BENCH-SCALE CO-PROCESSING ECONOMIC ASSESSMENT
Final Report

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For
U.S. Department of Energy
Pittsburgh Energy Technology Center
Pittsburgh, Pennsylvania

By
UOP
Des Plaines, Illinois

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**UOP
Des Plaines, Illinois**

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EXECUTIVE SUMMARY

The UOP Co-Processing scheme is a single-stage slurry catalyzed process in which petroleum vacuum resid and coal are simultaneously upgraded to a high-quality synthetic oil. A highly active dispersed catalyst has been developed which enables the operation of the co-processing unit at relatively moderate and high temperatures and relatively high pressure. Under the current contract, a multi-year research program was undertaken to study the technical and economic feasibility of this technology.

All the contractual tasks were completed. Autoclave experiments were carried out to evaluate dispersed vanadium catalysts, molybdenum catalysts, and a less costly UOP-proprietary catalyst preparation technique. Autoclave experiments were also carried out in support of the continuous pilot plant unit operation and to study the effects of the process variables (pressure, temperature, and metal loading on the catalyst).

A total of 24 continