concentrations in the feed are 3820 and 70 wppm, respectively. Both the sulfur and nitrogen content of the hydrotreated product will be less than 10 ppm. The feed naphtha is a blend of 4451 BPSD of crude unit naphtha and 11,875 BPSD of hydrovisbreaker naphtha. The unit will be capable of maintaining a 94% on-stream factor.

### Plant Processing Steps

The naphtha from the crude unit is fractionated in the Naphtha Hydrotreater feed splitter tower to obtain an overhead naphtha with a 490°F end-point. The feed splitter bottoms (490-650°F cut) is fed to the Distillate Hydrotreater. The overhead naphtha is combined with the 490°F end-point naphtha from the Hydrovisbreaker to form the total liquid feed to the Naphtha Hydrotreater. It is heated by exchange with the Naphtha Hydrotreater Reactor effluent to 690°F and then fed to the top of the reactor.

Recirculated hydrogen plus make-up hydrogen heated to 650-700°F by heat exchange and fired heat is also fed to the top of the reactor. The reactor operates at 650 to 750°F and 1565 psig over its operating cycle of about two years.

The reactor effluent is cooled and water-washed to prevent fouling and corrosion of the effluent aircooler. The unreacted hydrogen is separated and recycled to the reactor.

The hydrotreated naphtha product is stripped (debutanized) at the unit, and this permits it to be stored at atmospheric pressure in intermediate tankage, if needed. Thus a short emergency outage of the refinery Gas Plant

*fractionation will not force a shutdown of the Naphtha Hydrotreater. The hydrotreated naphtha can bypass the Absorber-Stripper and Debutanizer Towers at the Gas Plant and, instead, flow directly to the Main Fractionator at the Distillate Hydrocracker Unit.*

NAPHTHA HYDROTREATER UNIT

Utilities and Chemical Requirements

Saturated Steam Produced, lb/hr

Steam Generators:	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>
V-7 AT Heater H-1	-	8,750	-	8,750
V-8 at Heater H-2	-	6,200	-	6,200
V-9 at Heater H-3	-	4,050	-	4,040
E-5	-	-	10,000	-
Totals	-	19,000	10,000	29,000

Steam Used, lb/hr

	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>	<u>Condensate Recovered</u>
Heaters H1,H2,H3	1,000	-	-	1,000	-
Misc. heating	-	5,000	5,000	10,000	-
Total used	1,000	5,000	5,000	11,000	-
Net export stream	-1,000	14,000	5,000	18,000	

Boiler Feed Water

For steam produced	29,000 lb/hr
For 10% blowdown & wash water	9,000
Total gross BFW needed	<u>38,000 lb/hr</u>
Condensate recovered (reused as BFW)	-
Net BFW needed	<u>38,000 lb/hr=76 gpm</u>

<u>Cooling Water Circulated</u>	<u>Supply</u>	<u>Return</u>	<u>Duty</u>
None required			

Heater Fuel Fired

Fuel: Vacuum Residuum Product from the Hydrovisbreaker Unit.

Heater H-1	402 MMBTU/Hr
Heater H-2	28.3
Heater H-3	18.8
Total	<u>87.3 MMBTU/Hr=14.5 FOE Bbl/Hr</u> =5,080 lb/hr residuum fuel

NAPHTHA HYDROTREATER UNIT

Utilities and Chemical Requirements  
(continued)

Electrical Power

	<u>Brake Horsepower Operating</u>	<u>Brake Horsepower Connected</u>
P-1,1A Naphtha splitter feed pumps	30	60
P-2,2A Splitter bottoms pumps	30	60
P-3,3A Splitter reboiler pumps	70	140
P-4,4A Splitter reflux pumps	20	40
P-5,5A Reactor feed pumps	700	1,400
P-6,6A Product stripper feed pumps	75	150
P-7,7A Stripper reboiler pumps	75	150
P-8,8A Stripper reflux pumps	10	20
E-2,E-6, E-8 Fans	260	260
K-1 Recycle Gas Compressor	800	800
Total Brake Horsepower:	2,070	3,080 BHP
Kilowatts:	1,543 KW	2,297 KW

Air Requirements

Dry Instrument Air	50 psig	150 SCFM
Plant Air	120 psig	

Hydrotreating Catalyst

See table of Refinery Operating Requirements, Catalyst and Chemicals in the overall refinery description.

Type: Nickel-molybdenum-on-alumina  
 Initial fill: 59,784 lb at 47 lb/ft<sup>3</sup>  
 Life: 4 years total, includes one regeneration

# NAPHTHA HYDROTREATER UNIT

## Major Equipment List

### Heat Exchangers

E-1	Feed Splitter Tower Feed - Bottoms Exchanger
E-2	Feed Splitter Overhead Condenser (aircooler)
E-3	Reactor Effluent - Reactor Naphtha Feed Exchanger
E-4	Reactor Effluent - Reactor Gas Feed Exchanger
E-5	Reactor Effluent - 50 psig Steam Generator
E-6	Reactor Effluent Cooler (aircooler)
E-7	Stripper Feed - Bottoms Exchanger

### Fired Heaters

H-1	Feed Splitter Reboiler
H-2	Recycle Gas Heater
H-3	Stripper Reboiler

### Towers

T-1	Feed Splitter
T-2	Product Stripper (product debutanization)

### Reactor

R-1	Naphtha Hydrotreater Reactor
-----	------------------------------

### Compressor

K-1	Recycle Compressor
-----	--------------------

### Pumps

P-1,1A	Straight-run Naphtha Feed
P-2,2A	Feed Splitter Bottoms
P-3,3A	Feed Splitter Reboiler
P-4,4A	Feed Splitter Reflux
P-5,5A	Reactor Feed
P-6,6A	Product Stripper Feed
P-7,7A	Product Stripper Reboiler
P-8,8A	Product Stripper Reflux

### Vessels

V-1	Naphtha Hydrotreater Feed Drum (for straight-run naphtha)
V-2	Feed Splitter Overhead Accumulator Drum
V-3	Recycle Compressor Knockout Drum
V-4	High Pressure Separator
V-5	Low Pressure Separator
V-6	Stripper Overhead Accumulator Drum
V-7	Steam Drum for H-1 Heater
V-8	Steam Drum for H-2 Heater
V-9	Steam Drum for H-3 Heater

AD-A190 120

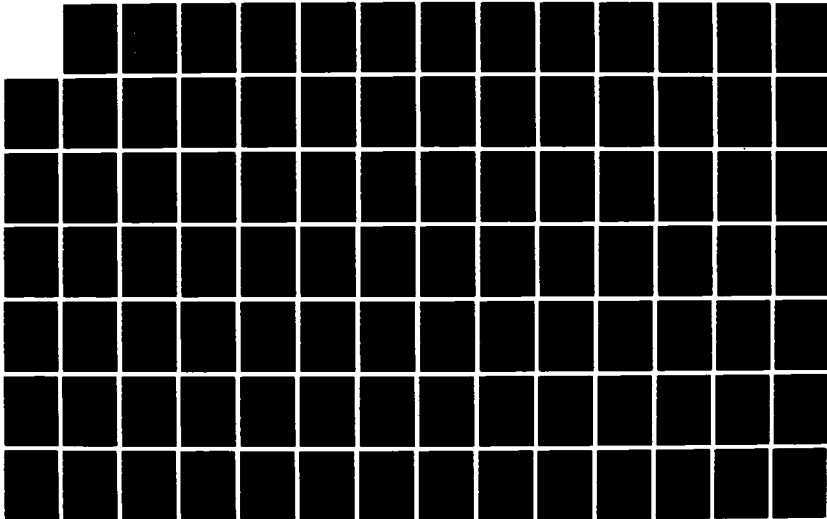
TURBINE FUELS FROM TAR SANDS BITUMEN AND HEAVY OIL  
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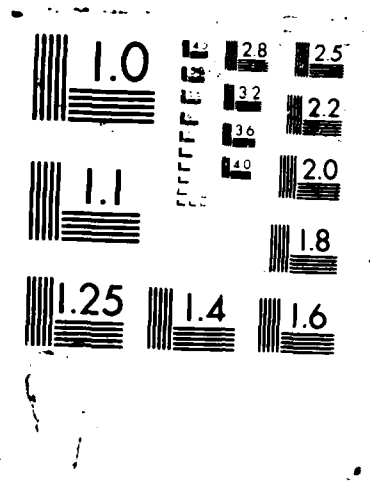
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ITEM NO.	DESCRIPTION	UNIT	QTY	UNIT PRICE	TOTAL	REMARKS
1	STRIPPER FLO. BOTTLES EXH.					
2	STRIPPER FLO. BOTTLES EXH.					
3	STRIPPER FLO. BOTTLES EXH.					
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42	STRIPPER FLO. BOTTLES EXH.					
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79	STRIPPER FLO. BOTTLES EXH.					
80	STRIPPER FLO. BOTTLES EXH.					
81	STRIPPER FLO. BOTTLES EXH.					
82	STRIPPER FLO. BOTTLES EXH.					

NOTES  
BR - STRESS RELIEVED  
2R - 2 RAYED

NO. DATE REVISION BY  
Sun Refining and Marketing Company  
HEAT EXCHANGER SCHEDULE  
PLANT AIR FORCE - MAPHYA  
HYDROTREATERS  
PLANT NO. UNIT NO.  
LOCATION  
BY APPD DRAWING NO. REV SHEET  
DATE 4/23/67  
2

HEATER SECTION		PROCESS DESIGN CONDITIONS	
HEATER TYPE	MANUFACTURER	SERVICE	HEAT ABSORPTION MM BTU/HR
H-1	SPLITTER REBOILER	H-2	RECYCLE GAS HEATER
34.1/482	24.44/28.34		
<b>HEATER DESIGN CONDITIONS</b> SERVICE: RADIANZ CONVECTION PROCESS: NAPIHTHA FLOW RATE (LBM/HR): 318.46 PRESSURE DROP (PSI) (ALLOWABLE): 3.0 PRESSURE DROP (PSI) (CALCULATED): 3.0 AVERAGE FLUID DENSITY (BTU/INCH SQ FT): 31.58 FOULING FACTOR: 1.0 <b>INLET CONDITIONS</b> TEMPERATURE (DEG F): 580 PRESSURE (PSIA) (PSIG): 60 LIQUID FLOW (LBM/HR): 318.46 VAPOR FLOW (LBM/HR): 0.0 LIQUID (DEG-FAH) (SP GR AT 60 F): 0.0 VAPOR MOLECULAR WEIGHT: 0.0 LIQUID VISCOSITY (CP): 0.0 <b>OUTLET CONDITIONS</b> TEMPERATURE (DEG F): 595 PRESSURE (PSIA) (PSIG): 70 LIQUID FLOW (LBM/HR): 318.46 VAPOR FLOW (LBM/HR): 0.0 LIQUID (DEG-FAH) (SP GR AT 60 F): 0.0 VAPOR MOLECULAR WEIGHT: 0.0 LIQUID VISCOSITY (CP): 0.0 <b>FUEL CHARACTERISTICS</b> TYPE OF FUEL: LOAFI HEATING VALUE (LHV): 2543 SULFUR/NITROGEN WEIGHT %: 0.0 <b>COIL DESIGN</b> HEATER SECTION: RADIANZ CONVECTION DESIGN PRESSURE (PSIG): 50 DESIGN FLUID TEMPERATURE (DEG F): 650 COMPOSITION: ALLOWANCE TUBES FITTINGS: 1/8" HYDROSTATIC TEST PRESSURE (PSIG): 200 NUMBER OF PASSES: 650 OVERALL TUBE LENGTH (FT): 1/8" EFFECTIVE TUBE LENGTH (FT): 1/8" BARE TUBES NUMBER: 200 TOTAL EXPOSED SURFACE SQ FT: 650 EXTENDED SURFACE TUBES NUMBER: 650 TUBE SPACING (CENTER TO CENTER IN (STAGGERED) IN LINE): 1/8" TUBE CENTER TO URNANCE WALL IN MIN: 1/8" HEAT TREATMENT: 1/8" WELD INSPECTION REQUIREMENTS: X-RAY OR OTHER: 1/8" <b>TUBES</b> TYPE: H 5CR WALL THICKNESS (MINIMUM) (AVERAGE) IN: 1/8" TUBE MATERIAL (ASTM SPECIFICATION AND GRADE): 1/8" TUBE END FINISH: 1/8" WALL THICKNESS (MINIMUM) (AVERAGE) IN: 1/8"			
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NOTES

NO. DATE REVISION BY

8/19/98 H-2

Sun Refining and Marketing Company

FIRED HEATER DESIGN

PLANT A/R FORCE - NAPIHTHA

PLANT NO. HYDROTREATER

LOCATION UNIT NO.

BY JAPP DRAWING NO. REV SHEET

LOC DATE 4/15/87 3

NOTES

SERVICE: H-3 STRIPPER REBOILER  
 TYPE: H-3 STRIPPER REBOILER  
 MANUFACTURER: N.A.P.M.T.A.  
 TOTAL DUTY OF HEATER MM BTU PER HR ABSORBED: 16,217,845

PROCESS DESIGN CONDITIONS

HEATER SECTION: RADIANT CONVECTION  
 SERVICE: PROCESS  
 HEAT ABSORPTION MM BTU/HR: 16,217,845  
 FLOW RATE LB/HR: 40,000 BFW  
 FLOW RATE BPD: 1578  
 PRESSURE ORGP PSI ALLOWABLE: 150  
 PRESSURE ORGP PSI CALCULATED: 152  
 ALLOWABLE DENSITY BTU/HR/FT<sup>2</sup>: 20,245  
 MANUFACTURER: N.A.P.M.T.A.  
 SPECIFICATIONS: 40,000 BFW  
 TEMPERATURE DEG F: 260  
 PRESSURE ORGP PSI: 152  
 WEIGHT FLOW LB/HR: 40,000  
 VOLUME FLOW GPM: 1578  
 SPECIFIC GRAVITY: 1.0  
 HEATER TYPE: H-3 STRIPPER REBOILER  
 HEATER MATERIAL: CS

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 HEATER TYPE: H-3 STRIPPER REBOILER  
 HEATER MATERIAL: CS

DATE: 8/1/1964  
 BY: M.S.  
 REVISION: 267  
 DIVISION: M.S.  
 COMPANY: Sun Refining and Marketing Company  
 DESIGN: FIRED HEATER DESIGN  
 PLANT NO: 10456 NAPTMA  
 UNIT NO: HYDROTREATER  
 LOCATION: NAPTMA  
 BY APPD: M.S.  
 DRAWING NO: 4  
 REV SHEET: 4  
 DATE: 4/22/67

ITEM NO.	TITLE	7-1	7-2	R-1	V-1	V-2	V-3	V-4	V-5	V-6	V-7
7-1	FEED SPLITTER										
7-2	STRIPPER										
R-1	REACTION										
V-1	FEED DRUM										
V-2	SPLITTER RECEIVER										
V-3	DOWNCOMER DRUM										
V-4	HI PRESSURE SEPARATOR										
V-5	LOW PRESSURE SEPARATOR										
V-6	STRIPPER RECEIVER										
V-7	STEAM DRUM										
ASSEMBLY											
DETAILS											
TRAYS											
PLATEWORK											
CODE											
PRESSURE											
TEMPERATURE											
CORROSION ALLOWANCE											
WIND LOADING											
HYDROSTATIC TEST PRESSURE											
HAMMER TEST PRESSURE											
STRESS RELIEVED											
RADIOGRAPHED											
VERTICAL OR HORIZONTAL											
REGULATED THICKNESS											
LENGTH BEAM TO BEAM											
LENGTH BASE SECTION											
HEIGHT OF SKIRT											
5 TO BOTTOM OF SUPPORTS											
LOCAL STRESS INCREASE											
HEAD THICKNESS											
SKIRT THICKNESS											
NO REGO											
SIZING											
TYPE											
FACE											
LOCATION											
NO TRAYS											
BEER & SPACING											
TYPE OF TRAY											
NO OF CAPS/TRAY B. TYPE											
SIZE OF CAPS											
SIZE OF RISERS											
TYPE OF DOWNCOMER											
NO TRAYS BEG B. SPACING											
TYPE OF TRAY											
TRAY B.											
SHELL B. HEADS											
SKIRT											
STRUCTURAL MEMBERS											
REQUISITION NO.											
PURCHASE ORDER											
PURCHASED FROM											
FABRICATION S/O											
SHIPPING WEIGHT											
EX. COST											

**NOTES**

- (1) SHELL 4.25 OF 24 CR. 1 INCH PLUS 3/16 OVERLAY CR. OVERLAY OVERLAY 1/8 CORROSION ALLOWANCE EX. OVERLAY 1/8 OVER 100 LBS HEADS 2.07 1/2 DIVERSITY MEMISPHRE/CAL
- (2) HEADS HULLISH/NO. CAL

REVISED 8/1/80 V. 2 V. 5 V. 6  
 1-1-1 L  
 2-13-80 HP VESSEL WTB 460

NO.	DATE	REVISION	BY

**Sun Refining and Marketing Company**

**TOWER & VESSEL SCHEDULE**

PLANT AIR FORCE - NAPHTHA HYDROTREATER

PLANT NO. UNIT NO.

LOCATION

BY APPD. DRAWING NO. REV. SHEET

DATE 4/15/81

5

ITEM NO.	TITLE	V-8	V-9	NOTES
DESIGN	STEAM DRUM	STEAM DRUM	STEAM DRUM	
SERVICE	CODE	200	200	
	TEMPERATURE OF	650	650	
	COMBUSTION ALLOWANCE	.115	.205	
	WIND LOADING			
	HYDROSTATIC TEST PRESSURE	300	300	
	HAMMER TEST PRESSURE			
	STRESS RECEIVED			
	RADIOGRAPHED	N.C.	N.C.	
	VERTICAL OR HORIZONTAL	SHRT	SHRT	
	INSULATION THICKNESS	H	H	
	O.D.	1' 2 1/2"	1' 2 1/2"	
	LENGTH FROM BEAM TO SEAM	12'	8'	
	LENGTH BASE SECTION			
	HEIGHT OF SHIRT			
	6 TO BOTTOM OF SUPPORTS			
TOTAL SHELL THICKNESS	4.50	4.50		
HEAD THICKNESS	4.50	4.50		
SHIRT THICKNESS				
NO. TRAYS	1	1		
SIZE	24"	24"		
SERIES & FACING				
TYPE				
THICKNESS				
LOCATION				
NO. TRAYS REQ'D & SPACING				
TYPE OF TRAY				
NO. OF CAPS / TRAY & TYPE				
SIZE OF CAPS				
SIZE OF RISERS				
TYPE OF DOWNCOMER				
NO. TRAYS REQ'D & SPACING				
TYPE OF TRAY				
TREY A				
TREY B				
TREY C				
TREY D				
TREY E				
TREY F				
REQUISITION NO.				
PURCHASE ORDER				
PURCHASE FROM				
FABRICATION S/O				
SHIPPING WEIGHT	4500#	2900#		
TEST WEIGHT				
SHELL & HEADS				
SHIRT	CS	CS		
STRUCTURAL MEMBERS				
BY APPD				
DRAWING NO				
REV SHEET			6	
BY JRS				
DATE			4/25/67	

PLANT NO. 400  
LOCATION  
BY APPD  
DRAWING NO  
REV SHEET

Sun Refining and Marketing Company  
TOWER & VESSEL SCHEDULE

PLANT AIR FORCE - NAPHTHA  
HYDRO REATER

PLANT NO. UNIT NO.  
LOCATION  
BY APPD  
DRAWING NO  
REV SHEET  
DATE  
4/25/67  
6

PUMP ITEM NO	SERVICE	P1,1A		P2,2A		P3,3A		P4,4A		P5,5A		P6,6A		P7,7A		P8,8A		K-1	NO. DATE	REVISION	BY	
		NAPHTHA FEED	SPLITTING BOTTOMS	SPLITTING REBOIL	SPLITTING REFLUX	REALTOR'S FEED	STRIPPERS FEED	STRIPPERS REBOIL	STRIPPERS REFLUX	STRIPPERS REFLUX	STRIPPERS REFLUX	STRIPPERS REFLUX	STRIPPERS REFLUX	STRIPPERS REFLUX	STRIPPERS REFLUX							
1	LIQUID																					
2	PUMPING TEMPERATURE °F	332	596	574	259	150	116	445	119	109.2												
3	SPECIFIC GRAVITY AT P.T.	1.77	1.69	1.69	1.71	1.76	1.76	1.58	1.57	1.44												
4	VISC. (CENTISTOGES) AT P.T.	1.5	1.4	1.4	1.5	1.5	1.5	1.4	1.7	1.38												
5	SPM AT P.T.	587	248	1000	248	541	539	886	105	892												
6	SUCTION PRESSURE (PSIG)	0	8	8	8	8	8	165	150	175												
7	DISCHARGE PRESSURE (PSIG)	70	110	70	68	1370	1370	245	211	210												
8	DIFFERENTIAL HEAD (FT)	210	391	208	204	187	187	317	241	240												
9	NPSH AVAILABLE	12	12	12	12	12	12	12	12	12												
10	PPM																					
11	BHP AT RATED RPM	263	263	607	140	126	126	67	67	76												
12	MAX BHP FOR INSTALLED IMP	30	30	70	20	700	700	35	35	10												
13	MANUFACTURER																					
14	TYPE AND SIZE																					
15	STAGES																					
16	SUCT. FLO. SIZE, RATING, FACE																					
17	DISCH. FLO. SIZE, RATING, FACE																					
18	NOZZLE ARRGT. SUCTION																					
19	NOZZLE ARRGT. DISCHARGE																					
20	IMPELLER DIA. REQUIRED																					
21	MAX IMP. DIA. AND HEAD																					
22	IMP. EYE AREA																					
23	NPSH REQUIRED																					
24	HYDROSTATIC TEST PRESSURE																					
25	THRUST BEARING																					
26	RADIAL BEARING																					
27	COUPLING																					
28	COUPLING GUARD																					
29	ROTATION FROM COUPLING END																					
30	MFG.																					
31	MECHANICAL SEALS																					
32	TYPE																					
33	AUX. BOX																					
34	BOX																					
35	WATER COOLED BEARING																					
36	PEDESTAL																					
37	WEIGHT OF PUMP AND BASE																					
38	CASING																					
39	CASE WEARING RINGS																					
40	STAGE PIECES																					
41	BUSHINGS																					
42	THROAT																					
43	INTERSTAGE																					
44	IMPELLER WEARING RINGS																					
45	IMPELLER																					
46	SHAFT SLEEVES																					
47	G. RING																					
48	INTERM. RINGS																					
49	BEARING HOUSING																					
50	BASE PLATE																					
51	GASKETS																					
52	SHAFTING: RING																					
53	SHAFTING: RING																					
54	SHAFTING: RING																					
55	SHAFTING: RING																					
56	MANUFACTURER'S SERIAL NO.																					
57	PERFORMABLE P.V.E.																					
58	DRAWING NO.																					
59	REV. SHEET																					
60	DATE																					
61	LOCATION																					
62	PLANT NO.																					
63	UNIT NO.																					
64	REVISION																					
65	BY																					

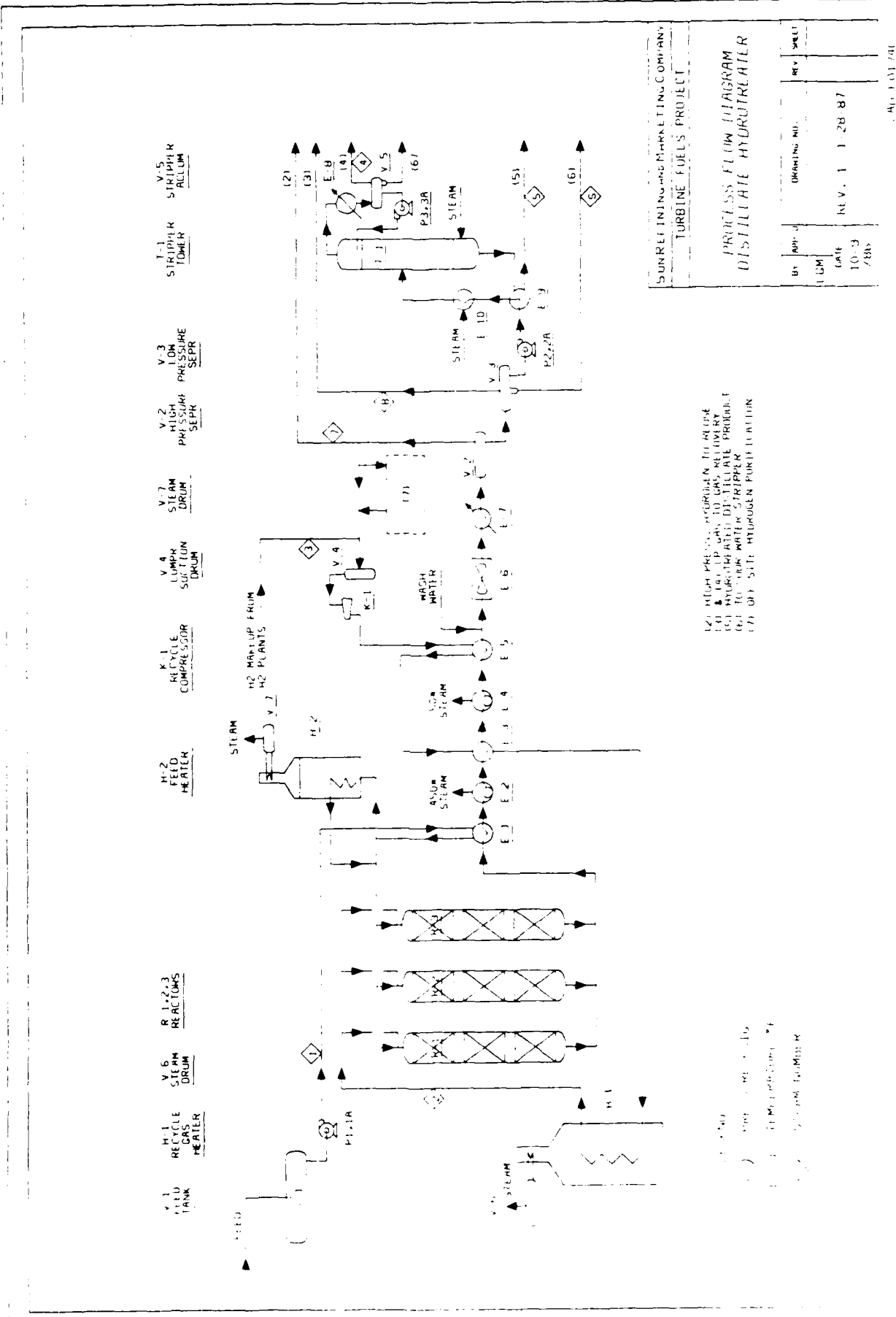
Sun Refining and Marketing Company  
 CENTRIFUGAL PUMP SCHEDULE & COMPRESSORS  
 PLANT AIR FORCE - NAPHTHA  
 HYDRO-TREATER  
 PLANT NO. \_\_\_\_\_ UNIT NO. \_\_\_\_\_  
 LOCATION \_\_\_\_\_  
 BY: \_\_\_\_\_ DRAWING NO. \_\_\_\_\_ REV. SHEET \_\_\_\_\_  
 DATE: 4/15/17  
 7

**PROCESS DESIGN SPECIFICATIONS**

for the

**DISTILLATE HYDROTREATING UNIT**





- V 1 RECYCLE TANK
- H 1 RECYCLE GAS HEATER
- V 6 STEAM DRUM
- R 1, 2, 3 REACTORS
- H 2 RECYCLE HEATER
- K 1 COMPRESSOR
- V 4 LAMP GLASS DRUM
- V 7 STEAM DRUM
- V 2 HIGH PRESSURE SEPR
- V 3 LOW PRESSURE SEPR
- I 1 STRIPPER
- V 5 STRIPPER ALLUM

(1) FEED  
 (2) REACTOR EFFLUENT  
 (3) REACTOR EFFLUENT  
 (4) REACTOR EFFLUENT  
 (5) REACTOR EFFLUENT  
 (6) REACTOR EFFLUENT  
 (7) REACTOR EFFLUENT  
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 (10) REACTOR EFFLUENT  
 (11) REACTOR EFFLUENT  
 (12) REACTOR EFFLUENT  
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 (17) REACTOR EFFLUENT

SUNBELT ENGINEERING-MARKETING COMPANY  
 TURBINE FUEL'S PROJECT  
 PROCESS FLOW DIAGRAM  
 DISTILLATE HYDROTREATER

BY	APP'D	DRAWING NO.	REV	SHEET
LGM			REV. 1	1
			10-3	
			1-28-87	
			7/88	

DISTILLATE HYDROTREATER UNIT  
MATERIAL BALANCE AND STREAM PROPERTIES

Stream Number	1	2	3	4	5	6	7	8
Stream Label	Liquid Feed	Reactor Gas Feed	Hydrogen Makeup	Stripper Tower Gas	Stripper Tower Bottoms	Sour Water from Low-Pressure Separator	High-Pressure Hydrogen Bleed Gas	Low-Pressure Separator Vent Gas
Stream Conditions								
Temperature, deg F	470	700	125	110	150	105	(normally)	121
Physical state	Liquid	Vapor	Vapor	Vapor	Liquid	Liquid	( none )	Vapor
API Gravity	18.7	-	-	-	31.1	10	-	-
Sp.Gr. @ 60 deg F	0.795	-	-	-	0.84	0.99	-	-
Sp.Gr. @ Temp.	-	2.00	2.72	32.8	-	-	-	7.32
Molecular Weight	2035	2585	2515	160	150	160	-	165
Pressure, psia	-	-	-	-	-	-	-	-
BBLS/DAY @ 60 deg F	30,652	-	-	-	33,354	19,141	-	-
MMSCFD	-	153.0	54.89	0.98	-	-	-	3.56
M lb/hr	420.82	47.13	16.40	3.51	422.85	279.14*	-	2.84
Vis., cSt @ Temp.	0.7	-	-	-	3.0	0.7	-	-

\* Does not include 5000 lb/hr sour water from the Stripper Overhead Accumulator Drum

## DISTILLATE HYDROTREATER DESIGN BASIS

### Distillate Feed Rate

The unit is designed to hydrotreat 30,652 BPSD of 18.7 °API San Ardo distillate with an ASTM end-point of 1000°F. The sulfur and nitrogen content of the feed is 0.85 and 0.79 weight percent. The sulfur and nitrogen content of the hydrotreated product will be less than 10 ppm. The feed distillate is a blend of 6,276 BPSD of crude unit light distillate and 24,376 BPSD of Hydrovisbreaker distillate. The unit will be capable of maintaining a 94% on-stream factor.

### Plant Processing Steps

All the straight-run naphtha and atmospheric gas oil from the crude unit boiling up to 650°F is fractionated for a 490°F cut point at the Naphtha Hydrotreater Feed Splitter. The 490-650°F bottoms product from the splitter is combined with all the 490-1000°F gas oil from the Hydrovisbreaker Unit to form the feedstock for the Distillate Hydrotreater.

The total distillate hydrotreater feed is heated by heat exchange with the reactor effluent followed by fired heating. This is fed to the top of the reactor at 700°F.

Recirculated hydrogen plus makeup hydrogen heated to 650 to 700°F by heat exchange and fired heat is also fed to the top of the reactor. A portion of the recirculated hydrogen is upgraded by a hydrogen purification unit from 92 to 99% hydrogen to remove excess methane from the system.

The reactor operates at 650 to 750°F and 2550 psig maximum over its operating cycle of about two years.

The reactor product is cooled and water washed to prevent fouling and corrosion of the final coolers. The unreacted hydrogen is separated and recycled to the reactor.

The distillate product is debutanized at the unit which permits the use of intermediate atmospheric pressure storage prior to the final recovery of the JP-4 and JP-8 by fractionation. Thus a short emergency outage of final fractionation will not force a shutdown of the Distillate Hydrotreater. Also the distillate can bypass the Gas Plant Debutanizer and can enter directly into the Main Fractionator at the Distillate Hydrocracker Unit.

## DISTILLATE HYDROTREATER UNIT

### Utilities and Chemical Requirements

#### Saturated Steam Produced, lb/hr

Steam Generators:	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>
E-2	44,970	-	-	44,970
E-4	-	-	46,700	46,700
V-6 at Heater H-1	-	10,600	-	10,600
V-7 at Heater H-2	-	4,240	-	4,240
<b>Totals</b>	<b>44,970</b>	<b>14,840</b>	<b>46,700</b>	<b>106,510</b>

#### Steam Used, lb/hr

	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>	<u>Condensate Recovered</u>
Heaters H1 & H2	8,800	-	-	8,800	-
Stripper tower	5,000	-	-	5,000	-
E-10 preheater	13,080	-	-	13,080	13,080
Misc. heating	-	5,000	10,000	15,000	-
<b>Total used:</b>	<b>26,880</b>	<b>5,000</b>	<b>10,000</b>	<b>41,880</b>	<b>13,080</b>
<b>Net export steam:</b>	<b>18,090</b>	<b>9,840</b>	<b>36,700</b>	<b>64,630</b>	

#### Boiler Feed Water

For steam produced 106,510 lb/hr  
 For 10% blowdown & wash water 10,651  
 Total gross BFW needed 117,161 lb/hr

Condensate recovered (reused as BFW) -13,000  
 Net BFW needed 104,161 lb/hr=208 gpm

<u>Cooling Water Circulated</u>	<u>Supply</u>	<u>Return</u>	<u>Duty</u>
E-7 2500 GPM	85°F	105°F	19.0 MMBTU/HR
E-8 473	<u>85</u>	110	<u>5.92</u> MMBTU/HR
<b>Total 2973 GPM</b>			<b>24.9 MMBTU/HR</b>

Make-up Cooling Water at 3% of total circulation = 90 GPM filtered water to replace loss to evaporation

## DISTILLATE HYDROTREATER UNIT

### Utilities and Chemical Requirements (continued)

#### Hydrotreating Catalyst

See Operating Requirements in overall refinery description section

#### Fuel Fired

Fuel: Vacuum Residuum Product from the Hydrovisbreaker Unit.

Heater H-1	48.6 MMBTU/HR	
Heater H-2	19.4	
Total	68.0 MMBTU/HR	= 11.3 FOE BBLs/HR = 3957 lb/hr Residuum Fuel

#### Electrical Power

	<u>Brake Horsepower Operating</u>	<u>Brake Horsepower Connected</u>
P-1,1A Feed Pumps	3000	6000
P-2,2A Stripper Feed	120	240
P-3,3A	3	6
E-6 Fans	80	80
 K-1 Recycle Compressor	<u>1200</u>	<u>1200</u>
Total	4403 BHP	7526 BHP
Kilowatts	3283 KW	5612 KW

#### Air Requirements

Dry Instrument Air	50 psig	250 SCFM
Plant Air	120 psig	

## DISTILLATE HYDROTREATING UNIT

### List of Major Equipment

#### Heat Exchangers

E-1	Reactor Feed - Effluent Exchanger
E-2	Reactor Effluent - 450 psig Steam Generator
E-3	Reactor Effluent - Recycle Gas Exchanger No. 2
E-4	Reactor Effluent - 50 psig Steam Generator
E-5	Reactor Effluent - Recycle Gas Exchanger No. 1
E-6	Reactor Effluent Aircooler
E-7	Reactor Effluent Trim Water Cooler
E-8	Stripper Overhead Condenser
E-9	Stripper Feed - Bottoms Exchanger
E-10	Stripper Feed Preheater

#### Fired Heaters

H-1	Recycle Gas Heater
H-2	Feed Heater

#### Distillation tower

T-1	Stripper Tower
-----	----------------

#### Reactors

R-1	Distillate Hydrotreater Reactor No. 1
R-2	Distillate Hydrotreater Reactor No. 2
R-3	Distillate Hydrotreater Reactor No. 3

#### Vessels

V-1	Feed Tank
V-2	High Pressure Separator
V-3	Low Pressure Separator
V-4	Compressor Suction Knockout Drum
V-5	Stripper Overhead Accumulator Drum
V-6	150 psig Steam Drum at Heater H-1
V-7	150 psig Steam Drum at Heater H-2

#### Pumps

P-1,1A	Feed Pumps
P-2,2A	Stripper Feed Pumps
P-3,3A	Stripper Reflux Pumps

#### Compressor

K-1	Recycle Gas Compressor
-----	------------------------

NOTES  
 DR - STEELS RELIEVED  
 DR - 1 RATED

ITEM NO	DESCRIPTION	UNIT	QTY	UNIT PRICE	TOTAL	REMARKS
1	REACTOR FEED EFFLUENT EXCH					
2	50 # STEAM GENERATOR					
3	#2 RECYCLE EFFLUENT EXCH					
4	50 # STEAM GENERATOR					
5	#3 RECYCLE EFFLUENT EXCH					
6	50 # STEAM GENERATOR					
7	#4 RECYCLE EFFLUENT EXCH					
8	50 # STEAM GENERATOR					
9	#5 RECYCLE EFFLUENT EXCH					
10	50 # STEAM GENERATOR					
11	#6 RECYCLE EFFLUENT EXCH					
12	50 # STEAM GENERATOR					
13	#7 RECYCLE EFFLUENT EXCH					
14	50 # STEAM GENERATOR					
15	#8 RECYCLE EFFLUENT EXCH					
16	50 # STEAM GENERATOR					
17	#9 RECYCLE EFFLUENT EXCH					
18	50 # STEAM GENERATOR					
19	#10 RECYCLE EFFLUENT EXCH					
20	50 # STEAM GENERATOR					
21	#11 RECYCLE EFFLUENT EXCH					
22	50 # STEAM GENERATOR					
23	#12 RECYCLE EFFLUENT EXCH					
24	50 # STEAM GENERATOR					
25	#13 RECYCLE EFFLUENT EXCH					
26	50 # STEAM GENERATOR					
27	#14 RECYCLE EFFLUENT EXCH					
28	50 # STEAM GENERATOR					
29	#15 RECYCLE EFFLUENT EXCH					
30	50 # STEAM GENERATOR					
31	#16 RECYCLE EFFLUENT EXCH					
32	50 # STEAM GENERATOR					
33	#17 RECYCLE EFFLUENT EXCH					
34	50 # STEAM GENERATOR					
35	#18 RECYCLE EFFLUENT EXCH					
36	50 # STEAM GENERATOR					
37	#19 RECYCLE EFFLUENT EXCH					
38	50 # STEAM GENERATOR					
39	#20 RECYCLE EFFLUENT EXCH					
40	50 # STEAM GENERATOR					
41	#21 RECYCLE EFFLUENT EXCH					
42	50 # STEAM GENERATOR					
43	#22 RECYCLE EFFLUENT EXCH					
44	50 # STEAM GENERATOR					
45	#23 RECYCLE EFFLUENT EXCH					
46	50 # STEAM GENERATOR					
47	#24 RECYCLE EFFLUENT EXCH					
48	50 # STEAM GENERATOR					
49	#25 RECYCLE EFFLUENT EXCH					
50	50 # STEAM GENERATOR					
51	#26 RECYCLE EFFLUENT EXCH					
52	50 # STEAM GENERATOR					
53	#27 RECYCLE EFFLUENT EXCH					
54	50 # STEAM GENERATOR					
55	#28 RECYCLE EFFLUENT EXCH					
56	50 # STEAM GENERATOR					
57	#29 RECYCLE EFFLUENT EXCH					
58	50 # STEAM GENERATOR					
59	#30 RECYCLE EFFLUENT EXCH					
60	50 # STEAM GENERATOR					
61	#31 RECYCLE EFFLUENT EXCH					
62	50 # STEAM GENERATOR					

ITEM NO	DESCRIPTION	UNIT	QTY	UNIT PRICE	TOTAL	REMARKS
63	STATIONARY TUBE SHEET					
64	FLOATING TUBE SHEET					
65	CROSS BAFFLES					
66	LONGITUDINAL BAFFLES					
67	TUBE SUPPORTS					
68	SHELL					
69	SHELL COVERS					
70	FLOATING HEAD COVER					
71	CHANNEL					
72	CHANNEL COVER					
73	CROSS BAFFLES TYPE SPACING					
74	LONGITUDINAL BAFFLE TYPE					
75	GASKETS					
76	WELONG FLANGES AND NOZZLES					
77	STUDS					
78	WEIGHT EXCHANGER					
79	WEIGHT BUNDLE					
80	CODE REQUIREMENTS					
81	SHR INC DRAWING NO					
82	REQUISITION NO					
83	PURCHASE ORDER NO					

HEAT EXCHANGER SCHEDULE  
 PLANT AIR FORCE  
 DISTILLATE HYDROTREATER  
 PLANT NO  
 LOCATION  
 BY APPD DRAWING NO REV SHEET  
 DATE 4/15/67  
 1



ITEM NO	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54	55	56	57	58	59	60	61	62
SERVICE																																																														
MANUFACTURER - NO SHELLS																																																														
SIZE AND TYPE																																																														
SURFACE/UNIT - SHELL																																																														
CONNECTED IN																																																														
FLUID CIRCULATED																																																														
QUANTITY																																																														
LARGE/SMALL																																																														
STEAM W/ATE %																																																														
FLUID VAPORIZED ON CONDENSER																																																														
STEAM CONDENSER																																																														
GRAVITY LIQUID (ISSUE)																																																														
VISCOSITY LIQUID (ISSUE)																																																														
SPEC HEAT LIQUID BTU/LB																																																														
LATENT HEAT VAPOR BTU/LB																																																														
OPER TEMP °F IN OUT																																																														
OPER PRESSURE PSIG																																																														
MINIMUM FOULING FACTOR																																																														
NUMBER OF PASSES																																																														
VELOCITY FT/SEC																																																														
PRESSURE DROP PSIG																																																														
HEAT EXCHANGER BTU/HR																																																														
HTD CORRECTED WEIGHTED																																																														
TRANSFER RATE SERVICE DESIGN																																																														
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CORROSION ALLOWANCE																																																														
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STUDS																																																														
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CODE REQUIREMENTS																																																														
SR & MC DRAWING NO																																																														
REQUISITION NO																																																														
PURCHASE ORDER NO																																																														
ITEM NO	6-6	E-7	E-A	F-9	F-10																																																									
SERVICE	EFFLUENT AIR COOLING	EFFLUENT WATER COOLING	STRIPPER CONDENSER	STRIPPER FEED HTMS EXCH.	STRIPPER FEED REHEATER																																																									
MANUFACTURER - NO SHELLS																																																														
SIZE AND TYPE	12X 40 AIR COOLING	19X 20 BBU	19X 20 BBU	40 X 20 ABS	25 X 20 BBU																																																									
SURFACE/UNIT - SHELL	65.10m SHELL	54" X 20" SHELL	54" X 20" SHELL	40 X 20 SHELL	25 X 20 SHELL																																																									
CONNECTED IN	AIR	WATER	WATER	HC	HC																																																									
FLUID CIRCULATED	AIR	WATER	WATER	HC	HC																																																									
QUANTITY	279	279	279	425.37	425.37																																																									
LARGE/SMALL	326	54	54	425.37	425.37																																																									
STEAM W/ATE %	3.0	2.0	2.0	0.91	0.91																																																									
FLUID VAPORIZED ON CONDENSER																																																														
STEAM CONDENSER	BH HC																																																													
GRAVITY LIQUID (ISSUE)																																																														
VISCOSITY LIQUID (ISSUE)																																																														
SPEC HEAT LIQUID BTU/LB																																																														
LATENT HEAT VAPOR BTU/LB																																																														
OPER TEMP °F IN OUT																																																														
OPER PRESSURE PSIG	205/164	165/157	165/145	110/360	450/450																																																									
MINIMUM FOULING FACTOR	2.0	2.0	2.0	1.0	1.0																																																									
NUMBER OF PASSES	2	1	1	1	1																																																									
VELOCITY FT/SEC	15	15	15	15	15																																																									
PRESSURE DROP PSIG	2.0	2.0	2.0	3.0	3.0																																																									
HEAT EXCHANGER BTU/HR	25,000,000	25,000,000	25,000,000	25,000,000	25,000,000																																																									
HTD CORRECTED WEIGHTED	35	35	35	35	35																																																									
TRANSFER RATE SERVICE DESIGN	35	35	35	35	35																																																									
DESIGN PRESSURE PSIG	20	20	20	20	20																																																									
TEST PRESSURE PSIG	30	30	30	30	30																																																									
DESIGN TEMPERATURE °F	100	100	100	100	100																																																									
CORROSION ALLOWANCE	1/8"	1/8"	1/8"	1/8"	1/8"																																																									
CONNECTIONS IN	1	1	1	1	1																																																									
SIZE STD TYPE OUT	4	4	4	4	4																																																									
MAT'L NO	35	35	35	35	35																																																									
TUBES O D	40	40	40	40	40																																																									
TUBES LENGTH	10	10	10	10	10																																																									
SHELL I D	120	120	120	120	120																																																									
STATIONARY TUBE SHEET																																																														
FLOATING TUBE SHEET																																																														
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4/15/67																																																														
UNIT NO																																																														
2																																																														

SERVICE	HEATER SECTION	PROCESS DESIGN CONDITIONS	NOTES
M-1 RECYCLE GAS HEATER 41.6546.5	M-2 HEATED HEATER 16.5749.35	<p>RADIANT CONVECTION</p> <p>PROCESS PRESSURE PSIG 2.38</p> <p>DESIGN FLUID TEMPERATURE DEG F 291.00</p> <p>ALLOWANCE TUBES FITTINGS</p> <p>HYDROSTATIC TEST PRESSURE PSIG 1000</p> <p>NUMBER OF PASSES 10</p> <p>OVERALL TUBE LENGTH FT 10</p> <p>EFFECTIVE TUBE LENGTH FT 10</p> <p>BARE TUBES NUMBER 10</p> <p>TOTAL EXPOSED SURFACE SQ FT 2830</p> <p>EXTENDED SURFACE TUBES NUMBER 1000</p> <p>TOTAL EXPOSED SURFACE SQ FT 2830</p> <p>TO BE SPALLING CENTER TO CENTER IN (STAGGERED) (IN LINE)</p> <p>TUBE CENTER TO SURFACE WALL IN MIN 1.0</p> <p>WELD INSPECTION REQUIREMENTS A HAY OR OTHER</p>	<p>H-2 HEATED HEATER</p> <p>11-2 ADAPT REVISION</p> <p>Sun Refining and Marketing Company</p> <p>FIRED HEATER DESIGN</p> <p>PLANT A/R FOLKE DISTILLATE</p> <p>HYDROTREATER UNIT NO</p> <p>LOCATION</p> <p>BY APP'D</p> <p>DATE 4/15/67</p> <p>DRAWING NO</p> <p>SHEET 3</p>
M-1 RECYCLE GAS HEATER	M-2 HEATED HEATER	<p>RADIANT CONVECTION</p> <p>PROCESS PRESSURE PSIG 4.50</p> <p>DESIGN FLUID TEMPERATURE DEG F 481.65</p> <p>ALLOWANCE TUBES FITTINGS</p> <p>HYDROSTATIC TEST PRESSURE PSIG 1000</p> <p>NUMBER OF PASSES 10</p> <p>OVERALL TUBE LENGTH FT 10</p> <p>EFFECTIVE TUBE LENGTH FT 10</p> <p>BARE TUBES NUMBER 10</p> <p>TOTAL EXPOSED SURFACE SQ FT 2830</p> <p>EXTENDED SURFACE TUBES NUMBER 1000</p> <p>TOTAL EXPOSED SURFACE SQ FT 2830</p> <p>TO BE SPALLING CENTER TO CENTER IN (STAGGERED) (IN LINE)</p> <p>TUBE CENTER TO SURFACE WALL IN MIN 1.0</p> <p>WELD INSPECTION REQUIREMENTS A HAY OR OTHER</p>	<p>H-2 HEATED HEATER</p> <p>11-2 ADAPT REVISION</p> <p>Sun Refining and Marketing Company</p> <p>FIRED HEATER DESIGN</p> <p>PLANT A/R FOLKE DISTILLATE</p> <p>HYDROTREATER UNIT NO</p> <p>LOCATION</p> <p>BY APP'D</p> <p>DATE 4/15/67</p> <p>DRAWING NO</p> <p>SHEET 3</p>
M-1 RECYCLE GAS HEATER	M-2 HEATED HEATER	<p>RADIANT CONVECTION</p> <p>PROCESS PRESSURE PSIG 4.50</p> <p>DESIGN FLUID TEMPERATURE DEG F 481.65</p> <p>ALLOWANCE TUBES FITTINGS</p> <p>HYDROSTATIC TEST PRESSURE PSIG 1000</p> <p>NUMBER OF PASSES 10</p> <p>OVERALL TUBE LENGTH FT 10</p> <p>EFFECTIVE TUBE LENGTH FT 10</p> <p>BARE TUBES NUMBER 10</p> <p>TOTAL EXPOSED SURFACE SQ FT 2830</p> <p>EXTENDED SURFACE TUBES NUMBER 1000</p> <p>TOTAL EXPOSED SURFACE SQ FT 2830</p> <p>TO BE SPALLING CENTER TO CENTER IN (STAGGERED) (IN LINE)</p> <p>TUBE CENTER TO SURFACE WALL IN MIN 1.0</p> <p>WELD INSPECTION REQUIREMENTS A HAY OR OTHER</p>	<p>H-2 HEATED HEATER</p> <p>11-2 ADAPT REVISION</p> <p>Sun Refining and Marketing Company</p> <p>FIRED HEATER DESIGN</p> <p>PLANT A/R FOLKE DISTILLATE</p> <p>HYDROTREATER UNIT NO</p> <p>LOCATION</p> <p>BY APP'D</p> <p>DATE 4/15/67</p> <p>DRAWING NO</p> <p>SHEET 3</p>



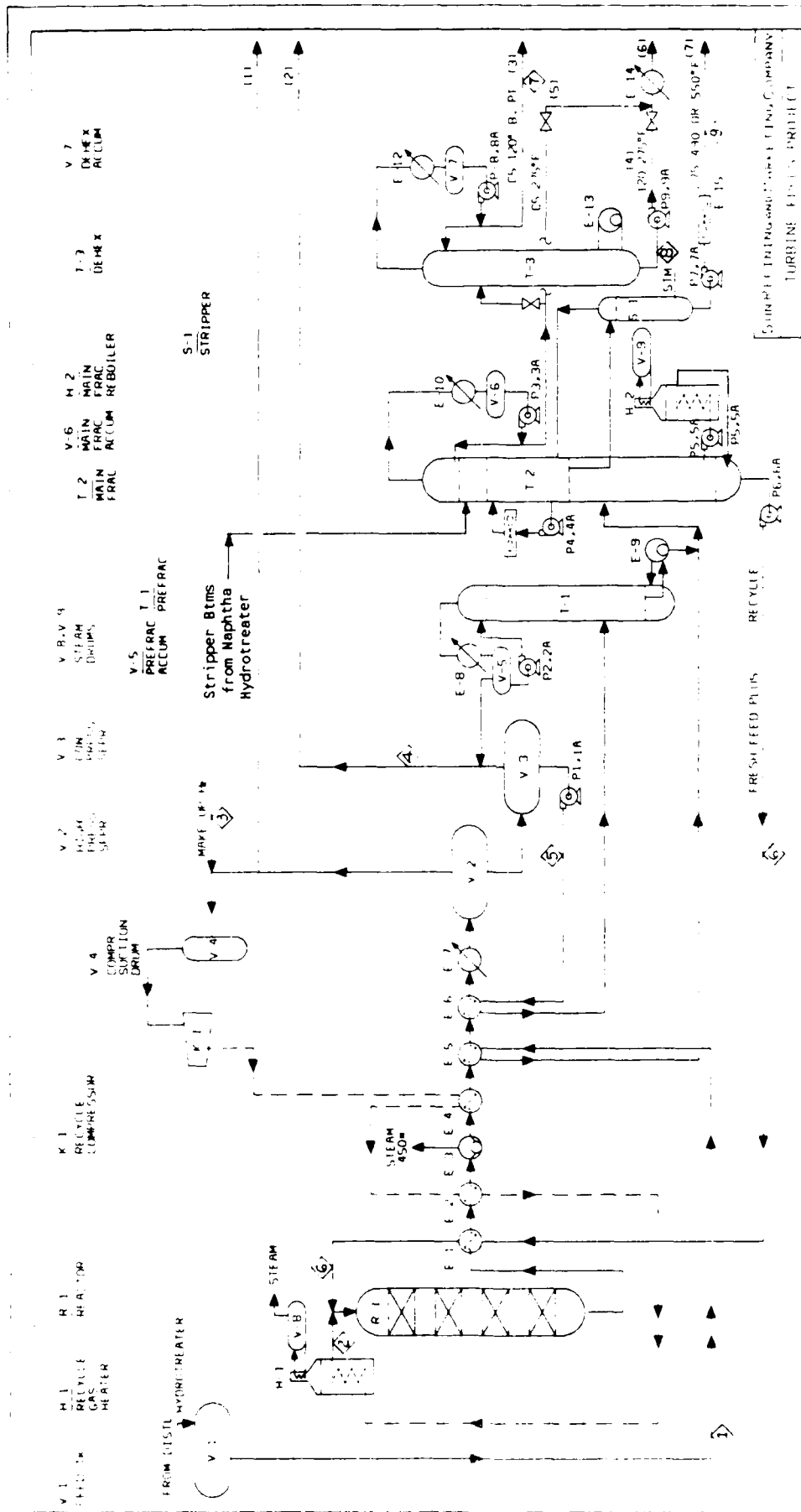
PUMP ITEM NO	SERVICE	COMPRESSOR		P.L.A. DISTL FEED	P2.2A STEAMP FEED	K3.3A STEAMP FEED	K-1 RECYCLE COMPRESSOR	NOTES
		LIQUID	VAPOR					
1		PUMPING TEMPERATURE	105					
2		SPECIFIC GRAVITY AT P.T.	1.295					
3		VISC. @ P.T.	1.0					
4		SPM AT P.T.	1109					
5		SUCTION PRESSURE (PSIB)	150					
6		DISCHARGE PRESSURE (PSIB)	244					
7		DIFFERENTIAL HEAD (FT)	255					
8		N.P.S.H. AVAILABLE	12					
9		RPM	175					
10		BHP AT RATED GPM	101					
11		MAX. BHP FOR RETALLED IMP.	130					
12		MANUFACTURER						
13		TYPE AND SIZE						
14		SUCT. FLG. SIZE, RATING, FACE						
15		DISCH. FLG. SIZE, RATING, FACE						
16		NOZZLE ARRGT. SUCTION						
17		NOZZLE ARRGT. DISCHARGE						
18		IMPELLER DIA. REQUIRED						
19		MAX. IMP. DIA. AND HEAD						
20		IMP. EYE AREA						
21		N.P.S.H. REQUIRED						
22		HYDROSTATIC TEST PRESSURE						
23		THRUST BEARING						
24		RADIAL BEARING						
25		COUPLING						
26		COUPLING GUARD						
27		ROTATION FROM COUPLING END						
28		MECHANICAL SEALS						
29		TYPE						
30		AUX. BOR.						
31		BOX						
32		WATER COOLED BEARING						
33		PEDESTAL						
34		WEIGHT OF PUMP AND BASE						
35		CASE						
36		CASE WEARING RINGS						
37		STAGE PIECES						
38		BUSHINGS						
39		INTERSTAGE						
40		IMPELLER WEARING RINGS						
41		SHAFT						
42		SHAFT SLEEVES						
43		G-RING						
44		LAND RINGS						
45		BEARING HOUSING						
46		BASE PLATE						
47		GASKETS						
48		STATIONARY RING						
49		SEAL						
50		ROTATING RING						
51		PACKING						
52		MANUFACTURER'S SERIAL NO.						
53		PERFORMANCE CURVE						
54		P. NUMBER						
55		30 & MC DRAWING NO.						
56								
57								

NO. DATE  
 REVISION  
 Sun Refining and Marketing Company  
 CENTRIFUGAL PUMP SCHEDULE  
 PLANT AIR FORCE DISTILLATE  
 MYMOTHEATER  
 LOCATION  
 BY APPD. DRAWING NO. REV SHEET  
 DATE 4/15/67  
 5

**PROCESS DESIGN SPECIFICATIONS**

**for the**

**DISTILLATE HYDROCRACKING UNIT**



ALL HIGH PRESSURE BLEED GAS NORMALLY FROM  
 (1) TO (10) BLEED GAS TO LAR 1.0 IN.  
 (11) TO (12) BLEED GAS TO LAR 1.0 IN. ISG  
 (13) TO (14) BLEED GAS TO WHEN FLOWING ISG  
 (15) TO (16) BLEED GAS TO WHEN FLOWING ISG  
 (17) TO (18) BLEED GAS TO WHEN FLOWING ISG  
 (19) TO (20) BLEED GAS TO WHEN FLOWING ISG  
 (21) TO (22) BLEED GAS TO WHEN FLOWING ISG

NOTE: ALL BLEED GAS FROM 1.0 IN. BLEED  
 PRESSURE TO LAR 1.0 IN.

PROCESS FLOW DIAGRAM  
 DIESEL FUEL HYDROTREATER

SURREY TRADING AND MARKETING COMPANY  
 TURBINE FUEL PROJECT

BY: _____	ENGR. NO. _____	REV. NO. _____
DATE: _____	_____	_____
10/17/57	_____	_____

DISTILLATE HYDROCRACKER UNIT  
MATERIAL BALANCE AND STREAM PROPERTIES

Stream Number	1	2	3	4	5	6	7	8	9
Stream Label	Unit Feed	Reactor Gas Feed	Makeup Hydrogen	Low Pressure Gas	Low Pressure Liquid	Total Reactor Liquid Feed	Dehexanizer Overhead Product	Dehexanizer Bottoms	Stripper Bottoms
Stream Conditions									
Temperature, deg F	150	110	125	150	118	500	112	242	120
Physical state	Liquid	Vapor	Vapor	Vapor	Liquid	Liquid	Liquid	Liquid	Liquid
API Gravity	33.3	-	-	-	55.9	30.2	91.1	58.6	46.6
Sp.Gr. @ 60 deg F	0.88	-	-	-	0.73	0.68	0.60	0.65	0.68
Sp.Gr. @ Temp.	-	4.0	2.7	30	-	-	-	-	-
Molecular Weight	150	2306	2306	105	105	31	27	34	75
Pressure, psia									
BBLS/DAY @ 60 deg F	33,503	-	-	-	50,000	44,184	5,506	19,650	26,051
MMSCFD	-	132.39	57.9	23.7	-	-	-	-	-
M lb/hr	422.9	-	-	-	500.2	503.6	51.0	213.2	301.8
Vis., cSt @ Temp.	3	-	-	-	0.4	0.3	0.2	0.2	0.5

## DISTILLATE HYDROCRACKER DESIGN BASIS

### Unit Feed Rate

The unit is designed to process 33,503 BPSD of fresh hydrotreated distillate from the Distillate Hydrotreater Debutanizer Bottoms. This stream is fed into the main fractionator where the <490°F boiling point feed and converted hydrocarbons are distilled off to final product.

The fresh feed plus recycle boiling above 490°F is routed from the bottom of the main fractionator to the reactor (26,505 plus 17,679 BPSD). The nominal 490°F cut point is used when producing JP-4; when producing JP-8 the cut point is 550°F. The above operation is based on 60% conversion per pass at the hydrocracker reactor.

The sulfur and nitrogen in the feed have been almost completely removed in the distillate hydrotreater before entering the hydrocracker.

### Plant Processing Steps

The fresh feed is routed through the main fractionator. The fresh feed plus recycle from the main fractionator bottoms is heat exchanged with reactor effluent and fed to the reactor mixed with preheated hydrogen. The mixture flows down through four beds of hydrocracking catalyst. The effluent is cooled by heat exchange, steam generation, and water cooling to 110°F. Hydrogen gas is separated in the high pressure separator, compressed, preheated and recycled to the reactor along with make-up hydrogen. The reactor operates at an inlet pressure of 2455 psig and 750°F. A liquid hourly space rate of 2.5/hour was selected based on pilot plant tests.

The liquid from the high pressure separator at 110°F and 2300 psig is flashed at 150 psig in the low pressure separator. The low pressure separator liquid is fed to the prefractionator. Prefractionator overhead product plus low pressure separator vapors are routed to the gas plant for butane recovery. The prefractionator bottoms is routed to the main fractionator to separate products from recycle liquid.



The main fractionator produces a <275°F cut overhead, a stripped side cut boiling between 275°F and 490°F (or 550°F for JP-8 production). The bottoms is reactor fresh feed plus recycle.

On JP-4 operation a downstream dehexanizer is operated to remove pentanes and hexanes overhead with the bottoms boiling between 120°F and 275°F blended into jet fuel boiling between 120°F and 490°F. The dehexanizer overhead is routed to naphtha storage for use as hydrogen plant furnace fuel and for sales for motor fuel blending.

When producing JP-8 which boils between 275°F and 550°F the dehexanizer is not operated and the main fractionator overhead C5-275°F cut is used for fuel at the hydrogen plant and is sold for motor fuel blending. The main fractionator stripped side cut boiling between 275°F and 550°F is the JP-8 product.

DISTILLATE HYDROCRACKER UNIT

Utilities and Chemical Requirements

Saturated Steam Produced, lb/hr

Steam Generators:	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>
E-3 Exchanger	51,000	-	-	51,000
V-8 at Heater H-1	-	3,400	-	3,400
V-7 at Heater H-2	-	40,000	-	40,000
Totals	51,000	43,400	-	94,400

Steam Used, lb/hr

	<u>450 psig</u>	<u>150 psig</u>	<u>50 psig</u>	<u>Total</u>	<u>Condensate Recovered</u>
Heaters H1 & H2	25,000	-	-	25,000	-
S-1 Stripper tower	-	2,000	-	2,000	-
T-20 Main Fract. Twr.	-	4,000	-	4,000	-
E-9 Reboiler	66,000	-	-	66,000	66,000
E-13 Reboiler	-	<u>28,500</u>	-	<u>28,500</u>	<u>28,500</u>
Totals	91,500	34,500	-	126,000	94,500
<u>Net Steam Import(export)</u>	40,500	(8,900)	-	31,600	

Boiler Feed Water

For steam produced	94,400 lb/hr
For 10% blowdown & wash water	<u>9,440 lb/hr</u>
Total gross BFW needed	103,840 lb/hr
Condensate recovered (reused as BFW)	<u>-94,500</u>
Net BFW needed	<u>9,340 lb/hr=19 gpm</u>

Cooling Water Circulated

<u>Exchanger</u>	<u>Rate</u>	<u>Supply</u>	<u>Return</u>	<u>Duty</u>
E-8	3536 GPM	85°F	110°F	44.2 MMBTU/HR
E-10	4627	85	115	69.4
E-12	1668	85	105	16.7
<u>E-14</u>	<u>1539</u>	<u>85</u>	<u>105</u>	<u>15.4</u>
Totals	11,370 GPM			145.7 MMBTU/HR

Make-up Cooling Water at 3% of total circulation = 341 GPM filtered water to replace loss to evaporation

## DISTILLATE HYDROCRACKER UNIT

### Utilities and Chemical Requirements (continued)

#### Heater Fuel Fired

Fuel: Vacuum Residuum Product from the Hydrovisbreaker Unit.

Heater H-1	55.44 MMBTU/HR	
Heater H-2	<u>182.87</u>	
	198.31 MMBTU/HR	= 33 FOE BBL/hr
		= 11,568 lb/hr Residuum Fuel

#### Electrical Power

	<u>Brake Horsepower Operating</u>	<u>Brake Horsepower Connected</u>
P-1,1A Prefractionator Tower Feed Pumps	175	350
P-2,2A Prefractionator Tower Reflux Pumps	55	110
P-3,3A Main Fract. Ovhd. Prod. & Reflux Pumps	115	230
P-4,4A Main Fract. Pumpharound Pumps	35	70
P-5,5A Main Fract. Reboiler Pumps	550	1100
P-6,6A Reactor Feed Pumps	4500	9000
P-7,7A Turbine Fuel Product Pumps	60	120
P-8,8A Dehexanizer Ovhd. Prod. & Reflux Pumps	20	40
P-9,9A Dehexanizer Bottoms Product Pumps	35	70
 K-1 Recycle Gas Compressor	 <u>1200</u>	 <u>1200</u>
Total Brake Horsepower	5,636 BHP	12,290 BHP
Kilowatts:	4,203 KW	9,165 KW

# DISTILLATE HYDROCRACKER UNIT

## List of Major Equipment

### Heat Exchangers

E-1	Reactor Feed - Effluent Exchanger
E-2	Reactor Effluent - Recycle Gas Exchanger
E-3	Reactor Effluent - 450 psig Steam Generator
E-4	Reactor Effluent - Recycle Gas Exchanger
E-5	Reactor Effluent - Main Fractionator Feed Exchanger
E-6	Reactor Effluent - Low Pressure Separator Liquid Exchanger
E-7	Reactor Effluent Cooler
E-8	Prefractionator Overhead Condenser
E-9	Prefractionator Reboiler
E-10	Main Fractionator Overhead Condenser
E-11	Main Fractionator Pumparound Reflux Aircooler
E-12	Dehexanizer Overhead Condenser
E-13	Dehexanizer Reboiler
E-14	Dehexanizer Bottoms Cooler
E-15	Turbine Fuel Product Aircooler

### Fired Heaters

H-1	Recycle Gas Heater
H-2	Main Fractionator Reboiler

### Fractionation Towers

T-1	Prefractionator
T-2	Main Fractionator
T-3	Dehexanizer
S-1	Main Fractionator Sidestripper

### Reactor

R-1	Distillate Hydrocracker Reactor
-----	---------------------------------

### Vessels

V-1	Feed Tank
V-2	High Pressure Separator
V-3	Low Pressure Separator
V-4	Compressor Suction Knockout Drum
V-5	Prefractionator Overhead Accumulator Drum
V-6	Main Fractionator Overhead Accumulator Drum
V-7	Dehexanizer Overhead Accumulator Drum
V-8	150 psig Steam Drum at Heater H-1
V-9	150 psig Steam Drum at Heater H-2

## DISTILLATE HYDROCRACKER UNIT

### List of Major Equipment (continued)

#### Pumps

P-1,1A Prefractionator Feed Pumps  
P-2,2A Prefractionator Reflux Pumps  
P-3,3A Main Fractionator Reflux Pumps  
P-4,4A Main Fractionator Pumparound Reflux Pumps  
P-5,5A Main Fractionator Reboiler Pumps  
P-6,6A Main Fractionator Bottoms (Reactor Feed) Pumps  
P-7,7A Sidestripper Bottoms Pumps  
P-8,8A Dehexanizer Reflux Pumps  
P-9,9A Dehexanizer Bottoms Pumps

#### Compressor

K-1 Recycle Gas Compressor

NOTES  
 SR - STRESS RELIEVED  
 SR - 1 RAYED

MANUFACTURER NO SHELLS  
 SIZE AND TYPE  
 SURFACE UNIT / SHELL

CONNECTED IN  
 FLUID CIRCULATED  
 QUANTITY VAPOR  
 LBS. HP STEAM  
 M / HR TOTAL  
 FLUID VAPORIZED OR CONDENSED  
 STEAM CONDENSED  
 GRAVITY LIQUID (SPEC. AT 100°F)  
 VISCOSITY LIQUID (SPEC. AT 100°F)  
 SPEC HEAT LIQUID BTU/LB  
 LATENT HEAT VAPOR BTU/LB  
 OPER TEMP °F IN OUT  
 OPER PRESSURE PSIG  
 MINIMUM FOULING FACTOR  
 NUMBER OF PASSES  
 VELOCITY FT/SEC  
 PRESSURE DROP PSIG  
 HEAT EXCHANGED BTU/HR (HT)  
 WTD CORRECTED WEIGHTED  
 TRANSFER RATE SERVICE CLEAN  
 DESIGN PRESSURE PSIG  
 TEST PRESSURE PSIG  
 DESIGN TEMPERATURE °F  
 CORROSION ALLOWANCE  
 CONNECTIONS IN  
 SIZE STD TYPE OUT  
 BATH  
 TUBES O D  
 B W G  
 LENGTH  
 PITCH  
 SHELL I D  
 G  
 STATIONARY TUBE SHEET  
 FLOATING TUBE SHEET  
 CROSS BAFFLES  
 LONGITUDINAL BAFFLES  
 TUBE SUPPORTS  
 SHELL COVERS  
 FLOATING HEAD COVER  
 CHANNEL COVER  
 CROSS BAFFLES TYPE SPACING  
 LONGITUDINAL BAFFLE TYPE  
 GASKETS  
 WELDING FLANGES AND NOZZLES  
 STUDS  
 WEIGHT EXCHANGER  
 WEIGHT BUNDLE  
 CODE REQUIREMENTS  
 BR & MC DRAWING NO  
 REQUISITION NO  
 PURCHASE ORDER NO

ITEM NO	DESCRIPTION	E-1	E-2	E-3	E-4	E-5
1	REFRACOR FIELD					
2	REFRACOR EFFLUENT					
3	REFRACOR EFFLUENT					
4	REFRACOR EFFLUENT					
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60	REFRACOR EFFLUENT					
61	REFRACOR EFFLUENT					
62	REFRACOR EFFLUENT					

CONSTRUCTION

NO DATE REVISION BY

Sun Refining and Marketing Company

HEAT EXCHANGER SCHEDULE

PLANT A/R FORCE  
 PLANT NO. HYDROCRACKER  
 UNIT NO.

LOCATION

BY APPD DRAWING NO REV SHEET

DATE 4/15/17

ITEM NO.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54	55	56	57	58	59	60	61	62
SERVICE																																	PERFORMANCE																CONSTRUCTION													
NOTES	SR - STRESS RELIEVED SR - STAYED																																																													

MANUFACTURER	KALAMAZOO																															
SIZE AND TYPE	12" X 20" BEU																															
SURFACE/UNIT	1/8" S																															
CONNECTED IN	E 7																															
FLUID CIRCULATED	KEROSENE																															
QUANTITY	492																															
FIXED BASES	132																															
STEAM	72.5																															
TOTAL	174																															
FLUID VAPORIZED OR CONDENSED	STEAM CONDENSED																															
GRAVITY LIQUID (SG) @ 60°F	0.85																															
VISCOSITY LIQUID (CP) @ 60°F	0.75																															
SPEC HEAT LIQUID (BTU/LB)	0.8																															
LATENT HEAT VAPOR (BTU/LB)	1000																															
OPER TEMP °F IN OUT	150 150																															
OPER PRESSURE PSIG	150																															
MINIMUM FOULING FACTOR	0.0001																															
NUMBER OF PASSES	1																															
VELOCITY FT/SEC	1.5																															
PRESSURE DROP PSIG	0.5																															
HEAT EXCHANGED (BTU/HR)	1000000																															
WGT CORRECTED WEIGHTED	1000000																															
TRANSFER RATE SERVICE CLEAN	1000000																															
DESIGN PRESSURE PSIG	150																															
TEST PRESSURE PSIG	225																															
DESIGN TEMPERATURE °F	150																															
CORROSION ALLOWANCE	0.015																															
CONNECTIONS IN	1/2"																															
CONNECTIONS OUT	1/2"																															
SIZE STD TYPE	1/2"																															
WALL	0.1875																															
OD	0.75																															
OBG	0.75																															
LENGTH	10																															
PITCH	10																															
SHIELD	0.015																															
STATIONARY TUBE SHEET	1/4"																															
FLOATING TUBE SHEET	1/4"																															
CROSS BAFFLES	1/4"																															
LONGITUDINAL BAFFLES	1/4"																															
TUBE SUPPORTS	1/4"																															
SHELL	1/4"																															
SHIELD COVERS	1/4"																															
FLOATING HEAD COVER	1/4"																															
CHANNEL	1/4"																															
CHANNEL COVER	1/4"																															
CROSS BAFFLES TYPE	SPACING																															
LONGITUDINAL BAFFLE TYPE	SPACING																															
GASKETS	GASKETS																															
WE DING FLANGES AND NOZZLES	GASKETS																															
STUDS	GASKETS																															
WEIGHT EXCHANGER	GASKETS																															
WEIGHT BUNDLE	GASKETS																															
CODE REQUIREMENTS	GASKETS																															
WEIGHT DRAWING NO	GASKETS																															
REQUISITION NO	GASKETS																															
PURCHASE ORDER NO	GASKETS																															

PLANT	11K	2000
UNIT NO	11K	2000
LOCATION	11K	2000
BY	11K	2000
DATE	11K	2000
REVISION	11K	2000

ITEM NO	DESCRIPTION	UNIT	QTY	UNIT PRICE	TOTAL	REMARKS
1	MANUFACTURER - NO SHELLS					
2	SIZE AND TYPE					
3	SURFACE/UNIT - /SHELL					
4	CONNECTED IN					
5	FLUID CIRCULATED					
6	LIQUID					
7	VAPOR					
8	QUANTITY					
9	PIPED GASES					
10	STEAM					
11	TOTAL					
12	FLUID VAPORIZED OR CONDENSED					
13	STEAM CONDENSED					
14	GRAVITY - LIQUID (SG/WT) AT 70					
15	VISCOSITY - LIQUID (SSU/CPI) AT 70					
16	SPEC HEAT - LIQUID (BTU/LB)					
17	LATENT HEAT - VAPOR (BTU/LB)					
18	OPER TEMP °F					
19	OPER PRESSURE PSIG					
20	MINIMUM FOULING FACTOR					
21	NUMBER OF PASSES					
22	VELOCITY FT/SEC					
23	PRESSURE DROP PSIG					
24	HEAT EXCHANGER DUTY/HR					
25	MT/CONNECTED WEIGHT					
26	TRANSFER RATE SERVICE CLEAN					
27	DESIGN PRESSURE PSIG					
28	TEST PRESSURE PSIG					
29	DESIGN TEMPERATURE °F					
30	CORROSION ALLOWANCE IN					
31	CONNECTIONS IN					
32	SIZE STD. TYPE					
33	OUT					
34	MATL					
35	TUBES					
36	OD					
37	BBG					
38	LENGTH					
39	PITCH					
40	SHELL I.D.					
41	OD					
42	STATIONARY TUBE SHEET					
43	FLOATING TUBE SHEET					
44	CROSS Baffles					
45	LONGITUDINAL Baffles					
46	TUBE SUPPORTS					
47	SHELL					
48	SHELL COVERS					
49	FLOATING HEAD COVER					
50	CHANNEL					
51	CHANNEL COVER					
52	CROSS Baffles TYPE					
53	LONGITUDINAL Baffle TYPE					
54	GASSETS					
55	WELDING FLANGES AND NOZZLES					
56	STUDS					
57	WEIGHT EXCHANGER					
58	DRY WET					
59	WEIGHT BURDLE					
60	CODE REQUIREMENTS					
61	SHR INC DRAWING NO					
62	REQUISITION NO					
63	PURCHASE ORDER NO					

NOTES  
 SA - STRESS RELIEVED  
 SR - R RATED

Sun Refining and Marketing Company

HEAT EXCHANGER SCHEDULE

PLANT AIR FORCE  
 DISTILLATE HYDROCRACKER  
 UNIT NO

LOCATION  
 BY APPD DRAWING NO REV SHEET  
 LCN DATE  
 4/25/67 3





ITEM NO	TITLE	SERIAL NO	DATE	NO	DATE	REVISION	BY
7-1	PRETRAC	200	1-2	V-3	V-4	V-5	V-6
7-2	MAIN FRAC	11	1-1	V-1	V-2	V-3	V-4
7-3	DE AREA	1	1-1	V-1	V-2	V-3	V-4
7-4	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-5	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-6	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-7	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-8	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-9	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-10	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-11	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-12	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-13	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-14	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-15	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-16	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-17	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-18	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-19	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-20	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-21	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-22	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-23	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-24	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-25	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-26	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-27	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-28	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-29	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-30	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-31	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-32	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-33	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-34	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-35	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-36	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-37	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-38	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-39	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-40	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-41	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-42	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-43	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-44	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-45	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-46	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-47	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-48	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-49	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-50	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-51	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-52	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-53	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-54	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-55	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-56	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-57	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-58	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-59	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-60	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-61	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-62	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-63	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-64	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-65	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-66	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-67	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-68	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-69	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-70	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-71	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-72	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-73	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-74	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-75	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-76	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-77	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-78	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-79	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-80	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-81	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-82	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-83	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-84	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-85	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-86	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-87	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-88	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-89	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-90	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-91	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-92	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-93	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-94	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-95	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-96	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-97	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-98	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-99	TRAYS	1	1-1	V-1	V-2	V-3	V-4
7-100	TRAYS	1	1-1	V-1	V-2	V-3	V-4

Sun Refining and Marketing Company

TOWER & VESSEL SCHEDULE

PLANT NO. 401 FORCE  
PLANT NO. 402 FORCE  
PLANT NO. 403 FORCE

LOCATION: ...  
DRAWING NO.: ...  
REV. SHEET: ...  
DATE: ...

NOTES

ITEM NO	V-7	V-8	V-9	1
TITLE	LEAK ACCUM	STEAM DRUM	STEAM DRUM	MANHOLES
SERIAL NO				
ASSEMBLY				
DETAILS				
TRAYS				
PLATFORMS				
CODE	AD	200	200	2735
PRESSURE	650	650	650	650
TEMPERATURE °F	1275	148	13	VIA INTRINSIC
CORROSION ALLOWANCE				
WIND LOADING				
HYDROSTATIC TEST PRESSURE				
HAMMER TEST PRESSURE				
STRESS RELIEVED	NO	NO	NO	YES
RADIOGRAPHED	SABT	SABT	SABT	YES
VERTICAL OR HORIZONTAL	HORIZ	HORIZ	HORIZ	VERT
INSULATION THICKNESS	5"	4"	7"	8"
INSULATION	7.5"	7D	20"	80"
LENGTH BEAM TO BEAM				
LENGTH BASE SECTION				
HEIGHT OF SKIRT				
E TO BOTTOM OF SUPPORTS	37.5"	1.50	7.5	8.87
TOTAL SHELL THICKNESS	30.5"	.50	1.25	5"
HEAD THICKNESS				
SKIRT THICKNESS	1"	1"	2.0"	5"
NO TRAYS	24"	2.0"	2.0"	2.0"
SIZE				
SERIES B FACING				
THICKNESS				
LOCATION				
NO TRAYS REQ'D B SPACING				
TYPE OF TRAY				
NO OF CAPS/TRAY B TYPE				
SIZE OF CAPS				
SIZE OF RISERS				
TYPE OF DOWNCOMER				
NO TRAYS REQ'D B SPACING				
TYPE OF TRAY				
TRAY B				
TRAY A				
NO				
DATE				
REVISION				
BY				

Sun Refining and Marketing Company

TOWER & VESSEL SCHEDULE

PLANT A, B FORCE  
 PLANT NO 1000000000  
 UNIT NO

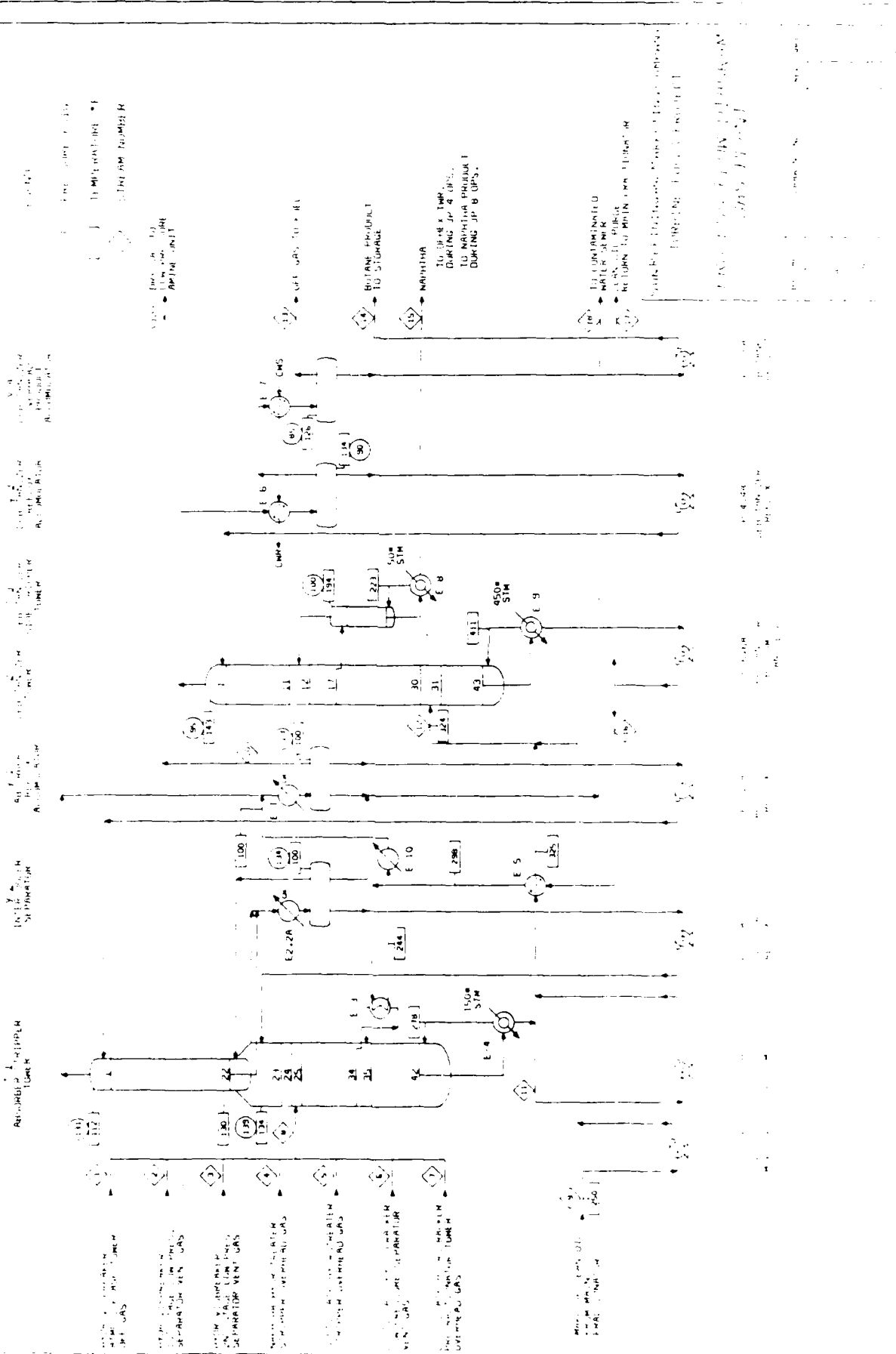
DATE  
 DRAWING NO  
 SHEET



**PROCESS DESIGN SPECIFICATIONS**

**for the**

**GAS PLANT**



GAS PLANT  
MATERIAL BALANCE AND STREAM PROPERTIES

Stream Number	8	9	10	11	13	14	16	17
Stream Label	Feed Gas	Lean Oil Makeup	Dry Gas	Rich Oil	Debut Gas (normally)	Butane Product	Naptha to Dehex	Lean Oil Purge
Stream Conditions					(none)			
Temperature, deg F	141	250	100	278	-	120	223	411
Physical state	Vapor	Liquid	Vapor	Liquid	-	Liquid	Liquid	Liquid
API Gravity	-	55.0	-	68.0	-	117	92	55
Sp.Gr. @ Temp.	-	.60	-	.58	-	.57	.52	.50
Molecular Weight	30	-	14.8	-	-	-	-	-
Pressure, psia	154	165	130	155	-	100	110	221
88LS/DAY @ 60 deg F	-	975	-	30,394	-	7100	1700	21519
MMSCFD	33.93	-	22.04	-	-	-	-	-
M lb/hr	111,910	10,750	30,700	312,920	-	59,520	16,590	9,540
Vis., cSt @ Temp.	-	0.3	-	0.2	-	0.3	0.3	0.3

NOTE: Stream 1 is composed of the following:

Stream No.	1	2	3	4	5	6	7	Total
lb/hr	2615	5857	5100	1204	3529	20,719	72,920	111,910
Molecular Weight	29.0	10.2	16.2	29.8	32.7	16.5	54.0	-

## GAS PLANT DESCRIPTION

### Feeds and Products

Each of the major processing plants in the refinery produces low-pressure vent gases and/or tower overhead gases, which are valuable sources of fuel gas, hydrogen plant feed, and liquid hydrocarbon products. Most also contain high levels of hydrogen sulfide which must be removed and converted to elemental sulfur to reduce environmental pollution.

Some of the vented refinery gases are sufficiently pure in hydrogen, methane, ethane and propane so that they can be used directly as fuel gas and hydrogen plant feed with no further processing. Those streams bypass the Gas Plant. Included in this group are:

- (1) Naphtha Hydrotreater Low-Pressure Vent Gas
- (2) Distillate Hydrotreater Low-Pressure Vent Gas
- (3) Hydrogen Purification Unit Vent Gas

However, the Gas Plant feed streams contain significant quantities of butane and naphtha which can be recovered as liquid products. These streams include:

- (1) Hydrovisbreaker Atmospheric Flash Tower Off-Gas
- (2) Hydrovisbreaker 1st-stage Low-Pressure Separator Vent Gas
- (3) Hydrovisbreaker 2nd-stage Low-Pressure Separator Vent Gas
- (4) Naphtha Hydrotreater Stripper Overhead Gas
- (5) Distillate Hydrotreater Stripper Overhead Gas
- (6) Distillate Hydrocracker Low-Pressure Separator Vent Gas
- (7) Distillate Hydrocracker Prefractionator Tower Overhead Gas



Consequently the purpose of the Gas Plant is to separate the gas feed mixture into three main products:

- (1) Dry Gas - enroute to sulfur removal prior to use as fuel gas or hydrogen plant feedstock
- (2) Butane - enroute to treating for sulfur removal prior to sale
- (3) Naphtha - to be mixed with the Main Fractionator naphtha product

#### Process Flow Description

The Gas Plant consists primarily of an Absorber-Stripper Tower and a Debutanizer Tower with a Sidestripper. The Absorber-Stripper has a total of 42 trays. It actually consists of two towers constructed in piggy-back fashion with a 22-tray absorber section on top of a 20-tray stripper section. The Debutanizer has 43 trays and its Sidestripper has 9 trays.

The Absorber-Stripper separates the dry gas from the butane and heavier hydrocarbon products with the aid of "lean oil" circulated from the Debutanizer bottoms. The Debutanizer and Sidestripper towers subsequently separate the butane, naphtha, and lean oil.

The feedstock gas mixture for the Gas Plant is fed to the Absorber-Stripper column near the top of the Stripper section. The only other feed stream entering the column is the lean oil (Debutanizer bottoms product) which is fed to the top of the Absorber to recover butane and heavier components that may be entrained in the dry gas.

The Absorber bottoms liquid and the Stripper overhead vapor are routed through the Absorber intercoolers and pass into a phase separator drum. Drum liquid is pumped to the Stripper top tray, and drum vapor is fed to the bottom of the Absorber.

The Absorber intercoolers serve to reduce the minimum required capacity (diameter) of the Absorber section by moving some of the overhead condenser duty to a point below the Absorber section. This condenses some of the higher-boiling components that would otherwise flow upward into the Absorber from the Stripper, and thereby reduces the vapor/liquid traffic through the entire Absorber.

Liquids traveling down the Stripper are stripped of light hydrocarbons by addition of heat from an interheater and a bottoms reboiler. The interheater is placed higher in the Stripper than the reboiler for reasons similar to those above concerning placement of the Absorber intercooler. By placing the interheater higher in the column, it strips out some of the lower-boiling components that would otherwise flow down through the Stripper section. Consequently this reduces the vapor/liquid traffic in the lower Stripper, which reduces the minimum required diameter of the Stripper section.

Debutanizer bottoms serves as the lean oil for the Absorber. It is fed to the Absorber overhead vapor stream upstream of the condenser and enters the Absorber with the liquid reflux. The hot Debutanizer bottoms first is cooled by heat exchange with the Stripper bottoms (rich oil), then the Stripper interheater, and finally with a lean oil cooler prior to being fed to the Absorber overhead condenser.

The heated Stripper bottoms (rich oil) is combined with make-up lean oil to form the Debutanizer feed. The make-up lean oil stream is drawn from an upper tray of the Main Fractionator Tower at the Distillate Hydrocracker Unit.

The Debutanizer overhead vapor is combined with lean oil and passed through the overhead condenser and into the reflux accumulator drum. The condensate with the lean oil is returned to the Absorber as reflux. The uncondensed vapors are sent to a secondary condenser to form bubble-point butane product in the overhead product accumulator drum.

Naphtha is drawn from the middle section of the Debutanizer and stripped in the Sidestripper by reboiler heat. The Sidestripper overhead is returned to a higher tray in the Debutanizer tower. The light naphtha from the Sidestripper bottoms is blended with other naphtha from the Main Fractionator overhead.

GAS PLANT  
DESIGN BASIS

FEED STREAMS

(See Material Balance and Stream Properties for more detail.)

1. Hydrovisbreaker Atmospheric Flash Tower Off-Gas
2. Hydrovisbreaker 1st-stage Low-Pressure Separator Vent Gas
3. Hydrovisbreaker 2nd-stage Low-Pressure Separator Vent Gas
4. Naphtha Hydrotreater Stripper Overhead Gas
5. Distillate Hydrotreater Stripper Overhead Gas
6. Distillate Hydrocracker Low-Pressure Separator Vent Gas
7. Distillate Hydrocracker Prefractionator Tower Overhead Gas
8. Lean Oil Make-Up

OPERATING SPECIFICATIONS

1. Absorber-Stripper

C4 recovery in Stripper bottoms (Rich Oil): 99% of C4 feed

C4<sup>+</sup> in Absorber overhead (Dry Gas): 2 mole% C4<sup>+</sup> MAX.

Adjust C3<sup>-</sup> in Stripper bottoms (Rich Oil)  
so that C3<sup>-</sup> in Debut Ovhd. product is: 2 mole% C3<sup>-</sup> MAX.

GAS PLANT  
DESIGN BASIS  
(continued)

2. Debutanizer

C4 recovery in overhead butane product:	99% of C4 feed
C3 <sup>-</sup> in Ovhd. product (see Abs-Stripper above):	2 mole% C3 <sup>-</sup> MAX.
C5 <sup>+</sup> in Ovhd. product	3 mole% C5 <sup>+</sup> MAX.
Sidedraw TBP boiling range	C5 - 175°F
Bottoms TBP boiling range	175°F <sup>+</sup>

3. Debutanizer Sidestripper

C4 <sup>-</sup> in stripped bottoms	5 mole% C4 <sup>-</sup> MAX.
Bottoms TBP boiling range	C5 - 175°F

UTILITIES

1. Cooling water: supply at 85°F / return at 105°F
2. Saturated steam pressures available: 450, 150, 50 psig

GAS PLANT  
LIST OF MAJOR EQUIPMENT

<u>TOWERS</u>		<u>Height</u>	<u>I.D.</u>	<u>No. of Trays</u>
T-1	Absorber-Stripper	123 ft.	6 ft/9 ft	42
T-2	Debutanizer	110 ft.	8.5 ft	43
T-3	Debutanizer Sidestripper	36 ft.	3.5 ft	9

<u>VESSELS</u> (all horizontal)		<u>I.D.</u>	<u>S-S Length</u>
V-1	Absorber Reflux Accumulator	6.5 ft	21 ft
V-2	Intercooler Separator	9 ft	20 ft
V-3	Debutanizer Reflux Accumulator	5 ft	11 ft
V-4	Debutanizer Overhead Product Accumulator	5 ft	12 ft

<u>HEAT EXCHANGERS, REBOILERS, CONDENSERS</u>		<u>Duty, MMBTU/hr</u>
E-1	Absorber Overhead Condenser	3.110
E-2,2A	Absorber Intercoolers	19.753
E-3	Stripper Interheater	8.000
E-4	Stripper Reboiler	34.072
E-5	Debutanizer Feed/Bottoms Heat Exchanger	17.543
E-6	Debutanizer Overhead Condenser	23.657
E-7	Debutanizer Overhead Product Condenser	8.105
E-8	Debutanizer Sidestripper Reboiler	4.021
E-9	Debutanizer Reboiler	25.470
E-10	Lean Oil Cooler	18.525

GAS PLANT  
LIST OF MAJOR EQUIPMENT  
(continued)

PUMPS

P-1,1A Absorber Reflux  
P-2,2A Stripper Reflux  
P-3,3A Stripper Bottoms (Rich Oil)  
P-4,4A Debutanizer Reflux  
P-5,5A Butane Product  
P-6,6A Debutanizer Bottoms (Lean Oil)  
P-7,7A Lean Oil Make-Up

GAS PLANT

UTILITY REQUIREMENTS

<u>Saturated Steam</u>	<u>lb/hr</u>
450 psig      E-9 Debutanizer Reboiler:	33294
150 psig      E-4 Stripper Reboiler:	39555
50 psig        E-8 Debutanizer Sidestripper Reboiler:	<u>4376</u>
Total Boiler Feed Water:	77225 lb/hr

<u>Cooling Water</u> (85°F supply; 105°F return)	<u>gal/min</u>
E-1      Absorber Condenser:	415
E-2,2A    Absorber Intercoolers:	1975
E-6      Debutanizer Overhead Condenser and	
E-7      Debutanizer Overhead Product Condenser:	3175
E-10     Lean Oil Cooler:	<u>1853</u>
Total Cooling Water:	7418 gpm

<u>Process Electricity</u> (requirement for one pump from each pair)	<u>KW</u>
P-1,1A    Absorber Reflux	50
P-2,2A    Intercooler Separator Liquid	55
P-3,3A    Stripper Bottoms (Rich Oil)	15
P-4,4A    Debutanizer Reflux	35
P-5,5A    Butane Product	25
P-6,6A    Debutanizer Bottoms	75
P-7,7A    Lean Oil Make-up	<u>5</u>
Total	260 KW



ITEM NO	DESCRIPTION	E-1	E-2, 2A	E-3	E-4	E-5
1	ABSORBER	CONDENSER	ABSORBER	STRIPPER	STRIPPER	DEBUT. REBO/BOTTOMS
2	CONDENSER	CONDENSER	ABSORBER	STRIPPER	STRIPPER	HEAT EXCHANGER
3	MANUFACTURER - NO	48" x 20'	48" x 20'	48" x 20'	48" x 20'	48" x 20'
4	SIZE AND TYPE	BEU	BEU	BEU	BEU	BEU
5	SURFACE/UNIT - /SHELL	3347	5710	1907	2039	2297
6	CONNECTED IN	TUBE	COILING	TUBE	TUBE	TUBE
7	FLUID CIRCULATED	COOLING WATER	COILING WATER	LEAN OIL	150 PIG	STRIPPER
8	STARTING QUANTITY	415 GPM	1975 GPM	247442	5870 STRAM	DEBUT
9	STEAM	58.934	168.135	531784	312773	247442
10	TOTAL	213.071	443.514	267442	39555	247442
11	STEAM COMBUSTED	2.498	55.853	487270	282.449	29041
12	GRAVITY - LIQUID (SG) AT 60°	7.3	7.2	6.5	7.1	7.1
13	VISCOSITY - LIQUID (CP) AT 60°	1.5	1.5	1.3	1.3	1.3
14	SPEC HEAT LIQUID BTU/LB	173.5/100	145/100	133/200	229/238	278/215
15	LATENT HEAT VAPOR BTU/LB	1031	1031	1031	1031	1031
16	OPER TEMP °F IN - OUT	131	131	131	131	131
17	OPER PRESSURE PSIG	1	1	1	1	1
18	NUMBER OF PASSES	1	1	1	1	1
19	NUMBER POLING FACTOR	1	1	1	1	1
20	VELOCITY FT/SEC	1	1	1	1	1
21	PRESSURE DROP PSIG	8	5	2	1	1
22	NET EXCHANGED BTU/HR (MM)	311012	145228	80000	340724	1756321
23	WTD CONNECTED WEIGHT	133	242	711	1074	318
24	TRANSFER RATE SERVICE CLEAN	70	70	70	70	70
25	DESIGN PRESSURE PSIG	156	164	165	175	184
26	DESIGN TEMPERATURE °F	234	244	248	262	271
27	ALLOWANCE	400	400	400	450	450
28	CONNECTIONS IN	1/2"	1/2"	1/2"	1/2"	1/2"
29	SIZE STD TYPE	IN	IN	IN	IN	IN
30	MATL	ADMICRIT	ADMICRIT	CS	CS	CS
31	NO	3/8" x 14	3/8" x 14	3/8" x 14	3/8" x 14	3/8" x 14
32	PITCH	20" x 1"	20" x 1"	20" x 1"	20" x 1"	20" x 1"
33	LENGTH	37'	41'	25'	66' x 33'	48'
34	SHELL I.D.	48"	48"	48"	48"	48"
35	STATIONARY TUBE SHEET					
36	FLOATING TUBE SHEET					
37	CROSS BAFFLES					
38	LONGITUDINAL BAFFLES					
39	TUBE SUPPORTS					
40	SHELL COVERS					
41	FLUENT HEAD COVER					
42	CHANNEL COVER					
43	CROSS BAFFLES TYPE - SPACING					
44	LONGITUDINAL BAFFLE TYPE					
45	BASKETS					
46	WELDING FLANGES AND NOZZLES					
47	STUDS					
48	WEIGHT EXCHANGER					
49	WEIGHT BUDDLE					
50	CODE REQUIREMENTS					
51	ASME DRAWING NO					
52	REQUISITION NO					
53	PURCHASE ORDER NO					

NO. 1 STRESS RELIEVED  
 IN - LAYED

MOISE  
 Sun Refining and Marketing Company  
 HEAT EXCHANGER SCHEDULE  
 PLANT GAS PLANT  
 PLANT NO. UNIT NO.  
 LOCATION  
 BY APPD. DRAWING NO. REV. SHEET  
 JR3 DATE 4/25/67  
 NO. DATE REVISION

ITEM NO	DESCRIPTION	UNIT	QTY	UNIT PRICE	TOTAL	REMARKS
1	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
2	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
3	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
4	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
5	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
6	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
7	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
8	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
9	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
10	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
11	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
12	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
13	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
14	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
15	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
16	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
17	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
18	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
19	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
20	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
21	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
22	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
23	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
24	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
25	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
26	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
27	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
28	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
29	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
30	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
31	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
32	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
33	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
34	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
35	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
36	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
37	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
38	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
39	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
40	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
41	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
42	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
43	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
44	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
45	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
46	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
47	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
48	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
49	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
50	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
51	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
52	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
53	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
54	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
55	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
56	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
57	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
58	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
59	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
60	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
61	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
62	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	

NOTES  
 SA - STRESS RELIEVED  
 SA - 11 RATED

ITEM NO	DESCRIPTION	UNIT	QTY	UNIT PRICE	TOTAL	REMARKS
63	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
64	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
65	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
66	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
67	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
68	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
69	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
70	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
71	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
72	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
73	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
74	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
75	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
76	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
77	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
78	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
79	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
80	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
81	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
82	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
83	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
84	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
85	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
86	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
87	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
88	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
89	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
90	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
91	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
92	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
93	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
94	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
95	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
96	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
97	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
98	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
99	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	
100	DEANIZER OVHD CONDENSER	36V	1	50.00	50.00	

CONSTRUCTION  
 STATIONARY TUBE SHEET  
 FLOATING TUBE SHEET  
 CROSS BAFFLES  
 LONGITUDINAL BAFFLES  
 TUBE SUPPORTS  
 SHELL  
 SHELL COVERS  
 FLOATING HEAD COVER  
 CHANNEL COVER  
 CHANNEL COVER  
 CROSS BAFFLES TYPE SPACING  
 LONGITUDINAL BAFFLE TYPE  
 BASKETS  
 WELDING FLANGES AND NOZZLES  
 STUDS  
 WEIGHT EXCHANGER  
 WEIGHT BUNDLE  
 CODE REQUIREMENTS  
 SHELL DRAWING NO.  
 REQUISITION NO.  
 PURCHASE ORDER NO.

REVISION  
 DATE  
 BY  
 JRS  
 DRAWING NO.  
 UNIT NO.  
 PLANT GAS PLANT  
 HEAT EXCHANGER SCHEDULE  
 Sun Refining and Marketing Company  
 4/15/87  
 2

ITEM NO.	T-1	T-2	T-3	V-1	V-2	V-3	V-4	NOTES		
	ABSORBER SKIPPER TOWER	DEBURNIZER TOWER	DEBURNIZER TOWER	ABSORBER ACRYLIC ACROMOLINA	INTERMEDIATE SEMI-TRA	DEBURNIZER ACROMOLINA	DEBURNIZER ACROMOLINA			
	SERIAL NO.									
	ASSEMBLY DETAILS									
	TRAYS									
	PLATFORMS									
DESIGN	CODE									
	PRESSURE PSIG	150	128	150	162	112	110			
	TEMPERATURE OF	650	650	650	650	650	650			
	CORROSION ALLOWANCE	185/16"	18"	133"	17"	15"	133"			
	WIND LOADING									
	HYDROSTATIC TEST PRESSURE PSIG	250	192	225	243	172	165			
	HAMMER TEST PRESSURE									
	STRESS RELIEVED	NO	NO	NO	NO	NO	NO			
	RADIOGRAPHED	SPOT	SPOT	SPOT	SPOT	SPOT	SPOT			
	VERTICAL OR HORIZONTAL	VERT	VERT	HORIZ	HORIZ	HORIZ	HORIZ			
	INSULATION THICKNESS									
	NO. OF SUPPORTS	9	7	6	5	4	3			
	LENGTH BEAM TO BEAM	111.0"	98.0"	111.0"	100.0"	110.0"	120.0"			
	LENGTH BASE SECTION									
	HEIGHT UP START	2.0'	1.0'	1.0'	1.0'	1.0'	1.0'			
	\$ TO BOTTOM OF SUPPORTS									
	TOTAL SHELL THICKNESS	5.25/815"	5.25/815"	5.25/815"	5.25/815"	5.00/815"	5.25/815"			
	HEAD THICKNESS	5.25/815"	5.25/815"	5.25/815"	5.25/815"	5.00/815"	5.25/815"			
	SHIRT THICKNESS	1.05/815"	1.05/815"	1.05/815"	1.05/815"	1.05/815"	1.05/815"			
	NO. REQ'D	5	2	1	1	1	1			
	SIZE	24"	24"	24"	24"	24"	24"			
	SERIES & FINISH	NONE	NONE	NONE	NONE	NONE	NONE			
	TYPE									
	THICKNESS									
	LOCATION									
	NO TRAYS REQ'D & SPACING	41 @ 24"	43 @ 24"	40 @ 24"						
	TYPE OF TRAY	TRAY	TRAY	TRAY	TRAY	TRAY	TRAY			
	NO OF CAPS / TRAY & TYPE	11 CA	11 CA	11 CA						
	SIZE OF CAPS	VALVE	VALVE	VALVE						
	SIZE OF ROVERS									
	TYPE OF DOWNCOMER									
	NO TRAYS REQ'D & SPACING									
	TYPE OF TRAY									
	TYPE OF HEAD									
	HEAD & HEADS									
	STRUCTURAL MEMBERS									
	REQUISITION NO.									
	PURCHASE ORDER									
	PURCHASED FROM									
	FABRICATION S/O									
	SHIPPING WEIGHT	112,776	14,455	13,000	48,000	19,500	4,500			
	TEST WEIGHT									
	NO. DATE									
	REVISION									
	BY									
	Sun Refining and Marketing Company									
	TOWER & VESSEL SCHEDULE									
	PLANT	CRAS PLANT								
	PLANT NO.									
	UNIT NO.									
	LOCATION									
	DRAWING NO.									
	REV.									
	SHEET									
	DATE	4/23/71								
										3

PUMP ITEM NO	SERVICE	P1.1A	P2.2A	P3.3A	P4.4A	P5.5A	P6.6A	P7.7A	NOTES	
									ABSORBER REFLUX	STRIPPER REFLUX
1	LIQUID	LEAN OIL	RKN OIL	RICH OIL	BUTANE	BUTANE	LEAN OIL	LEAN OIL		
2	PUMPING TEMPERATURE °F	100	100	278	134	126	411	280		
3	SPECIFIC GRAVITY AT P.1	7.318	6.822	5.781	5.770	5.188	5.807	4.693		
4	PIC SBU. CA CRIT. AT P.1	33 CP	33 CP	33 CP	33 CP	33 CP	33 CP	33 CP		
5	OPM AT P.1	717.3	1178.6	1079.2	621.3	227.1	881.5	32.3		
6	SUCTION PRESSURE (PSIA)	137	143	143	93	88	108	63		
7	DISCHARGE PRESSURE (PSIA)	204	193	157	159	205	206	184		
8	DIFFERENTIAL HEAD (FT)	246	193	37	277	321	402	314		
9	PSI/W AVAILABLE	12	13	1	12	12	12	118		
10	RPM									
11	RPM AT RATED QPM	55	65	77	40	33	85	3		
12	MAX. RPM FOR INSTALLED IMP	65	77	20	41	55	110	5		
13	MANUFACTURER									
14	TYPE AND SIZE									
15	STAGES									
16	SUCT. PLO. SIZE, RATINGS, FACE									
17	DISCH. PLO. SIZE, RATINGS, FACE									
18	NOZZLE ASSET SUCTON									
19	NOZZLE ASSET DISCHARGE									
20	IMPELLER DIA. REQUIRED									
21	MAX. IMP. DIA. AND HEAD									
22	IMP. EYE AREA									
23	RPM REQUIRED									
24	HYDROSTATIC TEST PRESSURE									
25	THRUST BEARING									
26	RADIAL BEARING									
27	COUPLING									
28	COUPLING GUARD									
29	ROTATION FROM COUPLING END									
30	MECHANICAL SEALS									
31	TYPE									
32	BOX									
33	WATER COOLED BEARING									
34	WEIGHT OF PUMP AND BASE									
35	CASTING									
36	SHAFT WEAR RINGS									
37	SHAFT PIECES									
38	ROTHINGS									
39	WEAR RINGS									
40	WEAR RINGS									
41	WEAR RINGS									
42	WEAR RINGS									
43	WEAR RINGS									
44	WEAR RINGS									
45	WEAR RINGS									
46	WEAR RINGS									
47	WEAR RINGS									
48	WEAR RINGS									
49	WEAR RINGS									
50	WEAR RINGS									
51	WEAR RINGS									
52	WEAR RINGS									
53	WEAR RINGS									
54	WEAR RINGS									
55	WEAR RINGS									
56	WEAR RINGS									
57	WEAR RINGS									

### Sun Refining and Marketing Company

EXTRAORDINARY PUMP SCHEDULE

PLANT: 3015 PLANT

LOCATION: 3015 PLANT

DATE: 4-1-81

DESIGNED BY: JRS

CHECKED BY: JRS

REVISED BY: JRS

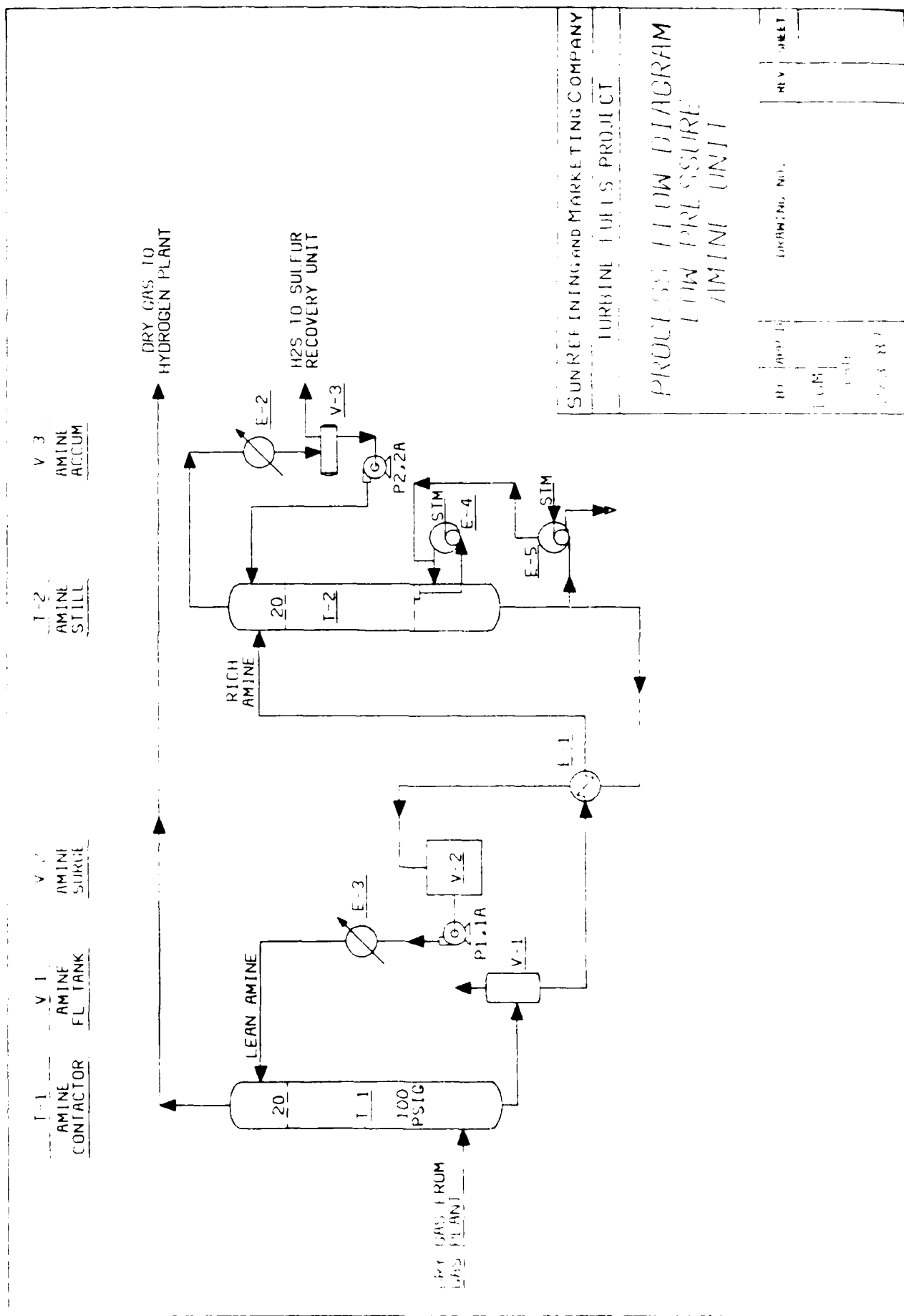
REVISED DATE: 4-1-81

REVISED SHEET: 4

**PROCESS DESIGN SPECIFICATIONS**

*for the*

**LOW PRESSURE AMINE UNIT**



- I-1 AMINE CONTACTOR
- V-1 AMINE FL. TANK
- V-2 AMINE SURGE
- I-2 AMINE STILL
- V-3 AMINE ACCUM

SUNREF INING AND MARKETING COMPANY  
TURBINE FUELS PROJECT

PROCESS FLOW DIAGRAM  
LOW PRESSURE  
AMINE UNIT

REV.	DATE	DESCRIPTION	BY	CHKD.	APP'D.
1					
2					
3					
4					
5					

## LOW PRESSURE AMINE UNIT

### Process Description

It is estimated that the combined dry gas from the gas plant plus bypassed lean gas will be about 48.4 million cubic feet per day containing about 1.22 gas volume percent hydrogen sulfide. This gas contains 25 short tons per stream day of sulfur. This must be removed to use this gas for making hydrogen. This is done by scrubbing the gas with 15 wt% monoethanolamine (MEA) in water. The gas is routed at 100°F and 100 psig through the amine absorber. The hydrogen sulfide is absorbed in the MEA. About a 99% removal is possible. The amine to the contactor top tray should be about 5°F above that of the incoming gas to avoid hydrocarbons condensing in the contactor. Condensation causes a "foam over" with amine carry over into the outlet gas.

The lean amine flows down through the 20-tray amine contactor countercurrent to the gas feed. The rich amine from the contactor is flashed to remove entrained hydrocarbons. The flashed amine is preheated by heat exchange and routed to the top of the amine still. Hot water vapor produced by the reboiler removes the hydrogen sulfide and carries it overhead. The water vapor is condensed in the overhead condenser and the hydrogen sulfide gas is piped to the sulfur recovery unit.

The stripped amine is cooled by heat exchange and water cooling for recirculation to the contactor.

A drag stream of lean amine is heated to a higher temperature to distill off the amine. Salts and impurities are withdrawn and discarded.

LOW PRESSURE AMINE UNIT

DESIGN BASIS

Feed Gas: 48.4 MMSCFD  
Hydrogen Sulfide fed 25 Short Tons/Stream Day  
Hydrogen Sulfide recovery 99%

Amine type: 15 wt% MEA (monoethanolamine)

Amine contactor pressure: 100 psig

Amine still bottoms: 235°F



LOW PRESSURE AMINE UNIT

Utilities and Chemicals

Steam

150 psig Steam	1,310 lb/hr
50 psig Steam	<u>12,500 lb/hr</u>
Steam condensate recovered	13,810 lb/hr

Cooling water

Supplied at 85°F, returned at 115°F

Circulation: 720 GPM

Electrical Power (440 and 220 volt) 23 KW

Chemicals

Monoethanolamine (MEA)  
at 100% concentration: 5 gal/day

LOW PRESSURE AMINE UNIT

List of Major Equipment

Exchangers

E-1	Rich Amine - Lean Amine Exchanger
E-2	Still Condenser
E-3	Lean Amine Cooler
E-4	Still Reboiler
E-5	Still Reclaimer

Towers and Vessels

T-1	Amine Contactor
T-2	Amine Still
V-1	Amine Flash Tank
V-2	Amine Surge Tank
V-3	Amine Accumulator

Pumps

P-1,1A	Lean Amine Pumps
P-2,2A	Still Reflux Pumps

**NOTES**  
 MR. STOKES RELEASED  
 IN - E RATED

NO DATE REVISION BY

**Sun Refining and Marketing Company**  
 HEAT EXCHANGER SCHEDULE

PLANT AIR FORCE  
 PLANT NO. MEASURE APARTS UNIT NO.

LOCATION: BY APPD DRAWING NO REV SHEET  
 DATE: 4/15/17

ITEM NO	SERVICE	MANUFACTURER	NO SHELLS	SIZE AND TYPE	SURFACE/UNIT	CONNECTED IN	FLUID CIRCULATED	QUANTITY	WEIGHT	REMARKS
1	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
2	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
3	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
4	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
5	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
6	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
7	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
8	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
9	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
10	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
11	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
12	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
13	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
14	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
15	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
16	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
17	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
18	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
19	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
20	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
21	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
22	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
23	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
24	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
25	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
26	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
27	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
28	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
29	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
30	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
31	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
32	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
33	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
34	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
35	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
36	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
37	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
38	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
39	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
40	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
41	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
42	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
43	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
44	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
45	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
46	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
47	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
48	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
49	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
50	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
51	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
52	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
53	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
54	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
55	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
56	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
57	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
58	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
59	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
60	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
61	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
62	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
63	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
64	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
65	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
66	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER
67	EXCHANGER	2	2	20" X 20"	1800	3	LEAN AMINE	2	1800	LEAN AMINE COOLER

WEIGHT EXCHANGER DRY NET  
 WEIGHT BUNDLE  
 CODE REQUIREMENTS  
 DRAWING NO  
 PURCHASE ORDER NO

NOTES

NO. DATE REVISION QT.  
 Sun Refining and Marketing Company  
 TOWER & VESSEL SCHEDULE

PLANT 4 N F 4006  
 LOCATION / AREA / UNIT NO  
 LOCATION  
 BY APPD DRAWING NO REV SHEET  
 DATE 4/15/63 2

ITEM NO.	T-1	T-2	V-1	V-2	V-3
	AMINE CONTRACT	AMINE STEEL	AMINE FL TANK	AMINE SURGE	AMINE ASSEMB
SERIAL NO.	275	275	275	275	275
ASSEMBLY	450	450	450	450	450
DETAILS	1254	1254	1254	1254	1254
TRAYS	265	265	265	265	265
PLATFORMS	765	765	765	765	765
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SHELL & HEADS					
SHIRT					
STRUCTURAL MEMBERS					
REQUISITION NO.					
PURCHASE ORDER					
PURCHASE FROM					
FABRICATOR S/O					
SHIPPING WEIGHT	1450				
TEST WEIGHT					

PUMP ITEM NO	SERVICE	DESCRIPTION	QTY	UNIT	REVISION	DATE	NOTES
1	LIQUID	PUMPING TEMPERATURE					<p>STILL</p> <p>RENEW</p> <p>WATER</p> <p>1200 F</p> <p>1.9</p> <p>1.8</p> <p>1.8</p> <p>1.8</p> <p>1.8</p> <p>1.8</p> <p>1.8</p> <p>1.8</p>
2		SPECIFIC GRAVITY AT P.T.					
3		WATER COR. COEFF. AT P.T.					
4		WATER COR. COEFF. AT P.T.					
5		WATER COR. COEFF. AT P.T.					
6		SUCTOR PRESSURE (PSID)					
7		DISCHARGE PRESSURE (PSID)					
8		DIFFERENTIAL HEAD (FT)					
9		NPSH AVAILABLE					
10		RPM					
11	GENERAL CONDITIONS	MAX. AT WATER GPM					
12		MAX. AT WATER GPM					
13		MAX. AT WATER GPM					
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56		MAX. AT WATER GPM					
57		MAX. AT WATER GPM					

Sun Refining and Marketing Company

CENTRIFUGAL PUMP SCHEDULE

PLANT: *91A FORCE*

COM: *FAB 3308* UNIT NO: *M/1116*

DATE: 4/13/47

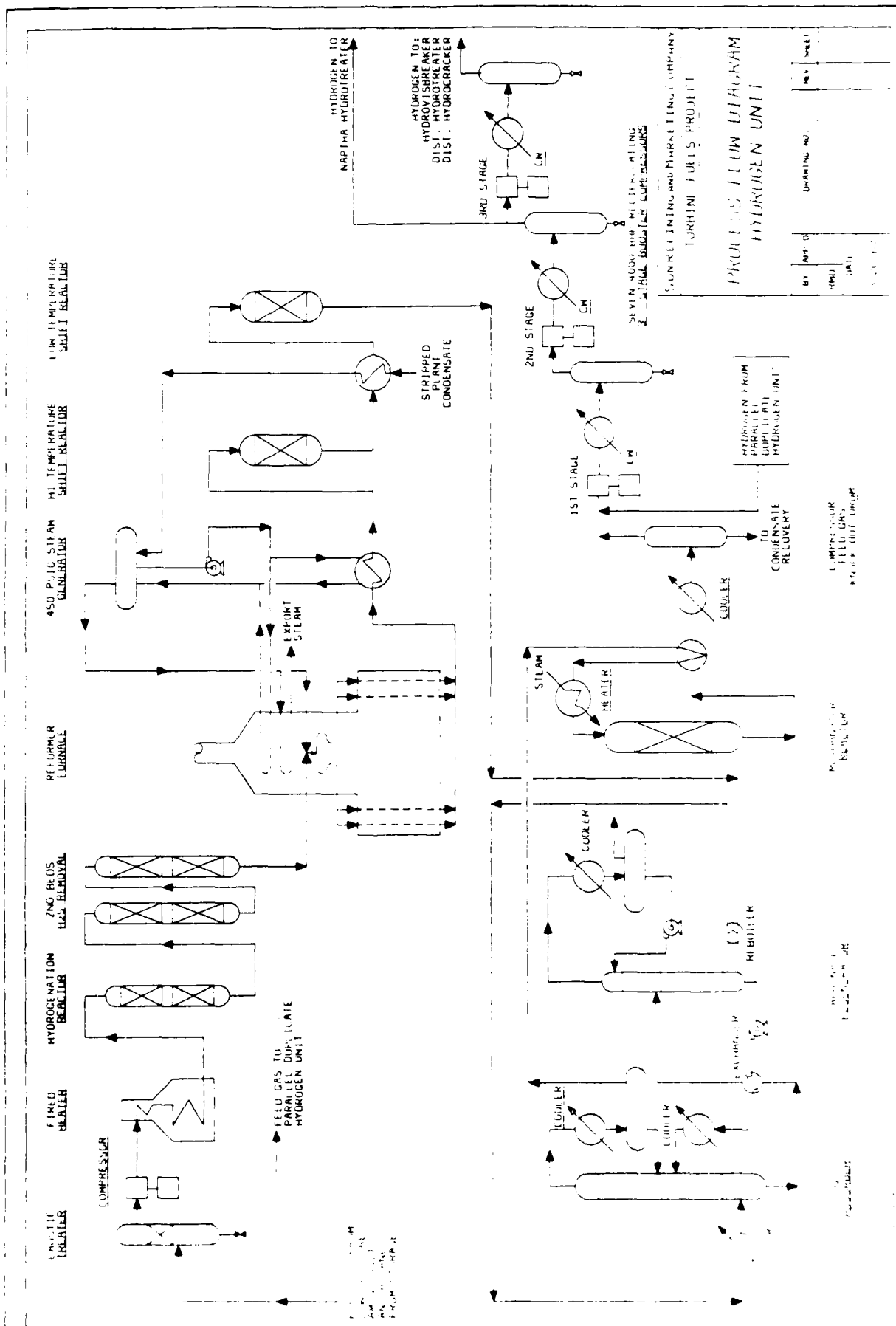
BY: *APPD* DRAWING NO. REV. SHEET

3

**PROCESS DESIGN SPECIFICATIONS**

for the

**HYDROGEN PLANT**



CONKRE ENGINEERING AND MARKETING COMPANY  
 TURBINE FUELS PROJECT  
 PROCESS FLOW DIAGRAM  
 HYDROGEN UNIT  
 SHEET NO. 1  
 DRAWING NO. 100-100-100-100

## HYDROGEN PLANTS

### Catalytic Steam Reforming of Hydrocarbons

#### PROCESS DESCRIPTION

The overall hydrogen production requirement for the refinery conversion processes is 160 MMSCFD of 95% purity hydrogen. This has been divided into two duplicate 80 MMSCFD hydrogen plants since this is approaching the maximum shop-fabricated equipment size. Also, the conversion processes - hydrovisbreaking, naphtha hydrotreating, distillate hydrocracking, and hydrocracking - can be kept on-stream if one hydrogen unit fails.

The feed to the units is amine-treated dry gas plus vaporized butane. These gases are caustic washed to remove residual hydrogen sulfide.

The feed gas is then compressed from 100 psig to 400 psig and preheated with a fired furnace to about 700°F. The preheated gas is then mixed with hydrogen and passed over cobalt-moly catalyst, which saturates olefins and converts any organic sulfur compounds to saturated hydrocarbons plus hydrogen sulfide.

The feed gas is then routed through two zinc oxide beds in series, which remove the small amount of hydrogen sulfide formed from organic sulfur compounds. The feed gas is then mixed with preheated steam and is fed to the reformer furnace to produce hydrogen plus carbon monoxide and carbon dioxide. The zinc oxide is converted to zinc sulfide. The zinc oxide is replaced periodically.

The reformer furnace contains vertical tubes with an inside diameter of approximately six inches packed with catalyst. The feed gas enters through the convection section after being premixed with superheated



steam. The steam is waste steam produced in the process which has also been preheated in the furnace convection section. The steam and hydrocarbon gas flow down over the catalyst in the radiant tubes where the conversion to hydrogen, carbon dioxide, carbon monoxide, and water vapor takes place. The converted gases leave the furnace at about 1525°F and are cooled to 650°F in the waste heat steam generator.

The gases are then routed through a high temperature (750°F) catalytic shift converter followed by a low temperature (400°F) catalytic shift converter. These catalytic reactors convert the carbon monoxide to carbon dioxide. The carbon dioxide is then absorbed in a regenerative potassium carbonate solution and thus removed from the process.

The carbon dioxide absorber tower operates at about 200°F and 300 psig. The rich solution is stripped of the absorbed carbon dioxide in a reboiled regenerative tower. The carbon dioxide is vented to the atmosphere. The scrubbed gas is now about 95% hydrogen but still contains some carbon dioxide and carbon monoxide, which are next removed by a catalytic methanator.

The catalytic methanator reacts carbon oxides with hydrogen to form methane plus water vapor. This removes the carbon monoxide which is hydrocracker catalyst poison. The methanator operates at about 500°F and about 300 psig.

The purified hydrogen product is then cooled and compressed from 240 psig to 2445 psig with electrically driven reciprocating compressors for use in the Hydrovisbreaker, Distillate Hydrotreater, and Distillate Hydrocracker Units. A portion of the hydrogen is removed at 1190 psig for use in the Naphtha Hydrotreater Unit.

# HYDROGEN PLANT

## DESIGN BASIS

Two (2) duplicate units

Feed (combined total for both units):

Refinery Dry Gas	48.42 MMSCFD	(containing 23.64 MMSCFD Hydrogen)
Vaporized Butane	1158 BPSD	
Steam consumed	146,000 lb/hr	

Steam export (estimated)      0 - 50,000 lb/hr

The detailed steam balance within the unit was estimated from similar units. A detailed hydrogen plant design was not developed.

Total Hydrogen Product (95% purity):      160 MMSCFH

### Hydrogen Required by the Hydroprocessing Plants

Naphtha Hydrotreater	9.0	MMSCFD
Hydrovisbreaker	40.0	
Distillate Hydrotreater	55.0	
Distillate Hydrocracker	52.0	
Internal use at Hydrogen Plt.	<u>4.0</u>	
Total	160.0	MMSCFD

## HYDROGEN PLANT

### Utilities, Chemicals, and Catalyst

#### Combined Requirements for both 80,000 MMSCFD plants

Export Steam Generated (450 psig):	0 - 50,000 lb/hr
Cooling water circulated	38,720 GPM
Electrical power	
Excluding product booster compressors	2,700 KW
Product booster compressors	<u>19,206 KW</u>
Total	21,906 KW

Process water 838 GPM

Fuel (from refinery butane  
and naphtha products)

Butane 917.5 MMBTU/HR = 5,743 BPSD

Naphtha 371.5 MMBTU/HR = 2,111 BPSD

#### Raw Materials

Caustic Soda	1000 lb/day
Zinc Oxide	4.2 lb/day
Cobalt Moly Desulfurization Catalyst	61.2 lb/day
Reforming Catalyst	86.4 lb/day
High Temperature Shift Catalyst	252 lb/day
Low Temperature Shift Catalyst	288 lb/day
Potassium Carbonate	1120 lb/day
Methanation Catalyst	33 lb/day

## HYDROGEN PLANT

### Major Equipment

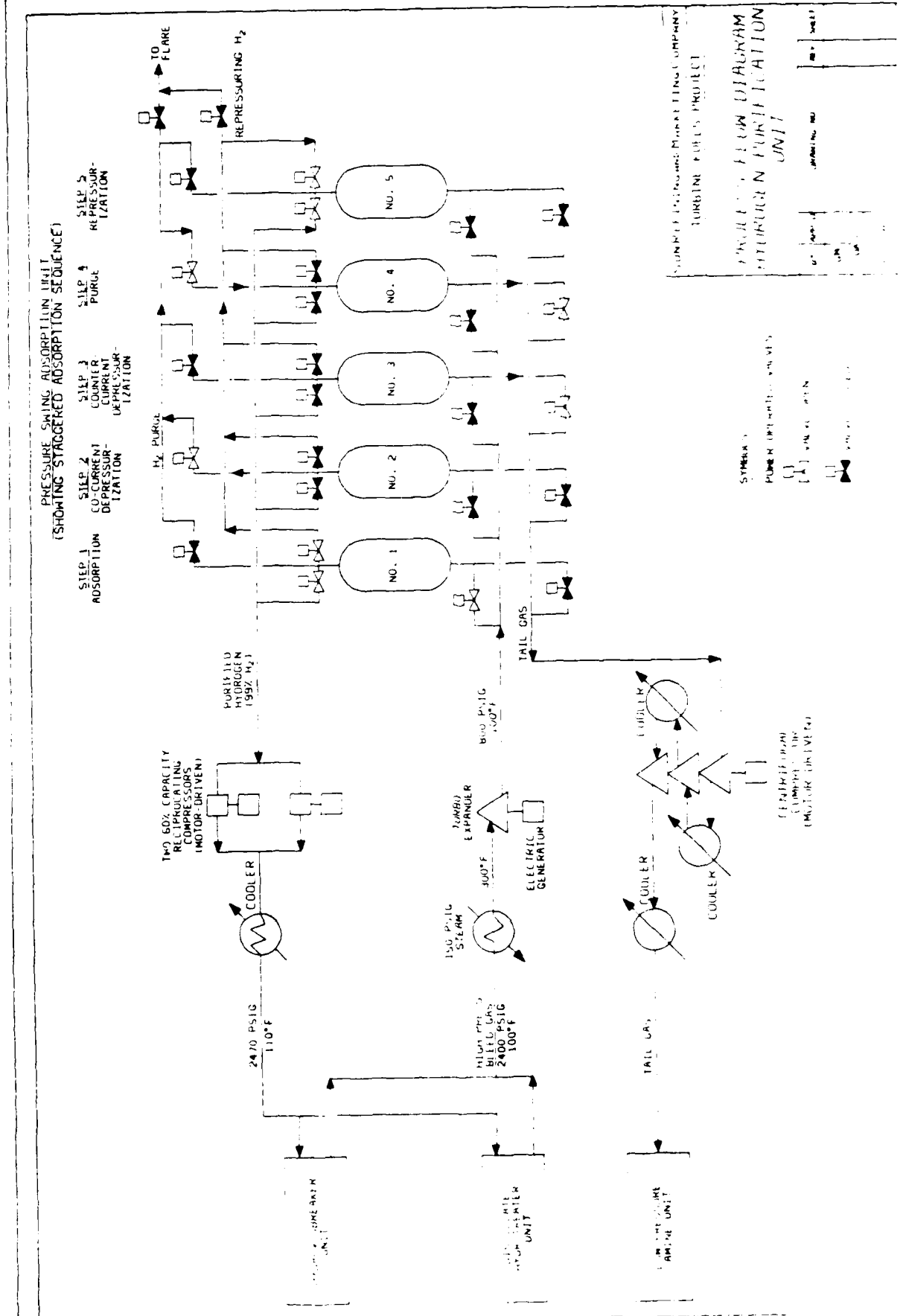
The hydrogen plant capital and utility requirements were based on data taken from Stanford Research Institute reports and from Sun Company hydrogen plant designs. Therefore, no detailed design or major equipment list was developed.

Additionally, no royalty is involved in purchasing the hydrogen plants.

**PROCESS DESIGN SPECIFICATIONS**

**for the**

**HYDROGEN PURIFICATION UNIT**



## HYDROGEN PURIFICATION UNIT

### Pressure Swing Absorption plus Recompression

#### Process Description

In hydroprocessing oil refineries hydrogen of 95 mole percent purity or higher is injected into the reactors. In the reaction methane and other light hydrocarbons are produced as byproducts. These are absorbed into the liquid product and removed from the system. In addition in hydrovisbreaking a bleed of the high pressure hydrogen recycle is required to maintain hydrogen purity (and hydrogen partial pressure) in the recycled gas system.

Recovering hydrogen from the high pressure bleed gas by use of pressure swing absorption is an economic method to reduce the size of the hydrogen production plant. This process developed by Union Carbide has been demonstrated world-wide in over 240 commercial units, and is available royalty-free.

Shown in the attached process flow diagram, it was found desirable to recover hydrogen from the distillate hydrotreater as well as the hydrovisbreaker. These bleed gases are combined at a pressure of 2400 psig and a temperature of 100°F. They must be reduced in pressure from 2400 psig to 800 psig for Pressure Swing Absorption (PSA) processing. A turbo-expander is used to recover valuable energy. Preheating ahead of expansion is required to obtain an expander outlet temperature of 100°F.

Hydrocarbons are absorbed on a molecular sieve in the PSA unit. The high purity hydrogen passes through the PSA, is recompressed and reused in the hydrotreater and hydrovisbreaker units. The PSA system is installed as a 5-vessel unit. While one vessel is absorbing hydrocarbons, four vessels are undergoing staged regeneration. A cycle is arranged so one vessel in a unit is always onstream.

Normal practice is to install two or more 5-vessel units in parallel depending on the capacity required. The cyclic sequence of a 5-vessel unit is:

<u>Vessel Number</u>	<u>Operating Status</u>	<u>Process Description</u>
1	On stream	<u>Absorption of hydrocarbons.</u> Upflow feed.
2	Off line	<u>Depressuring</u> - top vent - venting hydrogen to repressure vessel No. 5, followed by purge hydrogen to vessel No. 4.
3	Off line	<u>Final Depressuring</u> - bottom vent - venting absorbed hydrocarbon gases to tail gas at 0 psig and 110°F.
4	Off line	Purge hydrogen from vessel No. 2 into top of vessel. The purge sweeps hydrocarbon gases out bottom vent to tail gas at 0 psig and 110°F.
5	Off line	High purity, medium pressure hydrogen from vessel No. 2 is injected into the vessel followed by high pressure pure hydrogen. This repressures vessel No. 5, which then goes on stream as vessel No. 1 goes off line.

The cycle continues controlled automatically.

The hydrogen product at 99 mole percent purity is compressed from 800 psig to 2475 psig, cooled to 110°F and returned to the Hydrovisbreaker and Distillate Hydrotreating Unit.

The tail gas is compressed from 0 psig to 165 psig, cooled and routed to the low pressure amine unit for removal of hydrogen sulfide.



HYDROGEN PURIFICATION UNIT  
DESIGN BASIS

Purification Process: Polybed Pressure Swing Absorption Unit

	<u>Feed Gas (1)</u>	<u>Hydrogen Product Gas (2)</u>	<u>Fuel Gas (3)</u>
Flow Rate, MMSCFD	96.5	75.0	21.5
Pressure, psig	800	790	2
Temperature, °F	100	110	110

Composition, mol%

Hydrogen	87.85	99 minimum	49.1
Methane	9.36	1 maximum	38.4
Ethane	1.54	-	6.9
Propylene	0.05	-	0.2
Propane	0.69	-	3.1
Butane	0.29	-	1.3
Pentane	0.18	-	0.8
Water	<u>0.04</u>	<u>-</u>	<u>0.2</u>
Totals	100.00	100	100.0

Notes:

- (1) Feed gas enters the area at 2475 psig and 110°F. It is preheated to 300°F then expanded to 800 psig and 100°F.
- (2) Hydrogen product gas is compressed from 790 psig to 2470 psig and cooled to 110°F.
- (3) Fuel gas is compressed from 0 psig to 165 psig and cooled to 110°F.

## HYDROGEN PURIFICATION UNIT

### Utility and Chemical Requirements

Steam Used: 150 psig to Feed Gas Preheater 19,200 lb/hr

Instrument Air: 3500 SCFH at 100 psig

#### Electrical Power:

Power used by hydrogen compressors (Two 4000 BHP motors)	7026 BHP
Power used by fuel gas compressor	<u>3750</u> BHP
Total power required for compressors	10,776 BHP
Power generated by Turbo-Expander	<u>9,714</u> BHP
Net power required for compressors	1,062 BHP
Power for instrumentation, lighting, etc.	10 KW

#### Cooling Water

Supply temperature: 85°F

Return temperature: 105°F

Hydrogen Compressor Inter- and Aftercoolers	1,348 GPM
Fuel Gas Compressor Inter- and Aftercoolers	<u>319</u> GPM
Total	1,667 GPM

## HYDROGEN PURIFICATION UNIT

### Major Equipment

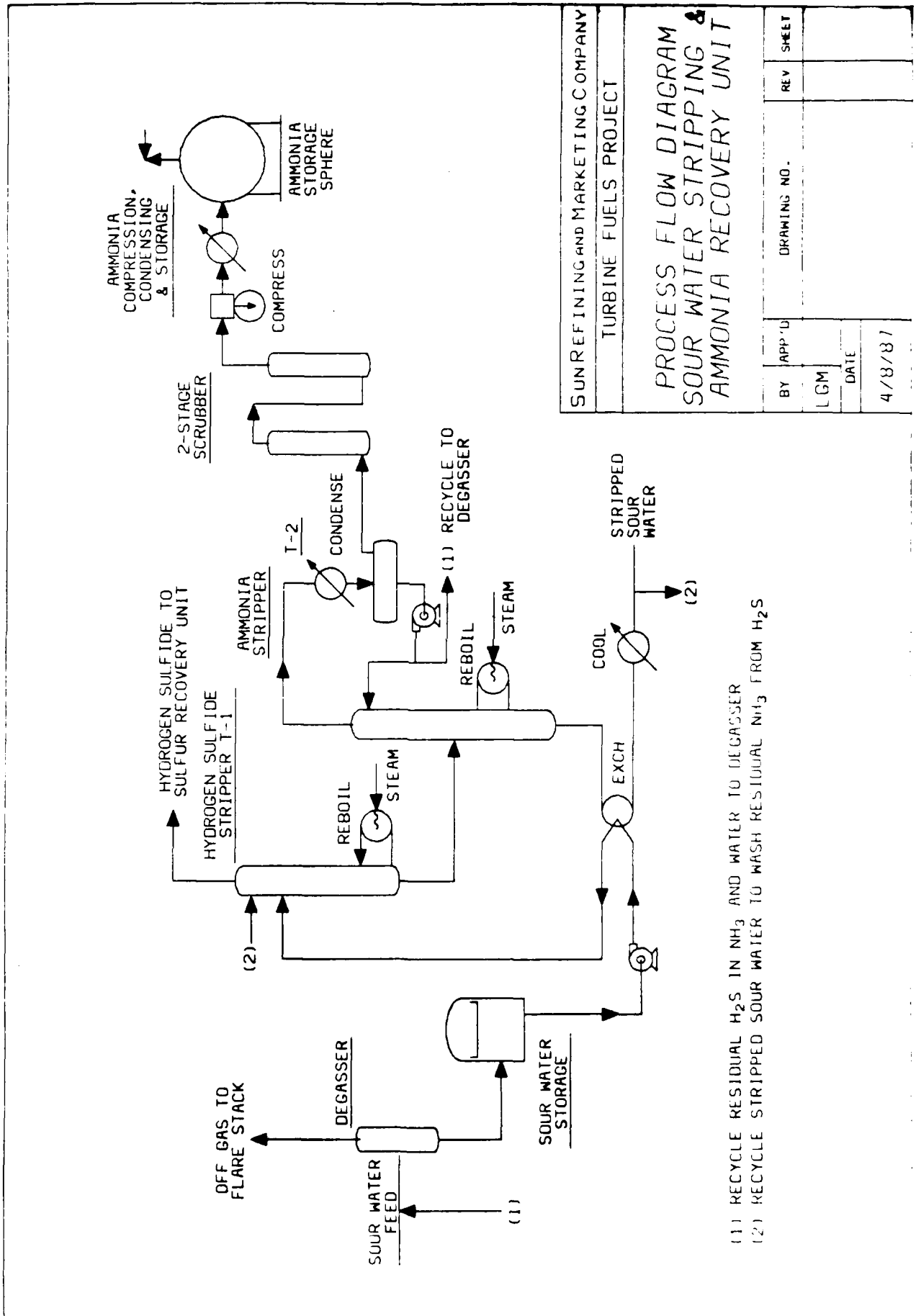
Because a package price for the Polybed Pressure Swing Adsorption Unit was obtained from Union Carbide, it was unnecessary to specify the equipment in detail. However, the turbo-expander feeding the unit and the compressor systems for the hydrogen product and tail gas were specified separately. The Hydrogen and Fuel Gas Compressor systems were priced by Dresser (equipment supplier). Sun estimated the horsepower and price of the Turbo-Expander.

Additionally, Union Carbide advised that the life of the molecular sieves in the pressure swing adsorption unit is sufficiently long that a replacement batch is not required. Their plant price included the cost of the molecular sieves. There is no royalty charge.

**PROCESS DESIGN SPECIFICATIONS**

**for the**

**SOUR WATER STRIPPER AND AMMONIA PLANT**



## SOUR WATER STRIPPER AND AMMONIA RECOVERY PLANT

### PROCESS DESCRIPTION

Approximately 1750 GPM of sour water is recovered from the hydroprocessing units. This is largely wash water used to eliminate condenser and cooler fouling in the reactor recycle loops. Extensive investigations in the petroleum industry show that the ammonium hydrosulfide concentration in water must be held to 2 wt% or less to eliminate equipment fouling. It is estimated that up to 100 short tons/day of ammonia and 143 short tons/day of hydrogen sulfide will be contained in the water.

The ammonia combines with an equal molar volume of hydrogen sulfide in the sour water. The above quantities occur at 49% nitrogen conversion to ammonia in the Hydrovisbreaker which was used as maximum design conversion.

Two stage sour water stripping is required to recover the ammonia separately from the hydrogen sulfide. Typically, caustic is added and the ammonia is steam stripped off. The alkaline water is then neutralized with acid and the hydrogen sulfide is steam stripped off. The resulting stripped sour water contains appreciable salts and is a disposal problem.

Two-stage stripping, separating ammonia from hydrogen sulfide is highly desirable for proper operation of the sulfur recovery unit. Also the ammonia has a worthwhile product value.

Chevron Research Company has developed a licensed process (WWT Process) which does not require the addition of caustic and acid to separately recover the ammonia and hydrogen sulfide by steam stripping. This permits re-using the stripped water for hydroprocessing injection, crude unit desalting water, and other process uses. A drag stream must be routed to disposal via waste water treating to oxidize the phenols and a small residual ammonia and hydrogen sulfide content.

The WWT Process is included in this plant since it reduces water usage and water disposal. The process consists of four main process steps:

1. Degassing and feed storage
2. Acid gas ( $H_2S$ ) steam stripping
3. Ammonia steam stripping
4. Ammonia purification and liquefaction

The sour water feed to the plant is combined with a recycle stream from the ammonia stripper, cooled and fed to a degasser where dissolved hydrogen, methane, and other light hydrocarbons are removed. The recycle stream is rich in ammonia, which helps keep acid gases in solution in the degasser. Thus, the release of acid gas and possible air pollution are minimized. The degassed sour water is pumped to an off-plot storage tank which serves to dampen flow rate and composition changes. It also provides the opportunity to remove entrained oil and solids.

From the feed tank the degassed sour water feed is pumped to the WWT unit, where it is heated by feed-bottoms exchange and fed to the acid gas or hydrogen sulfide stripper. This stripper is a steam-reboiled distillation column. The hydrogen sulfide, which is stripped overhead, is of high purity - an excellent feed for a sulfur or sulfuric acid plant. It contains negligible ammonia, less than 50 ppm, and very little hydrocarbon since the plant feed has been degassed. However, it does contain any carbon dioxide that is present in the feed. The hydrogen sulfide is available at about 20 psig and 100°F and will be saturated with water vapor.

The hydrogen sulfide stripper bottoms, containing all the ammonia in the feed and some hydrogen sulfide, is fed directly to the ammonia stripper, which is a steam-reboiled refluxed distillation column. In this column, essentially all the ammonia and hydrogen sulfide are removed from the water, which leaves as the column bottoms stream. After exchanging heat with the hydrogen sulfide stripper feed, this stripped water is cooled and sent off-plot for reuse or

treating. The stripped water contains less than 50 ppm of free ammonia and less than 10 ppm of free hydrogen sulfide. The stripped water will also contain traces of phenols and salts which entered with the feed. If the feed contains acidic compounds that "fix" ammonia, the fixed ammonia can be released and then stripped off.

The ammonia and hydrogen sulfide stripped from the water in the ammonia stripper are passed through an overhead condenser and are partially condensed. The liquid is used as column reflux, with a portion being recycled to the degasser and off-plot feed tank.

For production of anhydrous ammonia, the gas is passed through a two-stage scrubbing system to remove hydrogen sulfide and is then liquified to produce the anhydrous ammonia. The hydrogen sulfide content of the ammonia is typically less than 5 ppm.

The ammonia is compressed to 225 psia and condensed with cooling water. The ammonia product liquid is stored in a 42 foot diameter sphere which will hold 700 short tons (7 days of production).



SOUR WATER STRIPPER AND AMMONIA PLANT  
DESIGN BASIS

Feed: Maximum design

1750 GPM Sour Water at 100°F and 130 psig  
containing:

Ammonia	100.4 Short tons/stream day
Hydrogen sulfide	142.6 Short tons/stream day

Trace amounts of phenols, chloride, and salts

A pH of 8.5 to 9.0 is expected due to an excess of ammonia.

Hydrogen Sulfide Product

Hydrogen sulfide vapor containing 100 wt ppm ammonia, saturated with water vapor. The product is available at 20 psig and 100°F. This is feed to the Sulfur Recovery Unit.

Ammonia Product

Liquid ammonia (anhydrous) containing 5 wt ppm hydrogen sulfide and  
0.4 wt ppm of water.

Stripped Water

The water will contain 50 wt ppm of ammonia (maximum) and 10 wt ppm of hydrogen sulfide (maximum). Small amounts of phenols, chlorides, and salts will be present.

SOUR WATER STRIPPER AND AMMONIA PLANT  
UTILITY AND CHEMICAL REQUIREMENTS

Utilities

150 psig Steam used:	90,000 lb/hr
50 psig Steam used:	118,000 lb/hr
Steam condensate recovered:	208,000 lb/hr
Cold condensate used:	26,000 lb/hr

Electrical power used (440 and 220 Volt)	980 KW
--	--------

Cooling water circulated	1,720 GPM
Supplied at 85°F	
Returned at 100°F	

Chemicals and catalyst

None

SOUR WATER STRIPPER AND AMMONIA PLANT  
MAJOR EQUIPMENT

Because this is a licensed process, a detailed design and equipment list were not developed. These would be supplied by Chevron Research when needed.

Royalty cost basis

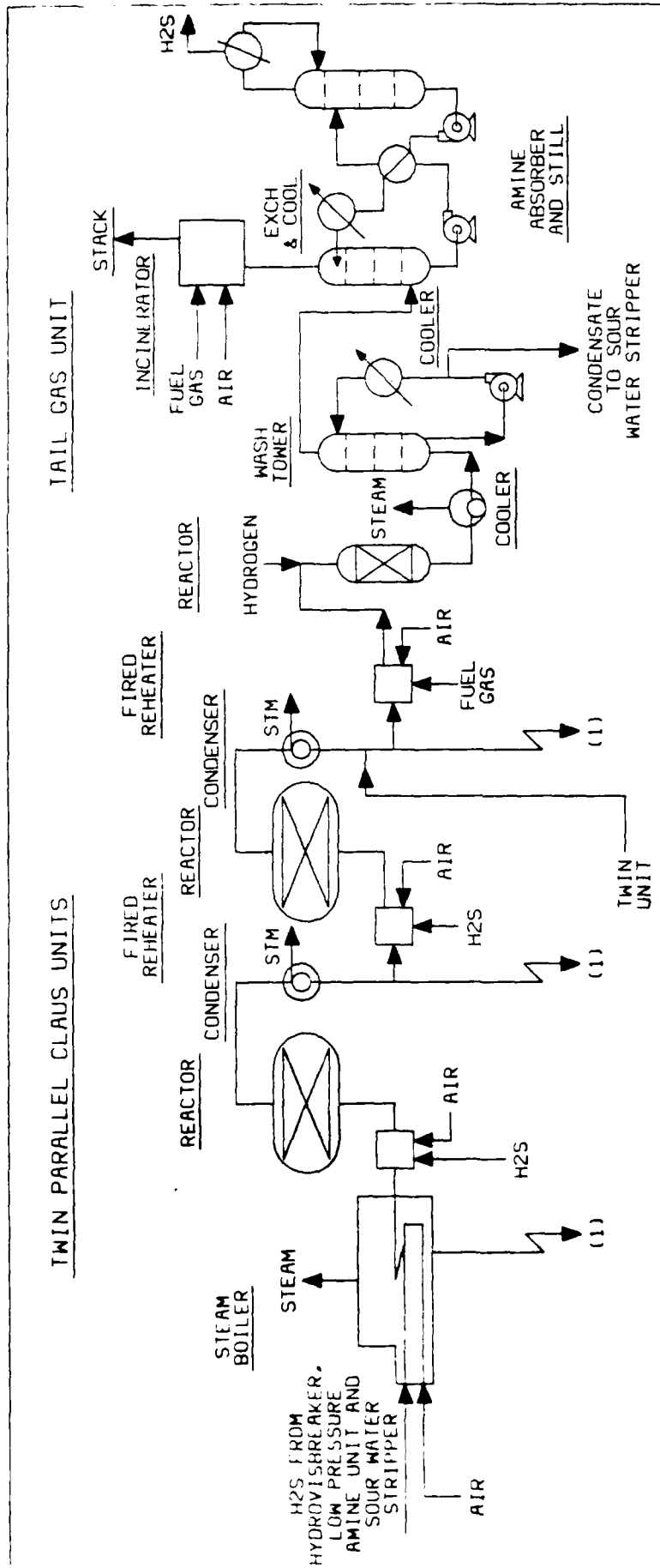
Fixed royalty - \$400,000 paid over 2 years

Running royalty - 87,500 per year (This would vary slightly with actual H<sub>2</sub>S and ammonia recovery.)

**PROCESS DESIGN SPECIFICATIONS**

for the

**SULFUR RECOVERY UNIT**



SUNREFINING AND MARKETING COMPANY  
TURBINE FUELS PROJECT

PROCESS FLOW DIAGRAM  
SULFUR RECOVERY UNIT

BY	APP'D	REV	SHEET
LCM			
	DATE		
	4/7/87		

(1) MOLTEN SULFUR DRAINS TO PIT

## SULFUR RECOVERY UNIT

including a  
Claus Unit and Tail Gas Unit

### Process Description

The Claus process is the most widely used method to recover sulfur from hydrogen sulfide ( $H_2S$ ). Conversion is achieved by combustion of one-third of the hydrogen sulfide to sulfur dioxide ( $SO_2$ ). Since the reaction is exothermic, it is carried out in a waste heat steam generator using air for combustion. The  $H_2S$  and  $SO_2$  then react over a catalyst to form sulfur and water vapor. Since the conversion is incomplete the gases are cooled, the sulfur is condensed out, and the reheated gases are fed to a second stage reactor. The gases are again cooled indirectly by low pressure steam generation, and the sulfur is condensed out. The Claus reaction takes place at about  $220^\circ C$  ( $428^\circ F$ ) and 3 to 8 psig. Alumina or bauxite is the normal catalyst for both reactors.

Since some hydrocarbon carryover and other problems with sulfur plants tend to plug the catalyst, it is normal practice to install two 100% capacity units to ensure 100% onstream time.

The twin units are followed by a single tail gas unit to ensure about 99.8% total hydrogen sulfide removal. The tail gas unit reheats the tail gas and introduces hydrogen as a reducing gas to convert all the sulfur compounds in the tail gas to hydrogen sulfide. The tail gas is then cooled, and the hydrogen sulfide is scrubbed out with amine. The amine is recycled to a still to strip out and recover the hydrogen sulfide. The conversion of sulfur compounds plus hydrogen to hydrogen sulfide takes place in a reactor containing a cobalt/molybdenum catalyst at about  $300^\circ C$  ( $572^\circ F$ ) and essentially atmospheric pressure.

The off-gas remaining after the tail gas has been scrubbed by amine is routed to a direct fired incinerator to combust any residual hydrogen sulfide or hydrocarbon gases.

These incinerated gases are then dispersed from a stack. The percentage sulfur removal from the tail gas can be maintained at 99.8% of the  $H_2S$  in. However, good instrumental control and maintenance of the unit is necessary.

## SULFUR RECOVERY UNIT

### Design Basis

Feed:	<u>Sulfur Equivalent</u>
Hydrogen sulfide from amine units	87.9 short tons/day
Hydrogen sulfide from sour water stripper	<u>77.3 short tons/day</u>
Total	165.2 short tons/day

The unit must be designed for a major amount of the hydrogen sulfide (about 86%) coming from the Sour Water Stripper when the Hydrovisbreaker operates at maximum nitrogen conversion. This is because the resulting ammonia in the Hydrovisbreaker wash water will absorb most of the hydrogen sulfide. At that time the Hydrovisbreaker amine unit duty will be light.

Product: Molten Sulfur      165 short tons/day = 330,000 lb/day

Two parallel Claus sulfur plants each designed for 165 short tons/day of sulfur production followed by one BSR/MDEA or Scott Tail Gas Unit plus an incinerator and stack.



SULFUR RECOVERY UNIT  
Utilities and Chemical Requirement

Utilities Produced

Excess 600 psig Steam at 600°F	8,400 lb/hr
Excess 50 psig Steam	28,900 lb/hr
Excess condensate produced	37,300 lb/hr

Utilities Required

Electrical power	192 KW
Cooling water circulation	3,390 GPM
Treated boiler feed water	45,841 lb/hr
Natural gas	21.4 MMBTU/hr

Catalyst and chemical cost \$32,430 per year

Includes: alumina Claus catalyst  
Co-Mo tail gas catalyst  
and alkanolamine

## SULFUR RECOVERY UNIT

### Major Equipment

A packaged price was obtained for both the Claus units and the tail gas unit from Ralph M. Parsons Company, a builder of many sulfur plants. Consequently it was unnecessary to develop a detailed process design or major equipment list.

There is no royalty on the Claus units, but there is a royalty on the tail gas unit. The royalty on the BSR/MDEA tail gas unit would be:

Tail Gas Unit Royalty, Ralph M. Parsons Co.  
for 165 tons/day sulfur production:

\$172,000 distributed as:	25% upon signing the license
	25% upon start of construction
	25% upon mechanical construction
	25% after performance guarantee is met

**PROCESS DESIGN SPECIFICATIONS**

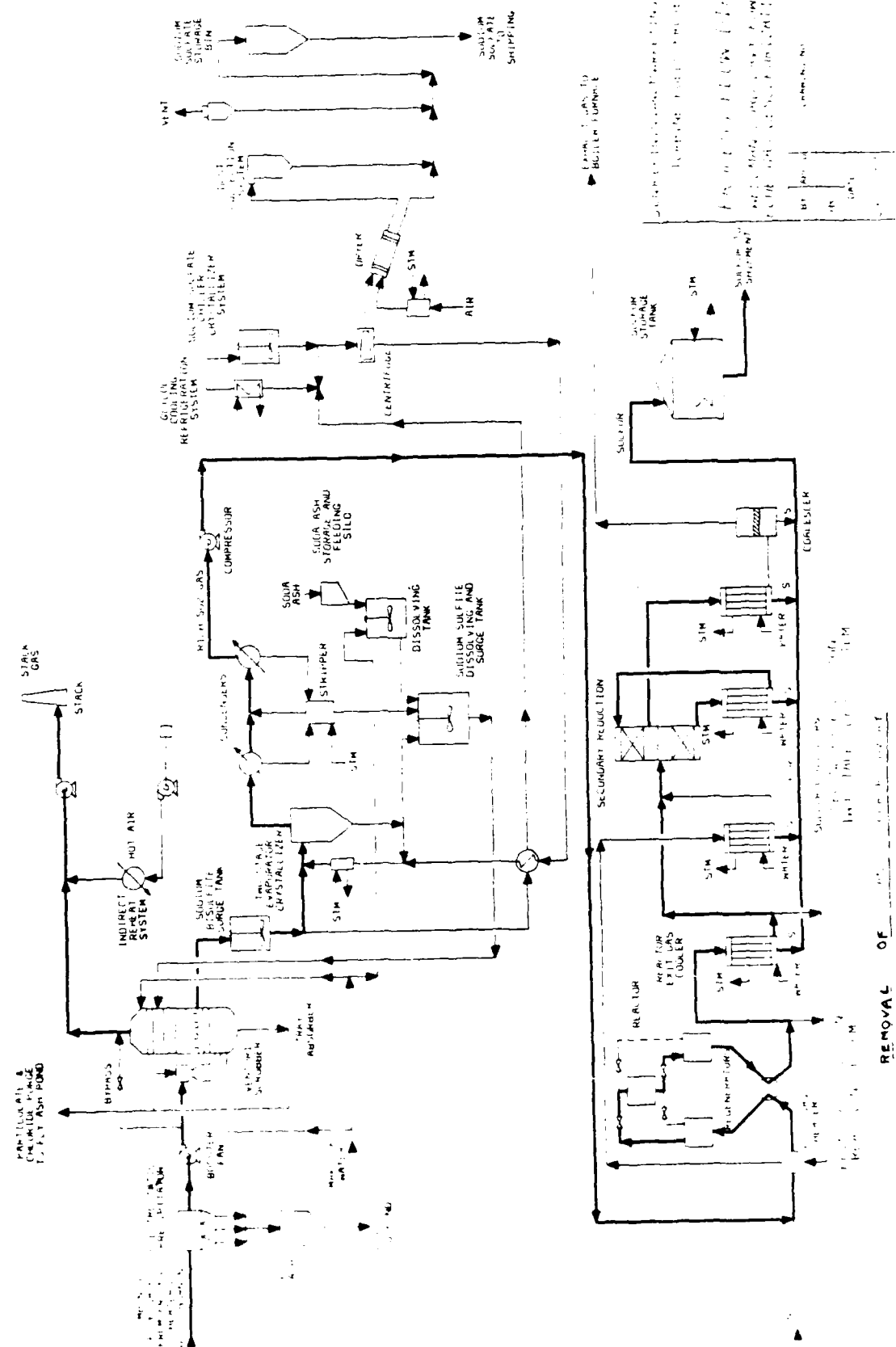
for the

**FLUE GAS DESULFURIZATION UNIT**

**SODIUM SULFATE CRYSTALLIZATION AND REMOVAL**

**SO<sub>2</sub> REGENERATION**

**GAS CLEARING AND ABSORPTION**



## Flue Gas Desulfurization Unit

### Process Description

The vacuum residuum produced from hydrovisbreaking will contain 1.34 weight percent sulfur based on the pilot plant data. There will be 4839 BPSD of this residuum produced. At maximum firing rate on the feed heaters and the boiler, 3862 BPSD of this residuum will be used as fuel. This will send an excessive amount of sulfur dioxide to the atmosphere (19.7 tons per day), unless stack gas scrubbing is installed. This fuel will also contain 612 lb/hr of molybdenum sulfide (MoS) derived from the Hydrovisbreaker coke-suppressing additive, which will become fly ash after fuel combustion. This must be removed by an electrostatic precipitator (ESP) ahead of the stack gas scrubber.

The Wellman-Lord stack gas scrubbing process was selected since this process has more proven commercial operation than any other regenerative process. Since it is a regenerative process it recovers sulfur and eliminates a large spent treating agent disposal problem which non-regenerative processes have. Also the make-up treating agent (soda ash) is a common low cost chemical.

Information from the Stanford Research Institute was used to develop the following information. A detailed design was not developed for the San Ardo crude stack gas scrubber.

A process flow diagram of the Wellman-Lord sulfite scrubbing process for removing  $\text{SO}_2$  from flue gases is attached. The scrubbing liquor is regenerated for recycling and the liberated rich  $\text{SO}_2$  gas stream is reduced to sulfur by methane using a proprietary process of Allied Chemical Corporation.

The flue gas enters the system at 450°F and is cooled to 200°F by exchanging heat with the effluent flue gas. The flue gas enters the

electrostatic precipitator where 99.5 to 99.8 wt% of the fly ash is removed and then proceeds through a booster fan to the scrubber system. The gas enters a venturi prescrubber at 200°F where the gas is humidified and cooled to about 130-135°F with a recirculating aqueous stream which also removes 95 to 99% of the  $\text{SO}_3$  and chlorides in the flue gas as hydrochloric and sulfuric acids. Any remaining particulate in the gas is substantially removed in this step. The removal of chlorides in this step keeps chlorides in the subsequent sulfite scrubbing at a very low level, thus reducing the potential occurrence of stress corrosion. Make-up process water is added at this point to the bottom of the absorber to replace water lost by vaporization in humidifying and cooling the gas and water removed in the purge stream from the venturi scrubber. This aqueous purge stream containing solids and acids collected by the venturi scrubber are pumped to an ash disposal pond where an alkaline reaction with the ash neutralizes the acidic constituents. The stream may alternately be neutralized with an excess of slaked lime slurry if necessary before disposal in the ash pond. The saturation of the flue gas in the venturi prescrubber system prevents the evaporation of any significant amounts of water in the  $\text{SO}_2$  scrubber itself.

The flue gas enters the bottom of the absorber almost saturated with water, and then passes through a chevron type demister to catch carryover of any entrained recycle wash solution into the main body of the absorber. The cooled humidified flue gas enters the scrubber through chimney type distributors where it flows upward through three valve trays. The lean sodium sulfite solution enters at the top of the scrubber and flows downward counter-currently to the rising flue gas from tray to tray, removing over 90% of the  $\text{SO}_2$  and forming sodium bisulfite in the liquor. The sodium sulfite solution has the capacity to absorb  $\text{SO}_2$  up to about 60 g/liter. The superficial gas velocity through the absorber is about 10 fps. Rich scrubber solution is collected on the bottom tray and flows to a surge tank of a capacity to allow the regenerator circuit to be shut down for up to 24 hours without halting scrubber operation.

Oxidation of the sodium sulfite in the scrubber solution occurs by reaction with the oxygen in the flue gas and inactive sodium sulfate is formed. About 8 to 10% of the sulfur absorbed from the flue gas is converted to  $\text{Na}_2\text{SO}_4$  by typical levels of excess oxygen in the flue gas. To keep this level of unreactive salt from building up to undesirable levels and causing crystallization in the scrubber, a purge of about 15% of the rich scrubber liquor is withdrawn from treatment to separate  $\text{Na}_2\text{SO}_4$  from the system and return the active solution to the scrubber. The warm purge stream is first cooled by exchange with cold returning scrubber solution and then cooled further to about 30-32°F in a chiller-crystallizer. A mixture of sodium sulfate and sodium sulfite crystallizes out which is separated from the liquor in a centrifuge at about 40% solids. With controlled crystallization,  $\text{Na}_2\text{SO}_4$  precipitates in a much greater proportion than the other sodium compounds. The solids are dried in a rotary drum with indirectly steam-heated air. The product is a crystalline mixture of  $\text{Na}_2\text{SO}_4$  (70%) and  $\text{Na}_2\text{SO}_3$  (30%) with very small amounts of thiosulfates, pyrosulfites and chlorides. This solid can be disposed of by sale as a salt cake mixture to a kraft paper mill. An ethylene glycol refrigeration system is used for cooling in the crystallization system.

The  $\text{SO}_2$  regeneration system consists of a double effect, forced circulation evaporator-crystallizer, with low pressure exhaust steam as the source of heat. Other equipment includes condensers, a condensate stripper and a sodium sulfite dissolving tank. Sodium sulfite is regenerated from sodium bisulfite plus the evolution of  $\text{SO}_2$  by reversing the absorption reaction by the addition of heat.

The rich scrubber liquor combined with the liquor from the  $\text{Na}_2\text{SO}_4$  centrifuges is split between the two evaporator effects, about 55% to effect I and about 45% to effect II. The liquor in the first effect is heated with exhaust steam in an external recirculation heat exchanger and operates at about 200°F under a small vacuum. The vapor from the first effect is used to similarly heat the second effect which operates at about 170°F and a higher vacuum. Water vapor and  $\text{SO}_2$  exit overhead from each evaporator effect and are

partially condensed to remove most of the water and concentrate the  $\text{SO}_2$ . Also, in each effect primarily sodium sulfite crystallizes out to a 45% solids concentration in the recirculating liquor as a slurry. A bleed stream of this slurry is withdrawn to a dissolving tank where the  $\text{Na}_2\text{SO}_3$  is dissolved with recycled condensate from the evaporator condensers. The regeneration reaction in the evaporators is limited by the equilibrium concentration of sulfite ion in solution. Since  $\text{Na}_2\text{SO}_3$  is less soluble than  $\text{NaHSO}_3$ , the crystallization of sulfite continuously removes it from the equilibrium allowing the regeneration reaction to proceed to substantial completion. Several hundred ppm of dissolved  $\text{SO}_2$  is present in the condensate and this is stripped with steam to drive the  $\text{SO}_2$  overhead. The  $\text{Na}_2\text{SO}_3$  slurry is dissolved in a dump tank with stripped condensate and water in a sodium carbonate solution added as sodium makeup to the system to replace that lost in the solid  $\text{Na}_2\text{SO}_4$  purge. The resulting solution is recycled to the scrubber as lean absorbent medium. The resulting  $\text{SO}_2$  exiting the regeneration system contains about 5 to 10% water and is compressed and transferred to the  $\text{SO}_2$  reduction section.

Reducing gas (natural gas) is added in a correct proportion to the rich gas stream, and this gas mixture is then passed through a feed gas heater where its temperature is raised above the dew point of the sulfur formed in the primary reduction system. The principle function of the catalytic reduction reaction is to form over 40% of the total recovered sulfur plus an amount of  $\text{H}_2\text{S}$  to attain an  $\text{H}_2\text{S}/\text{SO}_2$  ratio in the gas stream to approximate the stoichiometric ratio of 2:1 required for the Claus reaction.

The preheated  $\text{SO}_2$  and natural gas mixture enters the primary reduction reactor through a four-way flow reversing valve and is further preheated as it flows upward through a packed bed heat regenerator prior to entering the catalytic reactor. A bypass arrangement may be included in the design to keep a constant temperature on the gases entering the reactor by continuously bypassing a small quantity of cool gas mixture around the up-flow heat regenerator. The heat of reaction in this step is exothermic and is balanced throughout the system. A proprietary catalyst, developed by Allied Chemical,



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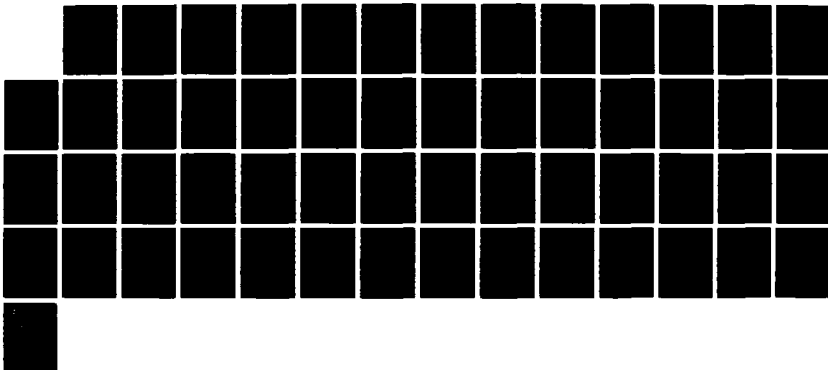
TURBINE FUELS FROM TAR SANDS BITUMEN AND HEAVY OIL  
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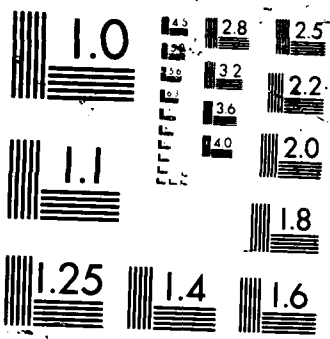
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remains stable at temperatures up to 2000°F and operates generally at a temperature above 1500°F. The main gas flow passes downward through a second heat regenerator yielding its heat to the packing before leaving the system through the four-way reversing valve. A heat balance can be maintained in the system by bypassing a small flow of the reactor hot gases around the heat regenerator and remixing with the main stream after the four-way valve before entering the gas cooler/sulfur condenser. The heat regenerators raise the temperature of the incoming gases so that the  $\text{SO}_2$ /natural gas reaction is initiated. The heating and cooling cycles of the two regenerators are periodically reversed by use of the four-way valve.

A portion of the reactor heat is used to heat the entering  $\text{SO}_2$ /natural gas mixture in an exchanger. The exit gases from the reactor may be too hot and too corrosive for the metal exchanger, so they are tempered by mixing with a portion of gases exiting the gas cooler/sulfur condenser before introduction into the feed gas heat exchanger.

Elemental sulfur formed in the primary reduction step is condensed in two shell-and-tube steam generating condensers. At this stage the condensed sulfur represents somewhat over 40% of the total sulfur recovered. The combined gas stream from the two condensers enters the first stage of a Claus conversion system where more sulfur is formed. The gases emerge at about 750-800°F. The gas is cooled in a steam generating condenser condensing more sulfur. Further conversion of  $\text{H}_2\text{S}$  and  $\text{SO}_2$  to sulfur and water occurs in the second stage Claus reactor at about 550-600°F with the sulfur condensed again in another steam generator. The remaining gases proceed to a coalescer containing mesh screens which remove liquid sulfur entrained in the gas. Molten sulfur from all the condensers is collected in a sulfur holding pit and transferred to sulfur storage at the Sulfur Recovery Unit.

Residual  $\text{H}_2\text{S}$  in the exiting gases from the system are exhausted to the boiler firebox where oxidation to  $\text{SO}_2$  occurs which recycles through the stack gas scrubber.

The desulfurized flue gas leaving the absorber is demisted of entrained droplets by a water wash across a chevron demister and then reheated in the heat exchanger. The reheated flue gas exits the stack at 380°F, which eliminates any water vapor plume.

## FLUE GAS DESULFURIZATION UNIT

### Design Basis

The flue gas is derived from the firing of residual fuel with 1.34 wt% sulfur content, fired at a maximum rate of 3862 BPSD with 30% excess air.

Maximum flue gas to electrostatic precipitator:	1,089,882 lb/hr @ 450°F
Fly ash removed in electrostatic precipitator:	612 lb/hr
Flue gas to absorber containing 1500 ppm SO <sub>2</sub> :	1,089,270 lb/hr
Flue gas from absorber containing 150 ppm SO <sub>2</sub> :	1,087,796 lb/hr
Sulfur recovered:	8.1 tons/day

## FLUE GAS DESULFURIZATION UNIT

### Utilities and Chemicals

450 psig Steam Used	10,000 lb/hr	(1)
Total power required	2,200 KW	(2)
Natural Gas	128,200 SCFD	
Makeup process water	100 GPM	(3)
Cooling water circulated	2,400 GPM	
Soda Ash makeup	221 lb/hr	
Purge sodium sulfate/sodium sulfite (for sale)	285 lb/hr	

- Notes: (1) Approximate, when the flue gas heat exchanger is used  
(2) Approximate from SRI data  
(3) Includes 3% makeup of cooling water circulated

## FLUE GAS DESULFURIZATION UNIT

### Major Equipment

Stanford Research Institute (SRI) information was used to develop capital and operating requirements. No detailed design was obtained since the design would be done by an engineering company (Davy Powergas) that sells the process and design.

Some adjustments were necessary to scale the SRI case to our unit size. Also their case was based on 300°F flue gas from a power boiler with extensive flue gas economizer heat exchange. Our entering flue gas temperature is 450°F which made a gas-to-gas heat exchanger necessary. This cools the inlet gas to 200°F and reheats the exit flue gas to 380°F. Reheating of scrubbed flue gas is necessary to prevent condensation of a stack plume near the ground. The heat exchange eliminates hot air injection into the stack proposed by SRI.

**PROCESS DESIGN SPECIFICATIONS**

**for**

**TANKAGE**



## TANKAGE DESIGN BASIS

All tankage specifications were based on refinery flow rates that were calculated for the production of JP-4 turbine fuel. The economic basis established by the U.S. Air Force included the following inventory specifications:

Crude inventory:	21 days storage capacity
	14 days inventory
Product inventory:	14 days storage capacity
	7 days inventory

## INTERMEDIATE PRODUCTS

Although separate tankage for intermediate refinery products was excluded purposely from the design basis at the direction of the U.S. Air Force, a plan was developed for their potential emergency storage. The plan designates part of the crude oil tank allotment as alternate emergency storage for intermediate products. Each intermediate stream that needs storage capacity is assigned to a tank, which is available for optional crude storage also. Since the design basis allowed for 21 days of crude storage capacity, but only required an inventory of 14 days, the extra seven days of inventory capacity could easily be put into this alternate service, especially if the refinery were supplied by pipeline.

The basis for sizing the intermediate tankage is a capacity of 3.5 days at half the normal flow rate.

Intermediate tankage is provided for the stabilized liquid feedstocks for each of the hydroprocessing units in the event of a downstream unit shutdown. This includes the feedstock for the Hydrovisbreaker, Naphtha Hydrotreater, Distillate Hydrotreater, and the combined Distillate Hydrocracker and Main Fractionator.

The basis for sizing intermediate tankage assumes that in the event of an emergency shutdown at one of the units, the production rate at upstream units would be cut in half. Consequently, the stabilized feedstock to the affected unit would be stored at half its normal production rate for up to 3.5 days.

#### JP-4 VERSUS JP-8 OPERATIONS

Because refinery operations upstream of the Distillate Hydrocracker and Main Fractionator are identical for both the JP-4 and JP-8 operations, the crude and intermediate tankage requirements are the same for both modes of operation.

However, in switching from JP-4 to JP-8 operations, the rate of naphtha production will increase. During JP-4 production the Dehexanizer bottoms, which boils nominally in the range of 120°-275°F, is added to the JP-4 product. But this material boils below the acceptable range for JP-8 turbine fuel, so during JP-8 production, the Dehexanizer is shutdown and the 120°-275°F distillate is routed to naphtha storage. Consequently, the naphtha storage tanks are specified to satisfy storage requirements for the JP-8 operations.

#### HYDROVISBREAKER VACUUM RESIDUUM

The pilot plant sample of Hydrovisbreaker vacuum tower bottoms, which represents the refinery Residuum product, did not blend with normal cutter stocks to a lower viscosity because it was of such a high-boiling nature. Therefore it is stored at 400-450°F to maintain it in a fluid state for use principally as furnace fuel oil for the process heaters and refinery boiler furnace.

#### BUTANE STORAGE

Since all butane is consumed as either feedstock for the Hydrogen Plant or as fuel for the Hydrogen Plant reformer furnace, the product storage capacity was somewhat arbitrarily reduced from 14 days to 3.5 days.

## TANKAGE SPECIFICATIONS

### Crude Oil Tankage

Number of tanks: 5  
Operating temperature: 150°F  
Diameter: 142 ft.  
Height: 56 ft.  
Volume: 157,956 BBL per tank (789,779 BBL total)  
Roof: Covered floater  
Other features: Heated and insulated  
Crude oil storage capacity: 15.29 days at 51653 BPD at 150°F

### Intermediate Hot Tankage

In addition to the stream designations listed below, these "hot" intermediate storage tanks are to be considered as optional storage capacity for Crude Oil.

#### Tank designations:

1. Hydrovisbreaker feed (crude unit vacuum residuum)
2. Distillate Hydrotreater feed
3. Distillate Hydrotreater Depropanizer bottoms (Dist. Hydrocracker Feed)

Number of tanks: 3  
Operating temperature: 200°F (or 150°F if crude oil storage)  
Operating pressure: atmospheric  
Diameter: 100 ft.  
Height: 56 ft.  
Volume (per tank): 439,823 cu.ft. = 78,336 BBL  
Volume total (3 tanks): 1,319,469 cu.ft. = 235,008 BBL  
Roof: Covered floater  
Other features: Heated and insulated  
Crude oil capacity: (at 51653 BPD crude at 150°F)  
1 tank: 1.52 days  
3 tanks: 4.55 days

## TANKAGE SPECIFICATIONS

(continued)

### Intermediate Cold Tankage

In addition to the stream designations listed below, these "cold" intermediate storage tanks are to be considered as optional storage capacity for Crude Oil.

Tank designations:

1. Naphtha Hydrotreater Feed
2. Naphtha Hydrotreater Stripper Bottoms

Number of tanks: 2  
Operating temperature: 100°F (or 150°F if crude oil storage)  
Operating pressure: atmospheric  
Diameter: 82 ft.  
Height: 32 ft.  
Volume (per tank): 168,993 cu.ft. = 30,099 BBL  
Volume total (2 tanks): 337,986 cu.ft. = 60,198 BBL  
Roof: Covered floater  
Other features: Heated and insulated for optional crude use  
Crude oil capacity: (at 51653 BPD crude at 150°F)  
0.58 days for 1 tank  
1.16 days for 2 tanks

### Note on Intermediate tank sizing:

Each tank exceeds the minimum storage requirement for the intermediate products listed above, which consisted of 3.5 days storage at half-flow rate of intermediate product. The tanks were sized larger to satisfy the overall crude oil storage capacity of 21-days.

### Summary of Crude Oil Storage Capacity

Basis: Crude oil at 51653 BPD at 150°F storage (50,000 Std. BPD)

<u>Tankage</u>	<u>Capacity</u>
Crude oil tanks (5):	789,779 BBL (15.29 days crude)
Hot intermediate tanks (3):	235,008 BBL ( 4.55 days crude)
Cold intermediate tanks (2):	60,198 BBL ( 1.16 days crude)
Total crude tankage:	1,084,985 BBL (21.00 days crude)

TANKAGE SPECIFICATIONS  
(continued)

Product Tankage

Butane Product sphere

Number of spheres: 1  
Operating pressure: 50 psig  
Operating temperature: 100°F  
Design pressure: 75 psig  
Design temperature: 650°F (carbon steel)  
Diameter: 65 ft.  
Volume: 143,793 cu.ft. = 25,611 BBL  
Shell thickness: 1.250 in. (includes 0.182" corrosion allowance)  
Sphere weight: 678,610 lb.

Note that all butane product will be consumed as either hydrogen plant feedstock or hydrogen plant fuel.

Naphtha Product tankage

Number of tanks: 1  
Operating temperature: 100°F  
Operating pressure: atmospheric  
Diameter: 114 ft.  
Height: 56 ft.  
Volume (per tank): 571,594 cu.ft. = 101,805 BBL  
Roof: Covered floater

Jet Fuel Product tankage

Number of tanks: 4  
Operating temperature: 100°F  
Operating pressure: atmospheric  
Diameter: 143 ft.  
Height: 56 ft.  
Volume (per tank): 899,394 cu.ft. = 160,188 BBL  
Roof: Covered floater

Residuum Product tankage

Number of tanks: 1  
Operating temperature: 450°F  
Operating pressure: atmospheric  
Diameter: 81 ft.  
Height: 32 ft.  
Volume: 164,896 cu.ft. = 29,369 BBL  
Roof: Cone roof  
Other features: Heated and insulated

TANKAGE SPECIFICATIONS  
(continued)

Sulfur Product storage (14 days total storage capacity)

Sulfur pit:

Location: Below grade, under the sulfur recovery unit  
Rectangular dimensions: 20' long x 20' wide x 8' deep  
Storage capacity: 3200 cu.ft. = 570 BBL  
Construction material: Either concrete or steel, lined with common  
brick in acid-proof cement  
Pit cover: Carbon steel plate  
Operating temperature: 260 - 280°F  
Pit heating: 75 psig steam coils

Sulfur Product tankage:

Number of tanks: 1  
Operating temperature: 260 - 280°F  
Operating pressure: atmospheric  
Tank diameter: 57 ft.  
Tank height: 16 ft.  
Head of sulfur at base: 12.4 psi max.  
Volume: 40828 cu.ft. = 7272 BBL  
Roof: Cone roof  
Tank heating: 75 psig steam coils (25 sq.ft. heating  
surface per sq.ft. of tank wall and roof)  
Other: Insulate entire outer shell and roof

Ammonia product storage sphere

Number of vessels: 1  
Type vessel: Sphere  
Operating temperature: 100°F  
Operating pressure: 200 - 210 psig  
Design temperature: 650°F (carbon steel)  
Design pressure: 225 psig  
Diameter: 42 ft.  
Volume: 39,000 cu.ft.  
Shell thickness: 2.25 in.  
Total weight: 510,000 lb.  
Storage capacity: 7 days of ammonia production

**APPENDIX**

**CAPITAL COST ESTIMATE**

TURBINE FUEL REFINERY  
CAPITAL COST ESTIMATE  
BASIS

The accompanying capital cost estimate for the Turbine Fuel Refinery is based upon the following:

1. The location of the refinery is Salt Lake City, Utah.
2. Mechanical completion occurs during the fourth quarter of 1985.
3. Estimates for the following onsite plants are based on specified equipment lists and flow diagrams:
  - a. Crude Unit
  - b. Hydrovisbreaker
  - c. Naphtha Hydrotreater
  - d. Distillate Hydrotreater
  - e. Distillate Hydrocracker
  - f. Gas Plant
  - g. Low Pressure Amine Unit
4. Estimates for the following onsites plants were factored from historical plant costs, with the exception of specified compressor systems for the Hydrogen Plant and Hydrogen Purification Unit:
  - a. Hydrogen Plant
  - b. Hydrogen Purification Unit
  - c. Sour Water Stripper and Ammonia Plant
  - d. Sulfur Recovery Unit
  - e. Flue Gas Desulfurization Unit
5. Tankage estimates were based on specified tank sizes.



6. The cost of off-sites excluding tankage was assumed to be 45% of the total installed on-site cost estimate.
7. All work is to be accomplished in a standard work week with no overtime.
8. This estimate is for economic evaluation and comparison only.
9. Capitalized spare parts are included as one percent of major equipment cost.
10. In general, royalties, catalysts and chemicals are not included in this estimate. However, the initial loading of molecular sieves for the Hydrogen Purification Unit are included in the cost.
11. Offsites have been estimated as 45% of the total onsite installed costs at the request of the U.S. Air Force.

TURBINE FUEL REFINERY  
CAPITAL COST SUMMARY

ONSITES	<u>Installed Cost</u> <sup>1</sup>
Crude Unit .....	\$ 19,968,000
Hydrovisbreaker Unit .....	168,119,000
Naphtha Hydrotreater Unit .....	24,154,000
Distillate Hydrotreater Unit .....	140,183,000
Distillate Hydrocracker Unit .....	94,724,000
Gas Plant .....	9,380,000
Hydrogen Plant .....	99,075,000
Hydrogen Purification Unit .....	60,915,000
Low Pressure Amine Unit .....	3,079,000
Sour Water Stripper and Ammonia Plant .....	33,091,000
Sulfur Recovery Unit .....	37,119,000
Flue Gas Desulfurization Unit .....	<u>54,328,000</u>
 Total Onsites .....	 \$ 744,135,000

OFFSITES

Tankage .....	\$ 45,061,000
Other: Specified by U.S. Air Force as 45% of onsite costs .....	334,861,000
 SPARE PARTS .....	 1,498,000
 ROUND UP TO NEAREST MILLION DOLLARS .....	 445,000

TOTAL REFINERY INSTALLED COST ..... \$ 1,126,000,000

<sup>1</sup> Based on 4th Quarter 1985 prices, Salt Lake City, Utah location

Factored Estimate Worksheet Summary

	Crude Unit	Naphtha Hydrotreater	Hydrovisbreaker	Distillate Hydrotreater	Distillate Hydrocracker	Gas Plant
Total Major Equipment	\$11,833,400	\$14,427,600	\$100,607,500	\$84,576,800	\$56,649,100	\$5,684,100
Instruments	2,058,000	2,375,200	16,344,400	12,941,600	9,245,600	840,400
Insulation	(400)	200	100	(400)	300	500
Rounding	13,890,000	16,803,000	116,952,000	97,518,000	65,895,000	6,525,000
Total Direct Installed Cost	2,084,000	2,520,000	17,543,000	14,628,000	9,884,000	979,000
Home Office Costs	15,974,000	19,323,000	134,495,000	112,146,000	75,779,000	7,504,000
Subtotal	3,994,000	4,831,000	33,624,000	28,037,000	18,945,000	1,876,000
Contingency	19,968,000	24,154,000	168,119,000	140,183,000	94,724,000	9,380,000
Total Installed Costs						
	Hydrogen Plant	Hydrogen Purification Plant	Stack Gas Scrubbing and Collecting Unit	Low Pressure Amine Unit	Sour Water Stripper and Ammonia Plant	Sulfur Recovery Unit
Total Major Equipment	\$52,705,000	\$32,685,900	\$31,276,800	\$1,864,000	\$18,993,500	\$22,132,800
Instruments	16,216,800	9,689,600	6,516,000	278,400	4,026,800	3,688,800
Insulation	200	500	200	(400)	(300)	400
Rounding	68,922,000	42,376,000	37,793,000	2,142,000	23,020,000	25,822,000
Total Direct Installed Cost	10,338,000	6,356,000	5,669,000	321,000	3,453,000	3,873,000
Home Office Costs	79,260,000	48,732,000	43,462,000	2,463,000	26,473,000	29,695,000
Subtotal	19,815,000	12,183,000	10,866,000	616,000	6,618,000	7,424,000
Contingency	99,075,000	60,915,000	54,328,000	3,079,000	33,091,000	37,119,000
Total Installed Costs						
	Total Onsites	Offsites excl. tankage	Tankage	Total Refinery Capital Cost		
Total Major Equipment	\$433,435,500	-	\$22,822,600	\$456,258,100		
Instruments	84,221,600	-	5,753,600	89,975,200		
Insulation	-	-	2,770,800	2,770,800		
Rounding	517,658,000	-	31,347,000	549,005,000		
Total Direct Installed Cost	77,648,000	-	4,702,000	82,350,000		
Home Office Costs	595,306,000	-	36,049,000	631,355,000		
Subtotal	148,829,000	-	9,012,000	157,841,000		
Contingency	744,135,000	334,861,000	45,061,000	1,124,057,000		
Total Installed Costs				1,124,057,000		
Spare Parts (1% of \$149,756,600 material cost of total major equipment)				1,498,000		
Roundup to nearest million dollars				445,000		
Refinery Total Installed Cost				1,126,000,000		

SUN REFINING AND MARKETING CO.

P. E. T. S. FACTORED ESTIMATE WORKSHEET

DATE: 16-Mar-87

PROJECT: TURBINE FUELS PROJECT  
CRUDE UNIT

ESTIMATE NO: 7263  
JOB NO: 848768

BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS	1					358,400				\$1,433,600
300R	REACTORS										
320A	AIR COOLED HEAT EXCHANGERS	1					37,000				\$160,200
320S	SHELL & TUBE HEAT EXCHANGERS	14					742,700				\$2,844,600
370H	FIRED HEATERS	2					1,198,500				\$2,696,600
400T	TANKS										
400V	VESSELS	7					656,700				\$2,646,400
450C	COMPRESSORS										
450P	PUMPS	16					427,000				\$2,028,500
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T PROCESS SYS. RALPH. M. PARSONS CO. SYSTEM	1					10,000				\$22,500
A	TOTAL MAJOR EQUIPMENT	41					3,430,300		3.45		\$11,832,400
600	INSTRUMENTS	15	% M.E.				514,500		4.00		\$2,058,000
620	INSULATION										
	ROUNDING										(\$-00)
A+B	TOTAL DIRECT INSTALLED COST										\$13,890,000
	HOME OFFICE COSTS		15 % D.I.C.								\$2,084,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$15,974,000
	CONTINGENCY		25 %								\$3,994,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/0/85								\$19,968,000
	ESCALATION		% /YR. FOR	YEAR							
	TOTAL INSTALLED COST		4/0/85								\$19,968,000

PROJECT: TURBINE FUELS PROJECT  
CRUDE UNITESTIMATE NO: 7263  
JOB NO: 848768BY: J.T. MARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	FACTOR HIGH	USED	INSTALLED COST
270P	T1 VACUUM TOWER 650 F FULL VAC. 12.5' X 63' CS/CS W/SKIRT 122,000# SOURCE: HOOPER TRAYS: VALVE TYPE ALLOY QTY 30 SOURCE: MUTTER	1	300,000	3 %	15,600	358,400	358,400	4.00	4.00	4.00	\$1,433,600
320A	E12 RECYCLE CLR. 350 F 50 PSIG 1 - 12' X 20' BAY CS SOURCE: HAPPY	1	34,000	3 %	2,000	37,000	37,000	4.10	5.00	4.33	\$160,200
320S	E1 CIRC REFLUX-DESALT PD. AES 450 F 450 PSIG 7,600 SF CS/CS SOURCE: WESTERN	2	72,200	3 %	2,900	77,300	154,600	3.50	4.80	3.83	\$592,100
	E2 FLASH TWR ON COND AJS 450 F 450 PSIG 6,000 SF CS/CS SOURCE: WESTERN	1	57,000	3 %	2,300	61,000	61,000	3.50	4.80	3.83	\$233,600
	E3 FLASH TWR PREHTR #1 AJS 450 F 450 PSIG 2,200 SF CS/CS SOURCE: WESTERN	1	24,200	3 %	1,000	25,900	25,900	3.50	4.80	3.83	\$99,200
	E4 FLASH TWR PREHTR #2 AES 550F 450 PSIG 1,940 SF CS/CS SOURCE: WESTERN	1	21,350	3 %	900	22,900	22,900	3.50	4.80	3.83	\$87,700
	E5 FLASH TWR PREHTR #3 AJS 625F 450 PSIG 5,000 SF CS/CS SOURCE: WESTERN	1	47,500	3 %	1,900	50,800	50,800	3.50	4.80	3.83	\$194,600
	E6 FLASH TWR PREHTR #4 AES 625F 400 PSIG 6,750 SF CS/SCR SOURCE: WESTERN	2	105,000	3 %	4,200	112,400	224,800	3.50	4.80	3.83	\$861,000
	E7 CIRC REFLUX TRIM CLR AES 300F 75 PSIG 760 SF CS/ADMIRALTY SOURCE: WESTERN	1	21,500	3 %	900	23,000	23,000	3.50	4.80	3.83	\$88,100
	E8 NAPTHA PROD CLR AES 300F 75 PSIG 2,400 SF CS/ADMIRALTY SOURCE: WESTERN	1	43,200	3 %	1,700	46,200	46,200	3.50	4.80	3.83	\$176,900
	E9 VAC TWR ON COND AES 300F 10 TO 75 PSIG 2,000 SF CS/ADMIRALTY SOURCE: WESTERN	1	36,000	3 %	1,400	38,500	38,500	3.50	4.80	3.83	\$147,500
	E10 FRESH - SALT WTR EXCH AES 350F 600 PSIG 880 SF CS/CS SOURCE: WESTERN	2	26,750	3 %	1,000	26,500	53,000	3.50	4.80	3.83	\$203,300
	E11 DESASTER EFF WTR CLR AES 300F 400 PSIG 1,400 SF CS/ADMIRALTY SOURCE: WESTERN	1	39,200	3 %	1,600	42,000	42,000	3.50	4.80	3.83	\$160,900
370H	H1 FLASH TWR FEED 99.2 MM BTU/HR H2 FLASH TWR FEED 46.2 MM BTU/HR SOURCE: BOURNE	1 1	1,115,000	3 %	50,000	1,198,500	1,198,500	2.00	3.00	2.25	\$2,696,600

PROJECT: TURBINE FUELS PROJECT  
CRUDE UNITESTIMATE NO: 7263  
JOB NO: 848768BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
400V	V1 RECYCLE SURGE DRUM 350 F 50 PSIG 7.5' X 16' CS/CS W/SADDLES 9,700# SOURCE: BUFFALO	1	12,500	3 %	1,725	14,600	14,600	4.00	4.10	4.03	\$58,800
	V2 DESALTER 350 F 350 PSIG 12' X 52' CS/CS W/SADDLES 175,000# SOURCE: BUFFALO	2	154,000	3 %	30,500	189,100	378,200	4.00	4.10	4.03	\$1,524,100
	V3 FLASH TWR 625 F 50 PSIG 9' X 25' CS/CS W/SKIRT 24,000# SOURCE: BUFFALO	1	52,500	3 %	4,700	58,800	58,800	4.00	4.10	4.03	\$237,000
	V4 VAC TWR ON REC 250 F 50 PSIG 3.5' X 6' CS/CS W/SADDLES 1,600# SOURCE: BUFFALO	1	4,000	3 %	350	4,500	4,500	4.00	4.10	4.03	\$18,100
	V5 FEED SURGE TANK 300 F 50 PSIG 16' X 65' CS/CS W/SKIRT 106,000# SOURCE: BUFFALO	1	167,500	3 %	17,300	189,800	189,800	4.00	4.10	4.03	\$764,900
	V6 STEAM DRUM 400 F 200 PSIG 5' X 10' CS/CS W/SADDLES 6,200# SOURCE: BUFFALO	1	9,500	3 %	1,000	10,800	10,800	4.00	4.10	4.03	\$43,500
450P	P1 CRUDE CYCLE FEED CS/12CR HOR 2310 GPM DISCH 350 PSIG DELTA 350 PSI - 885' PUMP TEMP 175 F @ 0.91 SG MOTOR HP 725 @ 3600 RPM SOURCE: UNITED	2	63,000	3 %	1,000	65,900	131,800	4.00	7.00	4.75	\$626,100
	P2 RECLE DILT CS/12CR HOR 600 GPM DISCH 45 PSIG DELTA 45 PSI - 135' PUMP TEMP 250 F @ 0.77 SG MOTOR HP 35 @ 3600 RPM SOURCE: UNITED	2	10,950	3 %	220	11,500	23,000	4.00	7.00	4.75	\$109,300
	P3 VAC TWR FEED CS/12CR HOR 1940 GPM DISCH 173 PSIG DELTA 165 PSI - 470' PUMP TEMP 572 F @ 0.81 SG MOTOR HP 310 @ 3600 RPM SOURCE: UNITED	2	32,300	3 %	450	33,700	67,400	4.00	7.00	4.75	\$320,200
	P4 CIRC REFLUX CS/12CR HOR 28.3 GPM DISCH 85 PSIG DELTA 80 PSI - 233' PUMP TEMP 368 F @ 0.79 SG MOTOR HP 130 @ 3600 RPM SOURCE: UNITED	2	21,300	3 %	350	22,300	44,600	4.00	7.00	4.75	\$211,900
	P5 RED'D CRUDE CS/12CR HOR 1600 GPM DISCH 115 PSIG DELTA 122 PSI - 338' PUMP TEMP 595 F @ 0.83 SG MOTOR HP 205 @ 3600 RPM SOURCE: UNITED	2	28,000	3 %	450	29,300	58,600	4.00	7.00	4.75	\$278,400
	P6 VAC TWR ON C CS/12CR HOR 28.3 GPM DISCH 70 PSIG DELTA 70 PSI - 192' PUMP TEMP 105 F @ 0.84 SG MOTOR HP 2.6 @ 3600 RPM SOURCE: UNITED	2	8,000	3 %	150	8,400	16,800	4.00	7.00	4.75	\$79,800

SUN REFINING AND MARKETING CO.

P. E. T. S. FACTORED ESTIMATE WORKSHEET

DATE: 16-Mar-87

PROJECT: TURBINE FUELS PROJECT  
CRUDE UNIT

ESTIMATE NO: 7263  
JOB NO: 848768

BY: J.T. MARLAM  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
	P7 DES W.W. BSTR CS/12CR HOR 150 GPM DISCH 355 PSIG DELTA 75 PSI - 188' PUMP TEMP 300 F @ 0.92 SG MOTOR HP 14.6 @ 3600 RPM SOURCE: UNITED	2	8,400	3 %	150	8,800	17,600	4.00	7.00	4.75	\$83,500
	P8 DES W.W. FEED CS/12CR HOR 150 GPM DISCH 335 PSIG DELTA 335 PSI - 786' PUMP TEMP 140 F @ 0.98 SG MOTOR HP 53 @ 3600 RPM SOURCE: UNITED	2	32,200	3 %	450	33,600	67,200	4.00	7.00	4.75	\$319,200
480	STEAM JET VACUUM SYS 665 GPM FLOW SOURCE: CROLL REYNOLDS	1	9,500	3 %	200	10,000	10,000	2.00	3.00	2.25	\$22,500

SUN REFINING AND MARKETING CO.

P. E. T. S. FACTORED ESTIMATE WORKSHEET

DATE: 16-Mar-87

PROJECT: TURBINE FUELS PROJECT  
 NAPTHA HYDROTREATER  
 ESTIMATE NO: 7269  
 JOB NO: 848768

BY: J. T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS	2					319,400				\$1,277,600
300R	REACTORS	1					603,900				\$2,415,600
320A	AIR COOLED HEAT EXCHANGERS	3					307,600				\$1,331,900
320S	SHELL & TUBE HEAT EXCHANGERS	14					1,195,500				\$4,578,800
370H	FIRED HEATERS	3					475,900				\$951,800
400T	TANKS										
400V	VESSELS	9					263,700				\$1,062,700
450C	COMPRESSORS	1					365,200				\$777,900
450P	PUMPS	16					427,600				\$2,031,300
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.U.T PROCESS SYS. RALPH M. PARSONS CO. SYSTEM										
A	TOTAL MAJOR EQUIPMENT	49					3,958,800		3.64		\$14,427,600
600	INSTRUMENTS	15 % N.E.					593,800		4.00		\$2,375,200
620	INSULATION										\$200
	ROUNDING										\$200
A+B	TOTAL DIRECT INSTALLED COST										\$16,803,000
	HOME OFFICE COSTS	15 % D.I.C.									\$2,520,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$19,323,000
	CONTINGENCY	25 %									\$4,831,000
	ROYALTIES										
	TOTAL INSTALLED COST	4/0/85									\$24,154,000
	ESCALATION		% /YR. FOR	YEAR							
	TOTAL INSTALLED COST	4/0/85									\$24,154,000



PROJECT: TURBINE FUELS PROJECT  
NAPHTHA HYDROTREATER

ESTIMATE NO: 7269  
JOB NO: 848768

BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	HIGH	USED	INSTALLED COST
270P	T1 FEED SPLITTER 650 F 75 PSIG 9' X 74' CS/CS W/SKIRT 68,000# SOURCE: HOOPER TRAYS: VALVE TYPE ALLOY QTY 30 SOURCE: MUTTER	1	142,400	3 %	11,400	182,100	182,100	4.00	4.00	4.00	\$728,400
	T2 STRIPPER 650 F 200 PSIG 6' X 56' CS/CS W/SKIRT 50,000# SOURCE: HOOPER TRAYS: VALVE TYPE ALLOY QTY 36 SOURCE: MUTTER	1	110,400	3 %	6,600	137,300	137,300	4.00	4.00	4.00	\$549,200
300R	R1 NAPHTHA HTR REACTOR 775 F 1815PSIG 8.5' X 22' 2.25CR W/O'LAY 185,400# SOURCE: HOOPER PACKED BEDS QTY 2 SOURCE:	1	556,200	3 %	31,000	603,900	603,900	4.00	4.00	4.00	\$2,415,600
320A	E2 SPLITTER COND. 650 F 75 PSIG 1 - 12' X 36' BAY CS SOURCE: HAPPY	1	60,700	3 %		62,500	62,500	4.10	5.00	4.33	\$270,600
	E6 R1 EFFLUENT A/C 650 F 1630 PSIG 2 - 18' X 40' BAYS 410 SS SOURCE: HAPPY	1	186,360	3 %		192,000	192,000	4.10	5.00	4.33	\$831,400
	E8 STRIPPER COND. 650 F 200 PSIG 1 - 12' X 36' BAY CS SOURCE: HAPPY	1	51,600	3 %		53,100	53,100	4.10	5.00	4.33	\$229,900
320S	E1 FEED BOTTOMS AES 650 F 150 PSIG 1120 SF CS/CS SOURCE: HUGHES ANDERSON	3	15,000	3 %		15,500	46,500	3.50	4.80	3.83	\$178,100
	E3 R1 EFF NA CLR AES 850 F 1395 PSIG 1440 SF 1CR O'LAY / 347 SOURCE: HUGHES ANDERSON	3	197,400	3 %		203,300	609,900	3.50	4.80	3.83	\$2,335,900
	E4 R1 EFF NAPHTHA AES 650 F 1440 PSIG 1100 SF 1CR O'LAY / 347 SOURCE: HUGHES ANDERSON	2	150,500	3 %		155,000	310,000	3.50	4.80	3.83	\$1,187,300
	E5 R1 EFF STM GEN AKT 650F 1315 PSIG 1250 SF CS/CS SOURCE: WESTERN	1	52,000	3 %		53,600	53,600	3.50	4.80	3.83	\$205,300
	E7 STRIP'R FD BTMS AES 650F 300 PSIG 1764 SF CS/CS SOURCE: WESTERN	5	34,050	3 %		35,100	175,500	3.50	4.80	3.83	\$672,200

PROJECT: TURBINE FUELS PROJECT  
NAPHTHA HYDROTREATER

ESTIMATE NO: 7269  
JOB NO: 848768

BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST
								LOW	HIGH	USED	
370H	H1 SPLITTER REBOILER 40.6 MM BTU/HR SOURCE BORN:	1	157,000	3 %		161,700	161,700	2.00	2.00	2.00	\$323,400
	H2 RECYCLE GAS HTR 28.34 MM BTU/HR SOURCE BORN:	1	155,000	3 %		159,700	159,700	2.00	2.00	2.00	\$319,400
	H3 STRIPPER REBOILER 18.85 MM BTU/HR SOURCE BORN:	1	150,000	3 %		154,500	154,500	2.00	2.00	2.00	\$309,000
400V	V1 FEED DRUM 650 F 40 PSIG 10' X 32' CS/CS W/SKIRT 24,500# SOURCE: BUFFALO	1	35,500	3 %	4,800	41,400	41,400	4.00	4.10	4.03	\$166,800
	V2 SPLITTER SEPARATOR 650 F 50 PSIG 8' X 16' CS/CS W/SADDLES 8,200# SOURCE: BUFFALO	1	14,000	3 %	1,700	16,100	16,100	4.00	4.10	4.03	\$64,900
	V3 COMP. K O DRUM 300 F 1650 PSIG 3' X 10' CS/CS W/SKIRT 12,000# SOURCE: NOOTER	1	18,000	3 %	2,100	20,600	20,600	4.00	4.10	4.03	\$83,000
	V4 HI PRESS SEP 300 F 1650 PSIG 5.5' X 18' CS/CS W/SKIRT 51,000# SOURCE: NOOTER	1	76,500	3 %	8,550	87,300	87,300	4.00	4.10	4.03	\$351,800
	V5 LOW PRESS SEP 300 F 175 PSIG 8' X 25' CS/CS W/SADDLES 28,000# SOURCE: BUFFALO	1	30,600	3 %	4,600	36,100	36,100	4.00	4.10	4.03	\$145,500
	V6 STRIPPER REC'R 650 F 200 PSIG 5' X 20' CS/CS W/SADDLES 10,000# SOURCE: BUFFALO	1	17,060	3 %	2,300	19,900	19,900	4.00	4.10	4.03	\$80,200
	V7 STEAM DRUM 650 F 200 PSIG 5' X 15' CS/CS W/SADDLES 8,000# SOURCE: BUFFALO	1	14,000	3 %	1,900	16,300	16,300	4.00	4.10	4.03	\$65,700
	V8 STEAM DRUM 650 F 200 PSIG 4' X 12' CS/CS W/SADDLES 4,500# SOURCE: BUFFALO	1	12,500	3 %	1,000	13,900	13,900	4.00	4.10	4.03	\$55,000
	V9 STEAM DRUM 650 F 200 PSIG 4' X 8' CS/CS W/SADDLES 2,900# SOURCE: BUFFALO	1	11,000	3 %	800	12,100	12,100	4.00	4.10	4.03	\$48,300

PROJECT: TURBINE FUELS PROJECT  
NAPHTHA HYDROTREATER

ESTIMATE NO: 7269  
JOB NO: 848768

BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST
								LOW	HIGH	USED	
450C	K1 RECYCLE COMP. CENT. 800 BHP 54,293 SCFM @ 1.38 K & 777 MOLE WT SUCTION 108 F @ 1190 PSIA DISCH. 126 F @ 1325 PSIA SPARES: SOURCE: I R	1	340,000	3 %		350,200	365,200	2.00	2.50	2.13	\$777,900
450P	P1 FEED PUMP CS/12CR HOR 387 GPM DISCH 60 PSIG DELTA 60 PSI - 180' PUMP TEMP 332 F @ 0.5 SG MOTOR HP 25 @ 3600 RPM SOURCE: UNITED	2	12,200	3 %		12,600	25,200	4.00	7.00	4.75	\$119,700
	P2 SPLTR BTMS 12CR/12CR HOR 265 GPM DISCH 100 PSIG DELTA 92 PSI - 308' PUMP TEMP 594 F @ 0.69 SG MOTOR HP 25 @ 3600 RPM SOURCE: UNITED	2	17,400	3 %		17,900	35,800	4.00	7.00	4.75	\$170,100
	P3 REBLR PUMP 12CR/12CR HOR 1000 GPM DISCH 70 PSIG DELTA 62 PSI - 208' PUMP TEMP 574 F @ 0.69 SG MOTOR HP 70 @ 3600 RPM SOURCE: UNITED	2	21,600	3 %		22,200	44,400	4.00	7.00	4.75	\$210,900
	P4 SPLTR REF 12CR/12CR HOR 211 GPM DISCH 68 PSIG DELTA 68 PSI - 204' PUMP TEMP 259 F @ 0.77 SG MOTOR HP 20 @ 3600 RPM SOURCE: UNITED	2	11,600	3 %		11,900	23,800	4.00	7.00	4.75	\$113,100
	P5 REACTOR FEED CS/12CR HOR 541 GPM DISCH 1370PSIG DELTA 1370PSI - 4187' PUMP TEMP 150 F @ 0.75 SG MOTOR HP 700 @ 3600 RPM SOURCE: UNITED	2	102,700	3 %		105,800	211,600	4.00	7.00	4.75	\$1,005,100
	P6 STRIPPER CHG CS/12CR HOR 327 GPM DISCH 270 PSIG DELTA 120 PSI - 360' PUMP TEMP 116 F @ 0.60 SG MOTOR HP 75 @ 3600 RPM SOURCE: UNITED	2	14,000	3 %		14,400	28,800	4.00	7.00	4.75	\$136,800
	P7 STRPR REBLR CS/12CR HOR 886 GPM DISCH 243 PSIG DELTA 78 PSI - 313' PUMP TEMP 445 F @ 0.56 SG MOTOR HP 75 @ 3600 RPM SOURCE: UNITED	2	17,400	3 %		17,900	35,800	4.00	7.00	4.75	\$170,100
	P8 STRPR REFLX CS/12CR HOR 105 GPM DISCH 211 PSIG DELTA 61 PSI - 270' PUMP TEMP 114 F @ 0.52 SG MOTOR HP 10 @ 3600 RPM SOURCE: UNITED	2	10,780	3 %		11,100	22,200	4.00	7.00	4.75	\$105,500

SUN REFINING AND MARKETING CO.

P. E. T. S. FACTORED ESTIMATE WORKSHEET

DATE: 16-Mar-87

PROJECT: TURBINE FUELS PROJECT  
HYDROVISBREAKER

ESTIMATE NO: 7273  
JOB NO: 848768

BY: J. T. HARLAM  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS	4					1,619,200				\$6,476,800
300R	REACTORS	3					9,360,000				\$37,440,000
320A	AIR COOLED HEAT EXCHANGERS	1					114,400				\$495,400
320S	SHELL & TUBE HEAT EXCHANGERS	26					5,053,500				\$19,354,800
370H	FIRED HEATERS	3					2,756,000				\$5,512,000
400T	TANKS	3					353,800				\$707,600
400V	VESSELS	13					4,354,700				\$17,264,600
450C	COMPRESSORS	3					1,466,400				\$3,123,400
450P	PUMPS	32					2,146,800				\$10,197,600
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T PROCESS SYS. RALPH M. PARSONS CO. SYSTEM	1					15,700				\$35,500
A	TOTAL MAJOR EQUIPMENT	89					27,240,500	3.69			\$100,607,500
600	INSTRUMENTS	15	% M.E.				4,086,100	4.00			\$16,344,400
620	INSULATION										\$100
A+B	TOTAL DIRECT INSTALLED COST										\$116,952,000
	HOME OFFICE COSTS	15	% D.I.C.								\$17,543,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$134,495,000
	CONTINGENCY	25	%								\$33,624,000
	ROYALTIES										
	TOTAL INSTALLED COST	4/0/85									\$168,119,000
	ESCALATION		% /YR. FOR	YEAR							
	TOTAL INSTALLED COST	4/0/85									\$168,119,000

PROJECT: TURBINE FUELS PROJECT  
HYDROVISBREAKER

ESTIMATE NO: 7273  
JOB NO: 848768

BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST
								LOW	HIGH	USED	
270P	T1 AMINE ABSORBER 650 F 2,700 PSIG 5.5' X 50' CS/CS W/SKIRT 285,000# SOURCE: TRAYS: VALVE TYPE 12CR QTY 20 SOURCE:	1	920,000	3 %	9,400	977,000	977,000	4.00	4.00	4.00	\$3,908,000
	T2 AMINE STILL 650 F 75 PSIG 6.5' X 50' CS/CS W/SKIRT 17,000# SOURCE: TRAYS: VALVE TYPE 12CR QTY 20 SOURCE:	1	46,200	3 %	700	68,300	68,300	4.00	4.00	4.00	\$273,200
	T3 ATMOSPHERIC TWR 650 F 75 PSIG 8/26/10' X 44/6.5/12' CS/CS WITH SKIRT 50,000# SOURCE: MOOTER TRAYS: VALVE TYPE 12CR QTY 32 SOURCE: EST.	1	244,000	3 %	2,800	286,100	286,100	4.00	4.00	4.00	\$1,144,400
	T4 VACUUM TOWER 750 F FULL VACUUM 20/10' X 20/12' CS/CS W/SKIRT 95000# SOURCE: MOOTER TRAYS: VALVE TYPE 12CR QTY 12 SOURCE: EST.	1	265,000	3 %	2,800	287,800	287,800	4.00	4.00	4.00	\$1,151,200
300R	HYDRO-VIS REACTORS 850 F 2805 PSIG 10' X 43' 2.25CR W/O LAY 838,100# SOURCE: C.B.I. SUPPORT GRIDS QTY 2 SOURCE: C.B.I.	3	3,000,000	3 %	30,000	3,120,000	9,360,000	4.00	4.00	4.00	\$37,440,000
320A	E16 VAC BTMS A/C 650 F 150 PSIG 1 - 12' X 40' BAY CS/5CR SOURCE: HOFFMAN	1	110,000	3 %	1,100	114,400	114,400	4.10	5.00	4.33	\$495,400
320S	E1 R FEED EFF. AEU 950 F 2740 PSIG 6100 SF 5 CR W/347 OL/347 TUBES SOURCE: EFCO	2	350,000	3 %	3,500	364,000	728,000	3.50	4.80	3.83	\$2,788,200
	E2 R EFF RECYCLE GAS AEU 800 F 2855 PSIG 2621 SF 2.25CR W/347 OL/347 TUBES SOURCE: EFCO	1	200,000	3 %	2,000	208,000	208,000	3.50	4.80	3.83	\$796,600
	E3 R EFF 450# STM GEN AKU 700 F 2730 PSIG 2236 SF CS /347 TUBES SOURCE: EFCO	1	235,000	3 %	2,400	244,500	244,500	3.50	4.80	3.83	\$936,400
	E4 HOT SEPR VAP 150#STM GEN AKU 700F 2740 PSIG 7944 SF CS/347 TUBES SOURCE: EFCO	1	450,000	3 %	4,500	468,000	468,000	3.50	4.80	3.83	\$1,792,400
	E5 HOT SEPR VAP 50# STM GEN AKU 650F 2710 PSIG 5556 SF CS/347 TUBES SOURCE: EFCO	1	350,000	3 %	3,500	364,000	364,000	3.50	4.80	3.83	\$1,394,100
	E6 HOT SEPR VAP RECYCLE GAS AEU 650F 2880 PSIG 3915 SF 2.25CR/347 TUBES SOURCE: EFCO	3	275,000	3 %	2,800	286,100	858,300	3.50	4.80	3.83	\$3,287,300
	E7 HOT SEPR VAP COOL WATER AEU 650F 2690 PSIG 5556 SF CS/MOMEL TUBES SOURCE: EST.	2	314,000	3 %	3,100	326,500	653,000	3.50	4.80	3.83	\$2,531,000

PROJECT: TURBINE FUELS PROJECT  
 HYDROVISBREAKER  
 ESTIMATE NO: 7273  
 JOB NO: B-8768

BY: J.T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	HIGH	USED	INSTALLED COST
	E5A HOT VAPOR SEPARATOR 1/2" DIA. 1/2" WALL 650F 175 PSIG 1084 SF 10847 TUBES SOURCE: EST.	1	17,000	3 %	200	17,700	17,700	3.50	4.80	3.83	\$67,800
	E5B HOT VAPOR SEPARATOR 1/2" DIA. 1/2" WALL 650F 175 PSIG 1084 SF 10847 TUBES SOURCE: EST.	1	55,000	3 %	600	57,300	57,300	3.50	4.80	3.83	\$219,500
	E9 AMINE PUMP 1/2" DIA. 1/2" WALL 650F 200 PSIG 2057 SF 10847 TUBES SOURCE: EST.	2	180,000	3 %	1,800	187,200	374,400	3.50	4.80	3.83	\$1,434,000
	E10 AMINE S... 175 PSIG 1084 SF 10847 TUBES 60 TUBES SOURCE: EST.	1	136,000	3 %	1,400	141,500	141,500	3.50	4.80	3.83	\$541,900
	E10A LEAN AMINE REGENERATOR 200 PSIG 1000 SF 10847 TUBES 60 TUBES SOURCE: EST.	2	152,000	3 %	1,500	158,100	316,200	3.50	4.80	3.83	\$1,211,000
	E11 AMINE STRIPPER 1/2" DIA. 1/2" WALL 650F 175 PSIG 4524 SF 10847 TUBES TUBES SOURCE: EST.	1	339,300	3 %	3,400	352,900	352,900	3.50	4.80	3.83	\$1,351,600
	E12 AMINE REGENERATOR 1/2" DIA. 1/2" WALL 650F 200 PSIG 450 SF 10847 TUBES TUBES SOURCE: EST.	1	45,000	3 %	500	46,900	46,900	3.50	4.80	3.83	\$179,600
	E12A ATM TURBINE 1/2" DIA. 1/2" WALL 650F 100 PSIG 2560 SF 10847 TUBES SOURCE: EST.	1	31,000	3 %	300	32,200	32,200	3.50	4.80	3.83	\$123,300
	E12B ATM TURBINE 1/2" DIA. 1/2" WALL 650F 150 PSIG 3330 SF 10847 TUBES TUBES SOURCE: EST.	2	47,000	3 %	500	48,900	97,800	3.50	4.80	3.83	\$374,600
	E13 ATM TURBINE 1/2" DIA. 1/2" WALL 650F 150 PSIG 1625 SF 10847 TUBES TUBES SOURCE: EST.	1	21,200	3 %	200	22,000	22,000	3.50	4.80	3.83	\$84,300
	E14 VAC TURBINE 1/2" DIA. 1/2" WALL 650F 100 PSIG 5012 SF 10847 TUBES SOURCE: EST.	1	52,000	3 %	500	54,100	54,100	3.50	4.80	3.83	\$207,200
	E15 VAC TURBINE 1/2" DIA. 1/2" WALL 650F 150 PSIG 1010 SF 10847 TUBES SOURCE: EST.	1	16,000	3 %	200	16,700	16,700	3.50	4.80	3.83	\$64,000

PROJECT: TURBINE FUELS PROJECT  
HYDROVISBREAKER

ESTIMATE NO: 7273  
JOB NO: B4876B

BY: J.T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	HIGH	USED	INSTALLED COST
370H	H1 RECYCLE GAS HTR 110.1 MM BTU/HR SOURCE: EST.	1	1,800,000	3 %	18,000	1,872,000	1,872,000	2.00	2.00	2.00	\$3,744,000
	H2 ATM TWR FEED HTR 101.0 MM BTU/HR SOURCE: EST.	1	500,000	3 %	5,000	520,000	520,000	2.00	2.00	2.00	\$1,040,000
	H3 VAC TWR FEED HTR 61.0 MM BTU/HR SOURCE: EST.	1	350,000	3 %	3,500	364,000	364,000	2.00	2.00	2.00	\$728,000
400T	TK 1 CRUDE RESIN HOR 12' X 50' W/SAD DES 650 F 30 PSIG CS W/SS OLAY SOURCE: BUFFALO	1	150,000	3 %	1,500	156,000	156,000	2.00	2.00	2.00	\$312,000
	TK 2 ADDITIVE TANK 20' DIA X 16' H. DES 150 F ATMOSPHERIC CONE BOTTOM FIXED ROOF SOURCE: BUFFALO	2	95,000	3 %	1,000	98,900	197,800	2.00	2.00	2.00	\$395,600
400V	V1 HOT HP SEP 650 F 2715 PSIG 9' X 20' 1.25CR W/SDLS 371,300# SOURCE: EST.	1	1,370,000	3 %	13,700	1,424,800	1,424,800	3.50	4.80	3.83	\$5,457,000
	V2 HOT LP SEP. 650 F 200 PSIG 7' X 26' CS/CS W/SDLS 21,500# SOURCE: BUFFALO	1	50,000	3 %	500	52,000	52,000	4.00	4.10	4.03	\$209,600
	V2A COOLED HT SEP VAP 500 F 200 PSIG 3' X 9' CS/CS W/SDLS 4,000# SOURCE: BUFFALO	1	6,250	3 %	100	6,500	6,500	4.00	4.10	4.03	\$26,200
	V3 COLD HI-PRESS SEP 650 F 2680 PSIG 8' X 24' CSW/347 OL W/SDLS 324,200# SOURCE: EST.	1	1,195,000	3 %	12,000	1,242,900	1,242,900	4.00	4.10	4.03	\$5,008,900
	V4 COLD LP SEP 650 F 200 PSIG 7' X 24' CS/CS W/SDLS 20,000# SOURCE: BUFFALO	1	31,000	3 %	300	32,200	32,200	4.00	4.10	4.03	\$129,800
	V5 COMP K.O. DRUM 650 F 2670 PSIG 4.5' X 10' CS/CS W/SKIRT 49,000# SOURCE: EST.	1	180,500	3 %	18,100	204,000	204,000	4.00	4.10	4.03	\$822,100
	V6 AMINE ACCUM 650 F 75 PSIG 3' X 15' SA516-70 W/SDLS 1,750# SOURCE: EST.	1	8,000	3 %	800	9,000	9,000	4.00	4.10	4.03	\$36,300
	V7 AMINE FLASH DRUM 650 F 200 PSIG 5' X 20' SA516-70 W/SDLS 10,000# SOURCE: EST.	1	37,000	3 %	3,700	41,800	41,800	4.00	4.10	4.03	\$168,500
	V8 ATMOS TWR ACCUM 650 F 75 PSIG 6' X 24' CS W/SDLS 12,000# SOURCE: BUFFALO	1	1,195,000	3 %	12,000	1,242,900	1,242,900	4.00	4.10	4.03	\$5,008,900
	V9 150# STM DRUM H-1 650 F 300 PSIG 6' X 15' CS/CS W/SDLS 11,000# SOURCE: BUFFALO	1	24,000	3 %	200	24,900	24,900	4.00	4.10	4.03	\$100,300
	V10 150# STM DRUM H-2 650 F 200 PSIG 5' X 15' CS/CS W/SDLS 7,000# SOURCE: BUFFALO	1	15,300	3 %	1,500	17,300	17,300	4.00	4.10	4.03	\$69,700

SUN REFINING AND MARKETING CO.

P. E. T. S. FACTORED ESTIMATE WORKSHEET

DATE: 16-Mar-87

PROJECT: TURBINE FUELS PROJECT  
HYDROVISBREAKERESTIMATE NO: 7273  
JOB NO: 848768BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
	V11 150# STM DRUM H-3 650 F 200 PSIG 4.5' X 12' CS/CS W/SOLS 5,400# SOURCE: BUFFALO	1	11,900	3 %	1,200	13,500	13,500	4.00	4.10	4.03	\$54,400
	V12 AMINE SURGE TANK 650 F 75 PSIG 10' X 25' SAS16-70 W/SOLS 10,500# SOURCE: EST.	1	38,000	3 %	3,800	42,900	42,900	4.00	4.10	4.03	\$172,900
450C	K1 RECYCLE COMP. CENT. 1395 BHP 159,396 SCFM @ 1.37 K & 777 MOLE WT SUCT. 125 F @ 2440 PSIA DISCH. 131 F @ 2620 PSIA SPARES: ROTOR SOURCE: EST.	1	850,000	3 %	8,500	884,000	884,000	2.00	2.50	2.13	\$1,882,900
	K2 ATMOS TWR OFF-GAS COMP. 50 BHP 537 SCFM @ 1.159 K & 777 MOLE WT SUCT. 100 F @ 34.7 PSIA DISCH. 237 F @ 164.7 PSIA SPARES: SOURCE: EST.	2	280,000	3 %	2,800	291,200	582,400	2.00	2.50	2.13	\$1,240,500
450P	P1 FEED PUMP CS/12CR HOR 1444 GPM DISCH 2700 PSIG DELTA 2695 PSI-7123' PUMP TEMP 514 F @ 0.894 SG MOTOR HP 3500 @ 3560 RPM SOURCE: UNITED	2	360,000	3 %	3,600	374,400	748,800	4.00	7.00	4.75	\$3,556,800
	P2 ADD & MIX CS/12CR HOR 100 GPM DISCH 30 PSIG DELTA 30 PSI - 88' PUMP TEMP 90 F @ 0.79 SG MOTOR HP 5 @ 3560 RPM SOURCE: UNITED	2	10,200	3 %	100	10,600	21,200	4.00	7.00	4.75	\$100,700
	P3,4,5 REA MIX 12CR/12CR H 14280 GPM DISCH 2515 PSIG DELTA 5 PSI - 16' PUMP TEMP 850 F @ 0.74 SG MOTOR HP 100 @ 3600 RPM SOURCE:	6	75,600	3 %	800	78,700	472,200	4.00	7.00	4.75	\$2,243,000
	P6 ATMOS TWR PA CS/12CR HOR 495 GPM DISCH 79 PSIG DELTA 46 PSI-152' PUMP TEMP 515 F @ 0.878 SG MOTOR HP 30 @ 3560 RPM SOURCE: UNITED	2	15,050	3 %	200	15,700	31,400	4.00	7.00	4.75	\$149,200
	P7 LEAN AMINE CS/12CR HOR 440 GPM DISCH 2480 PSIG DELTA 2480 PSI-5787' PUMP TEMP 145 F @ 0.99 SG MOTOR HP 1100 @ 3560 RPM SOURCE:	2	295,000	3 %	3,000	306,900	613,800	4.00	7.00	4.75	\$2,915,600



PROJECT: TURBINE FUELS PROJECT  
HYDROVISBREAKER

ESTIMATE NO: 7273

JOB NO: 848768

BY: J. T. HARLAN

FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
	P8 AMINE STILL RFX CS/12CR H 40 GPM DISCH 50 PSIG DELTA 50 PSI - 117' PUMP TEMP 105 F @ 0.99 SG MOTOR HP 4 @ 3600 RPM SOURCE:	2	5,000	3 %	100	5,300	10,600	4.00	7.00	4.75	\$50,400
	P9 ATMOS TWR OH CS/12CR H 519 GPM DISCH 76 PSIG DELTA 66 PSI - 204' PUMP TEMP 100 F @ 0.751 SG MOTOR HP 40 @ 3600 RPM SOURCE: UNITED	2	14,070	3 %	100	14,600	29,200	4.00	7.00	4.75	\$138,700
	P10 ATMOS TWR BTMS CS/12CR H 895 GPM DISCH 77 PSIG DELTA 78 PSI - 224' PUMP TEMP 619 F @ 0.806 SG MOTOR HP 75 @ 3600 RPM SOURCE: UNITED	2	19,320	3 %	200	20,100	40,200	4.00	7.00	4.75	\$191,000
	P11 ATMOS TWR DIST CS/12CR H 274 GPM DISCH 67 PSIG DELTA 44 PSI - 146' PUMP TEMP 543 F @ 0.694 SG MOTOR HP 15 @ 3600 RPM SOURCE: UNITED	2	14,630	3 %	100	15,200	30,400	4.00	7.00	4.75	\$144,400
	P12 VAC TWR DIST CS/12CR H 564 GPM DISCH 67 PSIG DELTA 81 PSI - 227' PUMP TEMP 542 F @ 0.824 SG MOTOR HP 50 @ 3600 RPM SOURCE: UNITED	2	16,330	3 %	200	17,000	34,000	4.00	7.00	4.75	\$161,500
	P13 VAC TWR PA CS/12CR H 792 GPM DISCH 51 PSIG DELTA 66 PSI - 185' PUMP TEMP 542 F @ 0.824 SG MOTOR HP 50 @ 3600 RPM SOURCE: UNITED	2	17,400	3 %	200	18,100	36,200	4.00	7.00	4.75	\$172,000
	P14 VAC TWR BTMS CS/12CR H 187 GPM DISCH 133 PSIG DELTA 47 PSI - 377' PUMP TEMP 631 F @ 0.901 SG MOTOR HP 40 @ 3600 RPM SOURCE: UNITED	2	17,070	3 %	200	17,800	35,600	4.00	7.00	4.75	\$169,100
	P15 VAC TWR OH WAT CS/12CR H 31 GPM DISCH 56 PSIG DELTA 70 PSI - 163' PUMP TEMP 100 F @ 0.99 SG MOTOR HP 5 @ 3600 RPM SOURCE: UNITED	2	10,000	3 %	100	10,400	20,800	4.00	7.00	4.75	\$98,800
	P16 VAC TWR OH DIS CS/12CR H 106GPM DISCH 56 PSIG DELTA 70 PSI - 180' PUMP TEMP 100 F @ 0.902 SG MOTOR HP 15 @ 3600 RPM SOURCE: UNITED	2	10,780	3 %	100	11,200	22,400	4.00	7.00	4.75	\$106,400
480	STEAM JET VACUUM SYS 665 GPM FLOW SOURCE: CROLL REYNOLDS	1	15,000	3 %	200	15,700	15,700	2.00	3.00	2.25	\$35,300

PROJECT: TURBINE FUELS PROJECT  
 DISTILLATE HYDROTREATER  
 ESTIMATE NO: 7281  
 JOB NO: 848768

BY: J.T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS	1					93,500				\$376,000
300R	REACTORS	3					12,792,000				\$51,168,000
320A	AIR COOLED HEAT EXCHANGERS	1					124,800				\$540,400
320S	SHELL & TUBE HEAT EXCHANGERS	16					5,772,400				\$22,108,400
370H	FIRED HEATERS										
400T	TANKS										
400V	VESSELS	7					991,100				\$3,983,100
450C	COMPRESSORS	1					811,200				\$1,727,900
450P	PUMPS	6					984,200				\$4,675,000
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE M.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T PROCESS SYS. RALPH M. PARSONS CO. SYSTEM										
A	TOTAL MAJOR EQUIPMENT	35					21,569,200	3.92			\$84,576,800
600	INSTRUMENTS		15 % M.E.				3,235,400	4.00			\$12,941,600
620	INSULATION										
	ROUNDING										(\$=00)
A+B	TOTAL DIRECT INSTALLED COST										\$97,518,000
	HOME OFFICE COSTS		15 % D.I.C.								\$14,628,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$112,146,000
	CONTINGENCY		25 %								\$28,037,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/0/85								\$140,183,000
	ESCALATION		% /YR. FOR		YEAR						
	TOTAL INSTALLED COST		4/0/85								\$140,183,000

PROJECT: TURBINE FUELS PROJECT  
DISTILLATE HYDROTREATER

ESTIMATE NO: 7281

JOB NO: 848768

BY: J.T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	FACTOR HIGH	FACTOR USED	INSTALLED COST
270P	T1 FEED STRIPPER 650 F 200 PSIG 5.5' X 42' CS/CS W/SKIRT 24,600# SOURCE: BRIGHTON TRAYS: VALVE TYPE 12CR QTY 16 SOURCE: BRIGHTON	1	85,000	3 %	900	93,500	93,500	4.00	4.00	4.00	\$374,000
300R	DIST HTR REACTORS 850 F 2805 PSIG 10' X 68' 2.25CR W/O'LAY 1,138,820# SOURCE: C.B.I. SUPPORT GRIDS QTY 2 SOURCE: C.B.I.	3	4,100,000	3 %	41,000	4,264,000	12,792,000	4.00	4.00	4.00	\$51,168,000
320A	E6 EFF. COOLER. 300 F 2710 PSIG 1 - 18' X 40' BAY CS SOURCE:	1	120,000	3 %	1,200	124,800	124,800	4.10	5.00	4.33	\$540,400
320S	E1 R FEED EFF. BEU 700 F 2750 PSIG 7945 SF 2.25CR W/347 OL/347 SOURCE: HTE	2	1,900,000	3 %	19,000	1,976,000	3,952,000	3.50	4.80	3.83	\$15,136,200
	E2 R EFF 450# STM GEN BKU 700 F 2740 PSIG 7950 SF CS/347 SOURCE: HTE	1	172,200	3 %	1,700	179,100	179,100	3.50	4.80	3.83	\$686,000
	E3 RECY EFF EX BEU 550 F 2720 PSIG 4223 SF 2.25CR W/347 OL/347 SOURCE: HTE	1	1,055,700	3 %	10,600	1,098,000	1,098,000	3.50	4.80	3.83	\$4,205,300
	E4 R EFF 50# STM GEN BKU 550F 2720 PSIG 7460 SF CS/347 SOURCE: HTE	1	80,800	3 %	800	84,000	84,000	3.50	4.80	3.83	\$321,700
	E5 #1 RECYCLE EFF EX BEU 400F 2715 PSIG 2170 SF CS/347 SOURCE: HTE	2	27,800	3 %	300	28,900	57,800	3.50	4.80	3.83	\$221,400
	E7 EFF TRIM CLR BED 300F 2690 PSIG 6600 SF CS/MONEL SOURCE: HTE	1	84,500	3 %	800	87,800	87,800	3.50	4.80	3.83	\$335,300
	E8 STRIPPER ON COND AEU 400F 115 PSIG 865 SF CS/ADM SOURCE: HTE	1	11,100	3 %	100	11,500	11,500	3.50	4.80	3.83	\$44,000
	E9 STRIPPER FD BOTTOMS AES 650F 320 PSIG 4190 SF CS/CS SOURCE: HTE	6	45,400	3 %	500	47,300	283,800	3.50	4.80	3.83	\$1,087,000
	E10 STRIPPER FD PRENTR BEU 650F 500 PSIG 1632 SF CS/CS SOURCE: HTE	1	17,700	3 %	200	18,400	18,400	3.50	4.80	3.83	\$70,500

PROJECT: TURBINE FUELS PROJECT  
DISTILLATE HYDROTREATER  
ESTIMATE NO: 7281  
JOB NO: 848768

BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	HIGH	USED	INSTALLED COST
400V	V1 FEED TANK 650 F 30 PSIG 12' X 50' CS/CS W/LEGS 52,000# SOURCE: BUFFALO	1	53,000	3 %	500	55,100	55,100	3.50	4.80	3.83	\$211,000
	V2 HI-PRESSURE SEP. 650 F 2685 PSIG 8' X 24' CS/CS W/LEGS 320,000# SOURCE: C.B.I.	1	730,000	3 %	7,300	759,200	759,200	4.00	4.10	4.03	\$3,059,600
	V3 LO-PRESSURE SEP. 650 F 200 PSIG 8' X 24' CS/CS W/LEGS 27,000# SOURCE: BUFFALO	1	29,000	3 %	300	30,200	30,200	4.00	4.10	4.03	\$121,700
	V4 COMP. SUCT. KO 650 F 2670 PSIG 3.5' X 10' CS/CS W/SKIRT 57,000# SOURCE: C.B.I.	1	130,800	3 %	1,300	136,000	136,000	4.00	4.10	4.03	\$548,100
	V5 STRIPPER ACCLM. 650 F 200 PSIG 4' X 10' CS/CS W/LEGS 3,300# SOURCE: BUFFALO	1	3,000	3 %	300	3,400	3,400	4.00	4.10	4.03	\$13,700
	V6 STEAM DRUM 650 F 200 PSIG 4.5' X 10' CS/CS W/LEGS 4,500# SOURCE: BUFFALO	1	4,200	3 %	400	4,700	4,700	4.00	4.10	4.03	\$18,900
	V7 STEAM DRUM 650 F 200 PSIG 3.5' X 8' CS/CS W/LEGS 2,500# SOURCE: BUFFALO	1	2,200	3 %	200	2,500	2,500	4.00	4.10	4.03	\$10,100
450C	K1 RECYCLE COMP. CENT. 777 BHP 106,439 SCFM @ 1.393 K & 777 MOLE WT SUCT. 123 F @ 2450 PSIA DISCH. 136 F @ 2650 PSIA SPARES: ROTOR, CPLG, SHAFT SOURCE:	1	780,000	3 %	7,800	811,200	811,200	2.00	2.50	2.13	\$1,727,900
450P	P1 FEED PUMP CS/12CR HOR 1165 GPM DISCH 2650 PSIG DELTA 2650 PSI - 7918' PUMP TEMP 470 F @ 0.795 SG MOTOR HP 3000 @ 7777 RPM SOURCE:	2	440,000	3 %	4,400	457,600	915,200	4.00	7.00	4.75	\$4,347,200
	P2 STRIP'R FD. CS/12CR HOR 1008 GPM DISCH 235 PSIG DELTA 94 PSI - 255' PUMP TEMP 165 F @ 0.847 SG MOTOR HP 200 @ 3600 RPM SOURCE:	2	30,700	3 %	300	31,900	63,800	4.00	7.00	4.75	\$303,100
	P3 STRPR REFLX CS/12CR HOR 30 GPM DISCH 200 PSIG DELTA 55 PSI - 199' PUMP TEMP 110 F @ 0.64 SG MOTOR HP 3 @ 3600 RPM SOURCE:	2	2,500	3 %		2,600	5,200	4.00	7.00	4.75	\$24,700

PROJECT: TURBINE FUELS PROJECT  
 DISTILLATE HYDROCRACKER  
 ESTIMATE NO: 7285  
 JOB NO: 848768

BY: J. T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS	4					767,100				\$3,068,400
300R	REACTORS	1					3,120,000				\$12,480,000
320A	AIR COOLED HEAT EXCHANGERS	3					457,600				\$1,981,400
320S	SHELL & TUBE HEAT EXCHANGERS	24					6,542,200				\$25,056,600
370N	FIRED HEATERS	2					1,716,000				\$3,432,000
400T	TANKS										
400V	VESSELS	7					1,101,300				\$4,438,200
450C	COMPRESSORS	1					728,000				\$1,550,600
450P	PUMPS	18					977,200				4,641,900
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T PROCESS SYS. RALPH M. PARSONS CO. SYSTEM										
A	TOTAL MAJOR EQUIPMENT	60					15,409,400		3.68		\$56,649,100
600	INSTRUMENTS		15 % M.E.				2,311,400		4.00		\$9,245,600
620	INSULATION										\$300
A+B	TOTAL DIRECT INSTALLED COST										\$65,895,000
	HOME OFFICE COSTS		15 % D.I.C.								\$9,884,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$75,779,000
	CONTINGENCY		25 %								\$18,945,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/8/85								\$94,724,000
	ESCALATION			% /YR. FOR	YEAR						
	TOTAL INSTALLED COST		4/8/85								\$94,724,000

PROJECT: TURBINE FUELS PROJECT  
 DISTILLATE HYDROCRACKER  
 ESTIMATE NO: 7285  
 JOB NO: 848768

BY: J. T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST
								LOW	HIGH	USED	
270P	T1 PREFRACTIONATOR 650 F 200 PSIG 8.5&10'X32&42' CS/CS W/SK 114,000# SOURCE: BRIGHTON TRAYS: VALVE TYPE 12CR QTY 16&16 SOURCE: BRIGHTON	1	156,000	3 %	1,600	178,300	178,300	4.00	4.00	4.00	\$713,200
	T2 MAIN FRACTIONATOR 650 F 41 PSIG 11&12.5&20&10'X24&8&4&2&30' CS/CS W/SKIRT 190,000# SOURCE: BRIGHTON TRAYS: VAL TYPE 12CR QTY 10&3&6&1&8&8 SOURCE: BRIGHTON	1	395,000	3 %	4,000	480,900	480,900	4.00	4.00	4.00	\$1,923,600
	T3 DEHEXAMIZER 650 F 50 PSIG 6&8'X30&24' CS/CS W/SK 25,000# SOURCE: BRIGHTON TRAYS: VALVE TYPE 12CR QTY 13&7 SOURCE: BRIGHTON	1	87,000	3 %	900	95,500	95,500	4.00	4.00	4.00	\$382,000
	S1 MN. FRAC. STRIPPER 650 F 100 PSIG 5'X 23' CS/CS W/SK 9,600# SOURCE: BRIGHTON TRAYS: VALVE TYPE 12CR QTY 8 SOURCE: BRIGHTON	1	11,000	3 %	100	12,400	12,400	4.00	4.00	4.00	\$49,600
300R	HYDROCR'R REACTORS 850 F 2805 PSIG 8' X 80' 2.25CR 840,000# SOURCE: CBI SUPPORT GRIDS QTY 4 SOURCE: CBI	1	3,000,000	3 %	30,000	3,120,000	3,120,000	4.00	4.00	4.00	\$12,480,000
320A	E11 MN FRAC PA CLR. 650 F 75 PSIG 1 - 12' X 40' BAY CS SOURCE: YUBA	1	120,000	3 %	1,200	124,800	124,800	4.10	5.00	4.33	\$540,400
	E15 MN FRAC TURB FUEL CLR. 650 F 100 2 - 16' X 36' BAY CS SOURCE:	2	160,000	3 %	1,600	166,400	332,800	4.10	5.00	4.33	\$1,441,000
320S	E1 R FEED EFF. BEU 800 F 2720 PSIG 5500 SF SCR 1/2MO/347 SOURCE: HTE QUOTE	3	1,333,000	3 %	13,300	1,386,300	4,158,900	3.50	4.80	3.83	\$15,928,600
	E2 #1 H TO EFF. EX. BEU 700 F 2740 PSIG 1700 SF 1CR 1/2 MO /347 SOURCE: HTE QUOTE	2	425,000	3 %	4,300	442,100	884,200	3.50	4.80	3.83	\$3,386,500
	E3 450# STM GEN BKU 650 F 2585 PSIG 3000 SF CS/CS SOURCE: HTE QUOTE	1	65,000	3 %	700	67,700	67,700	3.50	4.80	3.83	\$259,300
	E4 #2 H TO EFF. EX. BEU 650 F 2775 PSIG 1230 SF CS/CS SOURCE: HTE QUOTE	1	139,000	3 %	1,400	144,600	144,600	3.50	4.80	3.83	\$553,800
	E5 MN FRAC FD EX. BEU 2575 PSIG 2750 SF CS/CS SOURCE: HTE QUOTE	2	49,000	3 %	500	51,000	102,000	3.50	4.80	3.83	\$390,700
	E6 LP FL TO EFF EX. BEU 2565 PSIG 6050 SF CS/CS SOURCE: HTE QUOTE	4	96,500	3 %	1,000	100,400	401,600	3.50	4.80	3.83	\$1,538,100

PROJECT: TURBINE FUELS PROJECT  
DISTILLATE HYDROCRACKER  
ESTIMATE NO: 7285  
JOB NO: 848768

BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	HIGH	USED	INSTALLED COST
	E7 RX EFF CLR BEU 650F 254S PSIG 7190 SF CS/ADMIRALTY SOURCE: HTE QUOTE	2	138,000	3 %	1,400	143,500	287,000	3.50	4.80	3.83	\$1,099,200
	E8 PREFRAC COND BEU 400F 200 PSIG 3750 SF CS/ADMIRALTY SOURCE: HTE QUOTE	1	48,000	3 %	500	49,900	49,900	3.50	4.80	3.83	\$191,100
	E9 PREFRAC REBLR BKU 650F 500 PSIG 5400 SF CS/CS SOURCE: HTE QUOTE	2	58,500	3 %	600	60,900	121,800	3.50	4.80	3.83	\$466,500
	E10 MM FRAC OMD COND BEU 650F 100 PSIG 4343 SF CS/ADMIRALTY SOURCE: HTE QUOTE	2	57,500	3 %	600	59,800	119,600	3.50	4.80	3.83	\$458,100
	E12 DEHEX OMD COND BEU 650F 100 PSIG 4900 SF CS/ADMIRALTY SOURCE: HTE QUOTE	2	61,000	3 %	600	63,400	126,800	3.50	4.80	3.83	\$485,600
	E13 DEHEX REBLR BKU 650F 200 PSIG 1526 SF CS/CS SOURCE: HTE QUOTE	1	26,500	3 %	300	27,600	27,600	3.50	4.80	3.83	\$105,700
	E14 DEHEX BTM CLR BEU 650F 100 PSIG 3900 SF CS/ADMIRALTY SOURCE: HTE QUOTE	1	48,500	3 %	500	50,500	50,500	3.50	4.80	3.83	\$193,400
370H	H1 RECYCLE GAS HTR 13.28 MM BTU/HR SOURCE SELAS	1	250,000	3 %	2,500	260,000	260,000	2.00	2.00	2.00	\$520,000
	H2 MAIN FRAC REBLR 157.27 MM BTU/HR SOURCE SELAS	1	1,400,000	3 %	14,000	1,456,000	1,456,000	2.00	2.00	2.00	\$2,912,000
400V	V1 FEED TANK 650 F 175 PSIG 12' X 50' CS/CS W/LEGS 96,000# SOURCE: CBI	1	150,000	3 %	1,500	156,000	156,000	4.00	4.10	4.03	\$628,700
	V2 HI-PRESSURE SEP. 650 F 2530 PSIG 8' X 24' CS/CS W/LEGS 340,000# SOURCE: CBI	1	780,000	3 %	7,800	811,200	811,200	4.00	4.10	4.03	\$3,269,100
	V3 LO-PRESSURE SEP. 650 F 200 PSIG 8' X 24' CS/CS W/LEGS 27,000# SOURCE: BUFFALO	1	28,800	3 %	300	30,000	30,000	4.00	4.10	4.03	\$120,900
	V4 COMP. SUCT. KO 650 F 2530 PSIG 3' X 10' CS/CS W/SKIRT 25,000# SOURCE: CBI	1	55,000	3 %	600	57,300	57,300	4.00	4.10	4.03	\$250,900
	V5 PREFRAC. ACCUM. 650 F 200 PSIG 8' X 24' CS/CS W/LEGS 27,000# SOURCE: BUFFALO	1	28,800	3 %	300	30,000	30,000	4.00	4.10	4.03	\$120,900
	V6 MAIN FRAC ACCUM 650 F 70 PSIG 8' X 24' CS/CS W/SDLS 11,600# SOURCE: BUFFALO	1	10,000	3 %	100	10,400	10,400	4.00	4.10	4.03	\$41,900
	V7 DEHEX. ACCUM 650 F 40 PSIG 5' X 17.5' CS/CS W/SDLS 5,500# SOURCE: BUFFALO	1	6,100	3 %	100	6,400	6,400	4.00	4.10	4.03	\$25,800

PROJECT: TURBINE FUELS PROJECT  
DISTILLATE HYDROCRACKERESTIMATE NO: 7285  
JOB NO: 848768BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST
								LOW	HIGH	USED	
450C	K1 RECYCLE COMP. CENT. 1133 BHP 91,939 SCFM @ 1.37 K & 4.0 MOLE WT SUCTION 110 F @ 2305 PSIA DISCH. 125 F @ 2535 PSIA SPARES: ROTOR, CPLG, SHAFT SOURCE: IR	1	700,000	3 %	7,000	728,000	728,000	2.00	2.50	2.13	\$1,550,600
450P	P1 PREFR FD CS/12CR HOR 1686 GPM DISCH 255 PSIG DELTA 105 PSI - 33' PUMP TEMP 110 F @ 0.733 SG MOTOR HP 150 @ 7777 RPM SOURCE: UNITED	2	23,000	3 %	200	23,900	47,800	4.00	7.00	4.75	\$227,100
	P2 PREFR REF CS/12CR HOR 906 GPM DISCH 208 PSIG DELTA 63 PSI - 288' PUMP TEMP 167 F @ 0.505 SG MOTOR HP 50 @ 7777 RPM SOURCE: UNITED	2	15,300	3 %	200	16,000	32,000	4.00	7.00	4.75	\$152,000
	P3 HW FR REF&PR CS/12CR HOR 1064 GPM DISCH 114 PSIG DELTA 109 PSI - 370' PUMP TEMP 154 F @ 0.68 SG MOTOR HP 100 @ 7777 RPM SOURCE: UNITED	2	21,100	3 %	200	21,900	43,800	4.00	7.00	4.75	\$208,100
	P4 HW FR PA CS/12CR HOR 860 GPM DISCH 53 PSIG DELTA 40 PSI - 138' PUMP TEMP 339 F @ 0.670 SG MOTOR HP 30 @ 7777 RPM SOURCE: UNITED	2	16,900	3 %	200	17,600	35,200	4.00	7.00	4.75	\$167,200
	P5 HW FR REBOIL CS/12CR HOR 5799 GPM DISCH 115 PSIG DELTA 100 PSI - 345' PUMP TEMP 598 F @ 0.670 SG MOTOR HP 500 @ 7777 RPM SOURCE: UNITED	2	54,300	3 %	500	56,400	112,800	4.00	7.00	4.75	\$535,800
	P6 RX FEED CS/12CR HOR 1833 GPM DISCH 2640 PSIG DELTA 2625 PSI - 4320' PUMP TEMP 588 F @ 0.68 SG MOTOR HP 2000 @ 7777 RPM SOURCE: UNITED	2	294,400	3 %	2,900	306,100	612,200	4.00	7.00	4.75	\$2,908,000
	P7 TURB FUEL CS/12CR HOR 971 GPM DISCH 80 PSIG DELTA 63 PSI - 214' PUMP TEMP 322 F @ 0.680 SG MOTOR HP 60 @ 7777 RPM SOURCE: UNITED	2	19,300	3 %	200	20,100	40,200	4.00	7.00	4.75	\$191,000
	P8 DEHX OH R&P CS/12CR HOR 385 GPM DISCH 55 PSIG DELTA 40 PSI - 145' PUMP TEMP 112 F @ 0.6355 SG MOTOR HP 15 @ 7777 RPM SOURCE: UNITED	2	11,200	3 %	100	11,600	23,200	4.00	7.00	4.75	\$110,200
	P9 DEHX BTMS PR CS/12CR HOR 712 GPM DISCH 60 PSIG DELTA 40 PSI - 140' PUMP TEMP 242 F @ 0.660 SG MOTOR HP 30 @ 7777 RPM SOURCE: UNITED	2	14,500	3 %	100	15,000	30,000	4.00	7.00	4.75	\$142,500



SUN REFINING AND MARKETING CO.

P. E. T. S. FACTORED ESTIMATE WORKSHEET

DATE: 16-Mar-87

PROJECT: TURBINE FUELS PROJECT  
GAS PLANT

ESTIMATE NO: 7286  
JOB NO: 848768

BY: J.T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS	3					626,300				\$2,505,200
300R	REACTORS										
320A	AIR COOLED HEAT EXCHANGERS										
320S	SHELL & TUBE HEAT EXCHANGERS	11					499,300				\$1,912,200
370H	FIRED HEATERS										
400T	TANKS										
400V	VESSELS	4					57,100				\$230,100
450C	COMPRESSORS										
450P	PUMPS	14					218,200				\$1,036,600
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. M-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T PROCESS SYS. RALPH M. PARSONS CO. SYSTEM										
A	TOTAL MAJOR EQUIPMENT	32					1,400,900		4.06		\$5,684,100
600	INSTRUMENTS	15	% M.E.				210,100		4.00		\$840,400
620	INSULATION										
	ROUNDING										\$500
A+B	TOTAL DIRECT INSTALLED COST										\$6,525,000
	HOME OFFICE COSTS	15	% D.I.C.								\$979,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$7,504,000
	CONTINGENCY	25	%								\$1,876,000
	ROYALTIES										
	TOTAL INSTALLED COST	4/4/85									\$9,380,000
	ESCALATION		% /YR. FOR	YEAR							
	TOTAL INSTALLED COST	4/4/85									\$9,380,000

PROJECT: TURBINE FUELS PROJECT  
GAS PLANTESTIMATE NO: 7286  
JOB NO: 848768BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
270P	T1 ABSORBER STRIPPER 650 F 166 PSIG 6'89"X50&57" CS/CS W/SK 94,437# SOURCE: BRIGHTON TRAYS: VALVE TYPE 12CR QTY 22&20 SOURCE: BRIGHTON	1	225,000	3 %	2,300	261,100	261,100	4.00	4.00	4.00	\$1,044,400
	T2 DEBUTANIZER 650 F 130 PSIG 8.5' X 98' CS/CS W/SKIRT 79,236# SOURCE: BRIGHTON TRAYS: VAL TYPE 12CR QTY 43 SOURCE: BRIGHTON	1	292,000	3 %	2,900	327,700	327,700	4.00	4.00	4.00	\$1,310,800
	T3 DEBUT SIDE STRIP'R 650 F 128PSIG 3.5'X 24' CS/CS W/SK 6,507# SOURCE: BRIGHTON TRAYS: VALVE TYPE 12CR QTY 9 SOURCE: BRIGHTON	1	35,000	3 %	400	37,500	37,500	4.00	4.00	4.00	\$150,000
320S	E1 ABS'R ON COND BEU 400 F 156 PSIG 3341 SF CS / ADMIRALTY SOURCE: HTE QUOTE	1	38,000	3 %	400	39,500	39,500	3.50	4.80	3.83	\$151,300
	E2 ABS'R INTERCLR BEU 400 F 164 PSIG 5830 SF CS / ADMIRALTY SOURCE: HTE QUOTE	2	79,000	3 %	800	82,200	164,400	3.50	4.80	3.83	\$629,700
	E3 STRIPPER INTERHEATER BEU 650 F 165 PSIG 1607 SF CS/CS SOURCE: HTE QUOTE	1	17,500	3 %	200	18,200	18,200	3.50	4.80	3.83	\$69,700
	E4 STRIPPER REBOILER BKU 650 F 166 PSIG 2639 SF CS/CS SOURCE: HTE QUOTE	1	31,000	3 %	300	32,200	32,200	3.50	4.80	3.83	\$123,300
	E5 DEBUT FEED/BOTTOMS BEU 140 PSIG 6297 SF CS/CS SOURCE: HTE QUOTE	1	40,000	3 %	400	41,600	41,600	3.50	4.80	3.83	\$159,300
	E6 DEBUT ON COND. BEU 120 PSIG 6889 SF CS / ADMIRALTY SOURCE: HTE QUOTE	1	66,500	3 %	700	69,200	69,200	3.50	4.80	3.83	\$265,000
	E7 DEBUT ON PROD COND BEU 400F 2545 PSIG 2109 SF CS / ADMIRALTY SOURCE: HTE QUOTE	1	27,000	3 %	300	28,100	28,100	3.50	4.80	3.83	\$107,600
	E8 DEBUT SIDE STRIP REBLR BKU 650 F 128 PSIG 465 SF CS/CS SOURCE: HTE QUOTE	1	13,500	3 %	100	14,000	14,000	3.50	4.80	3.83	\$53,800
	E9 DEBUT REBLR BKU 650F 500 PSIG 4121 SF CS/CS SOURCE: HTE QUOTE	1	36,000	3 %	400	37,500	37,500	3.50	4.80	3.83	\$143,800
	E10 LEAN OIL CLR BEU 400F 115 PSIG 4947 SF CS/ADMIRALTY SOURCE: HTE QUOTE	1	52,500	3 %	500	54,600	54,600	3.50	4.80	3.83	\$209,100

PROJECT: TURBINE FUELS PROJECT  
GAS PLANTESTIMATE NO: 7286  
JOB NO: 848768BY: J. T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST
								LOW	HIGH	USED	
400V	V1 ABS'R ON ACCUM 650 F 150 PSIG 6.5' X 21' CS/CS W/SDLS 13,000# SOURCE: BUFFALO	1	11,200	3 %	100	11,600	11,600	4.00	4.10	4.03	\$46,700
	V2 INTERCOOLER SEP. 650 F 162 PSIG 9' X 20' CS/CS W/SDLS 26,000# SOURCE: BUFFALO	1	28,600	3 %	300	29,800	29,800	4.00	4.10	4.03	\$120,100
	V3 DEBUT REFLUX ACC 650 F 115 PSIG 7' X 17' CS/CS W/SDLS 10,500# SOURCE: BUFFALO	1	10,000	3 %	100	10,400	10,400	4.00	4.10	4.03	\$41,900
	V3 DEBUT ON PROD ACC 650 F 110 PSIG 5' X 12' CS/CS W/SDLS 4,500# SOURCE: BUFFALO	1	5,000	3 %	100	5,300	5,300	4.00	4.10	4.03	\$21,400
450P	P1 ABSORB REFLUX CS/12CR HOR 717 GPM DISCH 204 PSIG DELTA 77 PSI - 233' PUMP TEMP 100 F @ 0.7315 SG MOTOR HP 60 @ 7777 RPM SOURCE: UNITED	2	19,300	3 %	200	20,100	40,200	4.00	7.00	4.75	\$191,000
	P2 STRIP REFLUX CS/12CR HOR 1297 GPM DISCH 193 PSIG DELTA 56 PSI - 195' PUMP TEMP 100 F @ 0.622 SG MOTOR HP 73 @ 7777 RPM SOURCE: UNITED	2	17,100	3 %	200	17,800	35,600	4.00	7.00	4.75	\$169,100
	P3 STRIP'R BTMS CS/12CR HOR 1187 GPM DISCH 158 PSIG DELTA 15 PSI - 60' PUMP TEMP 278 F @ 0.5789 SG MOTOR HP 15 @ 7777 RPM SOURCE: UNITED	2	11,200	3 %	100	11,600	23,200	4.00	7.00	4.75	\$110,200
	P4 DEBUT REFLUX CS/12CR HOR 738 GPM DISCH 154 PSIG DELTA 61 PSI - 273' PUMP TEMP 135 F @ 0.5170 SG MOTOR HP 45 @ 7777 RPM SOURCE: UNITED	2	15,300	3 %	200	16,000	32,000	4.00	7.00	4.75	\$152,000
	P5 BUTANE PROD. CS/12CR HOR 252 GPM DISCH 205 PSIG DELTA 117 PSI - 521' PUMP TEMP 127 F @ 0.5188 SG MOTOR HP 30 @ 7777 RPM SOURCE: UNITED	2	16,900	3 %	200	17,600	35,200	4.00	7.00	4.75	\$167,200
	P6 DEBUT BTMS CS/12CR HOR 882 GPM DISCH 206 PSIG DELTA 98 PSI - 404' PUMP TEMP 411 F @ 0.5607 SG MOTOR HP 100 @ 7777 RPM SOURCE: UNITED	2	21,100	3 %	200	21,900	43,800	4.00	7.00	4.75	\$208,100
	P7 LEAN OIL M.U. CS/12CR HOR 36 GPM DISCH 154 PSIG DELTA 91 PSI - 316' PUMP TEMP 250 F @ 0.6643 SG MOTOR HP 5 @ 7777 RPM SOURCE: UNITED	2	4,000	3 %		4,100	8,200	4.00	7.00	4.75	\$39,000

PROJECT: TURBINE FUELS PROJECT  
 HYDROGEN PRODUCTION PLANTS - TWO @ 80 MMSCFD EACH  
 ESTIMATE NO: 7287  
 JOB NO: 848768

BY: J.T. HAPLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS						756,000		1.95		\$1,474,000
300R	REACTORS						2,142,000		1.95		\$4,177,000
320A	AIR COOLED HEAT EXCHANGERS										
320S	SHELL & TUBE HEAT EXCHANGERS						4,686,000		1.95		\$9,138,000
370H	FIRED HEATERS						10,800,000		1.95		\$21,060,000
400T	TANKS										
400V	VESSELS						921,000		1.95		\$1,796,000
450C	COMPRESSORS						7,480,000		1.95		\$14,586,000
450P	PUMPS						243,000		1.95		\$474,000
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.U.T PROCESS SYS. RALPH N. PARSONS CO. SYSTEM										
A	TOTAL MAJOR EQUIPMENT						27,028,000		1.95		\$52,705,000
600	INSTRUMENTS		15 % M.E.				4,054,200		4.00		\$16,216,800
620	INSULATION										
	ROUNDING										\$200
A+B	TOTAL DIRECT INSTALLED COST										\$68,922,000
	HOME OFFICE COSTS		15 % D.I.C.								\$10,338,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$79,260,000
	CONTINGENCY		25 %								\$19,815,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/0/85								\$99,075,000
	ESCALATION			% /YR. FOR	YEAR						
	TOTAL INSTALLED COST		4/0/85								\$99,075,000

SUN REFINING AND MARKETING CO.

P. E. T. S. FACTORED ESTIMATE WORKSHEET

DATE: 16-Mar-87

PROJECT: TURBINE FUELS PROJECT  
 HYDROGEN PURIFICATION UNIT  
 ESTIMATE NO: 7288  
 JOB NO: 848768

BY: J.T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS										
300R	REACTORS										
320A	AIR COOLED HEAT EXCHANGERS						127,000				\$486,400
320S	SHELL & TUBE HEAT EXCHANGERS	4									
370M	FIRED HEATERS										
400T	TANKS										
400V	VESSELS										
450C	COMPRESSORS	1					10,949,800				\$23,323,100
450P	PUMPS										
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T PROCESS SYS. RALPH M. PARSONS CO. SYSTEM	1					5,072,200				\$8,876,400
A	TOTAL MAJOR EQUIPMENT	6					16,149,000	2.02			\$32,685,900
600	INSTRUMENTS		15 % M.E.				2,422,400	4.00			\$9,689,600
620	INSULATION ROUNDING										\$500
A+B	TOTAL DIRECT INSTALLED COST										\$42,376,000
	HOME OFFICE COSTS		15 % D.I.C.								\$6,356,000
800	CATALYST & CHEMICALS										\$48,732,000
	SUB TOTAL										\$48,732,000
	CONTINGENCY		25 %								\$12,183,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/0/85								\$60,915,000
	ESCALATION			% /YR. FOR	YEAR						
	TOTAL INSTALLED COST		4/0/85								\$60,915,000

PROJECT: TURBINE FUELS PROJECT  
 HYDROGEN PURIFICATION UNIT  
 ESTIMATE NO: 7288  
 JOB NO: 848768

BY: J.T. HARLAM  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	HIGH	USED	INSTALLED COST
480	HYDROGEN PURIF. UNIT - UNION CARBIDE GAS RATE 74.9 MMSCFD HYDROGEN PRODUCT PURITY 99.0 % FEED GAS HYDROGEN RECOVERY 87.5 % PLANT FEED PRESSURE 800 PSIG TOTAL 147.6 LT/D	1	4,785,000	5 %	47,900	5,072,200	5,072,200	1.50	2.50	1.75	\$8,876,400
320S	E1 EXP FD PREHTR BEU 650 F 2640 PSIG 1,200 SF CS/CS SOURCE:	1	26,500	5 %	300	28,100	28,100	3.50	4.80	3.83	\$107,600
	E2 HYD. COMP. AFTERCLR BEU 400 F 2725 PSIG 1,872 SF CS/ADMIRALTY SOURCE:	1	46,800	5 %	500	49,600	49,600	3.50	4.80	3.83	\$190,000
	E3 BLEED GAS COMP INTERCLR BEU 400 F 100 PSIG 1,083 SF CS/ADMIRALTY SOURCE:	1	16,500	5 %	200	17,500	17,500	3.50	4.80	3.83	\$67,000
	E4 BLEED GAS COMP AFTERCLR BEU 400 F 190 PSIG 1,962 SF CS/ADMIRALTY SOURCE:	1	30,000	5 %	300	31,800	31,800	3.50	4.80	3.83	\$121,800
450C	GAS EXPANSION AND COMPRESSION UNIT CONSISTING OF THE FOLLOWING COMPRESSORS AND MOTOR:										
	K1 TURBOEXPANDER 9714 BHP 96.459MMSCFD @ 1.37K & 4.371MOLE WT. SUCT 300 F @ 2395 PSIG DISCH 108 F @ 800 PSIG	1	6,000,000	5 %	60,000	6,360,000	6,360,000	2.00	2.50	2.13	\$13,546,800
	K2 HYDROGEN COMPRESSOR 3513 BHP 51996 SCFM @ 1.287K & 2.156MOLE WT. SUCT 110 F @ 750 PSIG DISCH 340 F @ 2475 PSIG	2	1,000,000	5 %	10,000	1,060,000	2,120,000	2.00	2.50	2.13	\$4,515,500
	K3 BLEED GAS COMPRESSOR 3750 BHP T FIRST STAGE 2550 BHP 14989 SCFM @ 1.285K & 12.053MOLE WT. SUCT 110 F @ 0 PSIG DISCH 255 F @ 37 PSIG SECOND STAGE 2589 BHP 14989 SCFM @ 1.285K & 12.053MOLE WT. SUCT 110 F @ 32 PSIG DISCH 310 F @ 165 PSIG	1	2,330,000	5 %	23,300	2,469,800	2,469,800	2.00	2.50	2.13	\$5,260,700

SOURCE: K2 & K3 I.R. - K1 EST.

PROJECT: TURBINE FUELS PROJECT  
TANKAGE

ESTIMATE NO: 7289  
JOB NO: 848768

BY: J.T. KARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST
								LOW	HIGH	USED	
ESTIMATE SUMMARY											
270P	TOWERS										
300R	REACTORS										
320A	AIR COOLED HEAT EXCHANGERS										
320S	SHELL & TUBE HEAT EXCHANGERS										
370H	FIRED HEATERS										
400T	TANKS	18					9,589,300				\$22,822,600
400V	VESSELS										
450C	COMPRESSORS										
450P	PUMPS										
480	SPECIALIZED EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T PROCESS SYS. RALPH M. PARSON CO. SYSTEM										
A	TOTAL MAJOR EQUIPMENT	18					9,589,300		2.38		\$22,822,600
600	INSTRUMENTS		15 % M.E.				1,438,400		4.00		\$5,753,600
620	INSULATION		254,300 SQ. FT.			2.83	719,700		3.85		\$2,770,800
	ROUNDING										
A+B	TOTAL DIRECT INSTALLED COST										\$31,347,000
	HOME OFFICE COSTS		15 % D.I.C.								\$4,702,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$36,049,000
	CONTINGENCY		25 %								\$9,012,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/9/85								\$45,061,000
	ESCALATION		% /YR. FOR		YEAR						
	TOTAL INSTALLED COST		4/9/85								\$45,061,000

PROJECT: TURBINE FUELS PROJECT  
TANKAGEESTIMATE NO: 7289  
JOB NO: 848768BY: J. T. HARLAM  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
400	CRUDE TANKAGE OPER. TEMP. 150 F DIA. 142' HT. 56' ROOF=COV'D.FLOATER HEATED&INSULATED VOL. 157,956 BBLs.	5	631,864	5 %	15,800	679,300	3,396,500	2.00	3.50	2.38	\$8,083,700
	INTERMEDIATE HOT TANKAGE OPER. TEMP. 200 F (150 F CRUDE STOR) DIA. 100' HT. 56' ROOF=COV'D.FLOATER HEATED&INSULATED VOL. 439,823 CU FT.	3	362,695	5 %	9,100	389,900	1,169,700	2.00	3.50	2.38	\$2,783,900
	INTERMEDIATE COLD TANKAGE OPER. TEMP. 100 F (150 F CRUDE STOR) DIA. 82' HT. 32' ROOF=COV'D.FLOATER HEATED & INSULATED VOL. 30,099 BBLs.	2	245,306	5 %	6,100	263,700	527,400	2.00	3.50	2.38	\$1,255,200
	BUTANE PRODUCT SPHERE OPERATING PRESS. 50 PSIG @ 100 F DESIGN PRESS. 75 PSIG @ 650 F DIA. 65' VOL. 25,611 BBLs. SHELL THK. 1.250" WT. 678,610 LBS.	1	566,475	5 %	5,700	600,500	600,500	2.00	3.50	2.38	\$1,429,200
	AMMONIA STORAGE SPHERE OPERATING PRESS. 200 PSIG @ 100 F DESIGN PRESS. 225 PSIG @ 650 F DIA. 42' VOL. 39,000 CU FT SHELL THK. 2.250" WT. 510,000 LBS.	1	380,000	5 %	3,800	402,800	402,800	2.00	3.50	2.38	\$958,700
	NAPHTHA PRODUCT TANKAGE OPER. TEMP. 100 F DIA. 114' HT. 56' ROOF=COV'D.FLOATER VOLUME 101,805 BBLs.	1	545,675	5 %	5,500	578,500	578,500	2.00	3.50	2.38	\$1,376,800
	JET FUEL PRODUCT TANKAGE OPER. TEMP. 100 F DIA. 143' HT. 56' ROOF=COV'D.FLOATER VOLUME 160,188 BBLs.	4	640,752	5 %	6,400	679,200	2,716,800	2.00	3.50	2.38	\$6,466,000
	RESIDIUM PRODUCT TANKAGE OPER. TEMP. 450 F DIA. 81' HT. 32' ROOF=CONE ROOF HEATED&INSULATED VOL. 29,369 BBLs.	1	185,900	5 %	1,900	197,100	197,100	2.00	3.50	2.38	\$469,100



PROJECT: TURBINE FUELS PROJECT  
 STACK GAS SCRUBBING AND COLLECTING UNIT  
 ESTIMATE NO: 7291  
 JOB NO: 848768

BY: J.T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS										
300R	REACTORS										
320A	AIR COOLED HEAT EXCHANGERS										
320S	SHELL & TUBE HEAT EXCHANGERS										
370H	FIRED HEATERS										
400T	TANKS										
400V	VESSELS										
450C	COMPRESSORS										
450P	PUMPS										
480	SPECIALIZED EQUIPMENT										
	CROLL REYNOLDS STM JET SYS										
	UNION CARBIDE H.P.U.										
	W-L/DP/ACP/SRI SYSTEMS	10					7,843,900				\$22,590,700
	ELEC. PRECIP. & FLY ASH COLL. SYS	2					3,018,000				\$8,686,100
	CHEVRON W.W.T PROCESS SYS.										
	RALPH M. PARSONS CO. SYSTEM										
A	TOTAL MAJOR EQUIPMENT	12					10,859,900		2.88		\$31,276,800
600	INSTRUMENTS		15 % N.E.				1,629,000		4.00		\$6,516,000
620	INSULATION										
	ROUNDING										\$200
A+B	TOTAL DIRECT INSTALLED COST										\$37,793,000
	HOME OFFICE COSTS		15 % D.I.C.								\$5,669,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$43,462,000
	CONTINGENCY		25 %								\$10,866,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/0/85								\$54,328,000
	ESCALATION		% /YR. FOR		YEAR						
	TOTAL INSTALLED COST		4/0/85								\$54,328,000

PROJECT: TURBINE FUELS PROJECT  
 STACK GAS SCRUBBING AND COLLECTING UNIT  
 ESTIMATE NO: 7291  
 JOB NO: 84-8768

BY: J. T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	HIGH	USED	INSTALLED COST
480	RAW MATERIAL STORAGE HANDLING AND PREPARATION COMPLETE. (ELEVATOR BINS, FEEDER, TANK AGITATION, AND PUMPS)	1	110,700	3 %	2,800	116,800	116,800	2.50	4.00	2.88	\$336,400
	FLUE GAS COOLING (WINDTUNES)	1	584,400	3 %	15,800	1,647,700	1,647,700	2.50	4.00	2.88	\$4,745,400
	SO2 SCRUBBER SYSTEM (TRAY SCRUBBERS, DEMISTERS, TANKS, AGITATORS PUMPS)	1	2,918,500	3 %	29,200	3,035,300	3,035,300	2.50	4.00	2.88	\$8,741,700
	CHLORIDES AND SOLIDS PURGE SYSTEM (TANKS AND PUMPS)	1	11,500	3 %	100	11,900	11,900	2.50	4.00	2.88	\$34,300
	STACK GAS REHEAT SYSTEM	1	525,500	3 %	5,300	546,600	546,600	2.50	4.00	2.88	\$1,574,200
	SO2 REGENERATION SYSTEM (EVAPORATOR-CRYSTALLIZERS, HEATER, CONDENSERS, STRIPPERS, COMPRESSORS, TANKS, AGITATORS AND PUMPS)	1	679,000	3 %	6,800	706,200	706,200	2.50	4.00	2.88	\$2,033,900
	SO2 REDUCTION SYSTEM	1	654,000	3 %	6,500	680,100	680,100	2.50	4.00	2.88	\$1,958,700
	SULFUR STORAGE (PIT AND STORAGE TANK, PUMPS AND HEATERS)	1	75,500	3 %	800	78,600	78,600	2.50	4.00	2.88	\$226,400
	SODIUM SULFATE PURGE SYSTEM (CHILLER CRYSTALLIZER, CENTRIFUGE, ROTARY DRYER, TANKS, PUMPS, CONVEYING EQUIPMENT, BIN, AND LOADOUT)	1	314,500	3 %	3,100	327,000	327,000	2.50	4.00	2.88	\$941,800
	BOOSTER AND ID FANS	1	667,000	3 %	6,700	693,700	693,700	2.50	4.00	2.88	\$1,997,900
	SOURCE FOR THE ABOVE IS: WELLMAN-LORD/DAVY POWERGAS/ ALLIED CHEMICAL PROCESS/SRI										
	ELECTROSTATIC PRECIPITATION	1	2,500,000	3 %	25,000	2,600,000	2,600,000	2.50	4.00	2.88	\$7,488,000
	FLY ASH COLLECTION AND REMOVAL SYSTEM CONSISTING OF:	1	400,000	3 %	4,000	416,000	416,000	2.50	4.00	2.88	\$1,198,100
	SILO 12' DIA X 35'		80,000								
	SILO UNLOADING ROOM 12' DIA X 15'		20,000								
	BOTTOM STORAGE HOPPER 15'X15'X15'		40,000								
	CONVEYOR SYSTEM		260,000								

SOURCE: UNITED CONVEYOR

PROJECT: TURBINE FUELS PROJECT  
 LOW PRESSURE AMINE PLANT  
 ESTIMATE NO: 7292  
 JOB NO: 848768

BY: J.T. HARLAM  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS	2					211,700				846,800
300R	REACTORS										
320A	AIR COOLED HEAT EXCHANGERS										
320S	SHELL & TUBE HEAT EXCHANGERS	7					179,000				685,500
370M	FIRED HEATERS										
400T	TANKS										
400V	VESSELS	3					22,400				90,300
450C	COMPRESSORS										
450P	PUMPS	4					50,800				241,400
480	MISC. EQUIPMENT CROLL REYNOLDS STM JET SYS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELE. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T PROCESS SYS. RALPH W. PARSONS CO. SYSTEM										
A	TOTAL MAJOR EQUIPMENT	16					463,900		4.02		\$1,864,000
600	INSTRUMENTS		15 % M.E.				69,600		4.00		\$278,400
620	INSULATION										
	ROUNDING										(\$-00)
A+B	TOTAL DIRECT INSTALLED COST										\$2,142,000
	HOME OFFICE COSTS		15 % D.I.C.								\$321,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$2,463,000
	CONTINGENCY		25 %								\$616,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/0/85								\$3,079,000
	ESCALATION		% /YR. FOR		YEAR						
	TOTAL INSTALLED COST		4/0/85								\$3,079,000

PROJECT: TURBINE FUELS PROJECT  
 LOW PRESSURE AMINE PLANT  
 ESTIMATE NO: 7292  
 JOB NO: 848768

BY: J. T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR LOW	FACTOR HIGH	FACTOR USED	INSTALLED COST
270P	T1 AMINE CONTACTOR 650 F 175 PSIG 6.5' X 50' CS/CS W/SK 30,000# SOURCE: BRIGHTON TRAYS: VALVE TYPE 12CR QTY 20 SOURCE: BRIGHTON	1	136,500	3 %	1,400	147,400	147,400	4.00	4.00	4.00	\$589,600
	T2 STILL 650 F 75 PSIG 4' X 50' CS/CS W/SKIRT 11,000# SOURCE: BRIGHTON TRAYS: VAL TYPE 12CR QTY 20 SOURCE: BRIGHTON	1	59,300	3 %	600	64,300	64,300	4.00	4.00	4.00	\$257,200
320S	E1 LEAN/RICH EXCH AES 650 F 100 PSIG 900 SF CS / 304 SS SOURCE: ESTIMATED	2	22,000	3 %	200	22,900	45,800	3.50	4.80	3.83	\$175,400
	E2 STILL COND BEU 400 F 75 PSIG 680 SF CS / BIMETAL ADMIRALTY/CS SOURCE: ESTIMATED	1	24,500	3 %	200	25,400	25,400	3.50	4.80	3.83	\$97,300
	E3 LEAN AMINE CLR BEU 400 F 175 PSIG 762 SF CS / BIMETAL ADMIRALTY/CS SOURCE: ESTIMATED	2	26,500	3 %	300	27,600	55,200	3.50	4.80	3.83	\$211,400
	E4 STILL REBOILER BKU 650 F 175 PSIG 1810 SF CS/304 SS SOURCE: ESTIMATED	1	43,400	3 %	400	45,100	45,100	3.50	4.80	3.83	\$172,700
	E5 AMINE RECLAIMER BKU 200 PSIG 180 SF CS/18-8 SOURCE: ESTIMATED	1	7,200	3 %	100	7,500	7,500	3.50	4.80	3.83	\$28,700
400V	V1 FLASH TANK 650 F 175 PSIG 3.5' X 10' CS/CS W/LEGS 2,500# SOURCE: ESTIMATED	1	3,500	3 %		3,600	3,600	4.00	4.10	4.03	\$14,500
	V2 SURGE TANK 650 F 0 PSIG 10' X 20' CS/CS W/LEGS 12,200# SOURCE: ESTIMATED	1	15,500	3 %	200	16,200	16,200	4.00	4.10	4.03	\$65,300
	V3 REFLUX TANK 650 F 75 PSIG 2.5' X 10' CS/CS W/SOLS 1,450# SOURCE: ESTIMATED	1	2,500	3 %		2,600	2,600	4.00	4.10	4.03	\$10,500
450P	P1 LEAN AMINE CS/12CR HOR 192 GPM DISCH 150 PSIG DELTA 150 PSI-35' PUMP TEMP 140 F @ 0.99 SG MOTOR HP 35 @ 7777 RPM SOURCE: ESTIMATED	2	19,300	3 %	200	20,100	40,200	4.00	7.00	4.75	\$191,000
	P2 STILL REFLUX CS/12CR HOR 18 GPM DISCH 40 PSIG DELTA 40 PSI - 95' PUMP TEMP 120 F @ 0.99 SG MOTOR HP 1 @ 7777 RPM SOURCE: ESTIMATED	2	5,000	3 %	100	5,300	10,600	4.00	7.00	4.75	\$50,400

PROJECT: TURBINE FUELS PROJECT  
 SOLAR WATER STRIPPER AND AMMONIA PLANT  
 ESTIMATE NO: T203  
 JOB NO: 8-8768

BY: J. T. HARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	FACTOR HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
200	TOWERS										
300	REACTORS										
300A	AIR COOLED HEAT EXCHANGERS										
300S	SHELL & TUBE HEAT EXCHANGERS										
300H	FIRED HEATERS										
400	TANKS	1					669,100				\$1,592,500
400V	VESSELS										
450C	COMPRESSORS										
450P	PUMPS										
450	SPECIALIZED EQUIPMENT CROLL REYNOLDS STM JET STS UNION CARBIDE H.P.U. WILCOX AC/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T. PROCESS SYSTEM RALPH M. PARSONS CO. SYSTEM	1					6,042,000				\$17,401,000
A	TOTAL MAJOR EQUIPMENT	1					6,711,100		2.83		\$18,993,500
500	INSTRUMENTS	15 % M.E.					1,006,700		4.00		\$4,026,800
620	INSULATION										
	ROUNDING										(\$300)
A-B	TOTAL DIRECT INSTALLED COST										\$23,020,000
	HOME OFFICE COSTS	15 % D.I.C.									\$3,453,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$26,473,000
	CONTINGENCY	25 %									\$6,618,000
	ROYALTIES										
	TOTAL INSTALLED COST	4/0/85									\$33,091,000
	ESCALATION		% /YR. FOR		YEAR						
	TOTAL INSTALLED COST	4/0/85									\$33,091,000

PROJECT: TURBINE FUELS PROJECT  
 SOUR WATER STRIPPER AND AMMONIA PLANT  
 ESTIMATE NO: 7293  
 JOB NO: 848768

BY: J.T. HARLAM  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST
								LOW	HIGH	USED	
480	ANHYDROUS AMMONIA RECOVERY SYSTEM W.W.T. PLANT PRODUCTS AMMONIA 100.4 ST/D HYDROGEN SULFIDE 142.6 ST/D STRIPPED WATER 1,747 GPM @ 140 F	1	5,700,000	5 %	57,000	6,042,000	6,042,000	2.50	4.00	2.88	\$17,401,000
440T	SOUR WATER FEED TANK OPER. TEMP. 100 F DIA. 140' HT. 48' ROOF=COV'D.FLOATER VOLUME 120,000 BBL'S.	1	631,200	5 %	6,300	669,100	669,100	2.00	3.50	2.38	\$1,592,500

PROJECT: TURBINE FUELS PROJECT  
 SULFUR RECOVERY UNIT  
 ESTIMATE NO: 7294  
 JOB NO: 848768

BY: J. T. MARLAN  
 FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN ALLOW.	FREIGHT	UNIT COST	MATERIAL COST	LOW	HIGH	USED	INSTALLED COST
ESTIMATE SUMMARY											
270P	TOWERS										
300R	REACTORS										
320A	AIR COOLED HEAT EXCHANGERS										
320S	SHELL & TUBE HEAT EXCHANGERS										
370H	FIRED HEATERS										
400T	TANKS										
400V	VESSELS										
450C	COMPRESSORS										
450P	PUMPS										
480	SPECIALIZED EQUIPMENT CROLL REYNOLDS STM JET STS UNION CARBIDE H.P.U. W-L/DP/ACP/SRI SYSTEMS ELEC. PRECIP. & FLY ASH COLL. SYS CHEVRON W.W.T. PROCESS SYSTEM RALPH M. PARSONS CO. SYSTEM	1					6,148,000				\$22,132,800
A	TOTAL MAJOR EQUIPMENT						6,148,000		3.60		\$22,132,800
600	INSTRUMENTS		15 % M.E.				922,200		4.00		\$3,688,800
620	INSULATION										
	ROUNDING										\$-00
A+B	TOTAL DIRECT INSTALLED COST										\$25,822,000
	HOME OFFICE COSTS		15 % D.I.C.								\$3,873,000
800	CATALYST & CHEMICALS										
	SUB TOTAL										\$29,695,000
	CONTINGENCY		25 %								\$7,424,000
	ROYALTIES										
	TOTAL INSTALLED COST		4/9/85								\$37,119,000
	ESCALATION		% /YR. FOR		YEAR						
	TOTAL INSTALLED COST		4/9/85								\$37,119,000

SUN REFINING AND MARKETING CO.

P. E. T. S. FACTORED ESTIMATE WORKSHEET

DATE: 16-Mar-87

PROJECT: TURBINE FUELS PROJECT  
SULFUR RECOVERY UNIT

ESTIMATE NO: 7294  
JOB NO: 848768

BY: J.T. HARLAN  
FOR: L. MAGILL

COST CODE	DESCRIPTION	QTY	QUOTE	DESIGN		UNIT COST	MATERIAL COST	FACTOR			INSTALLED COST	
				ALLOW.	FREIGHT			LOW	HIGH	USED		
480	SULFUR RECOVERY UNIT		1	5,800,000	5 %	58,000	6,148,000	6,148,000	2.50	4.00	3.60	\$22,132,800
	S.R.U. PLANT FEED											
	AMINE ACID GAS		78.6	LT/D								
	S.U.S. ACID GAS		68.8	LT/D								
	TOTAL		147.4	LT/D								



END  
DATE  
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4-88  
DTIC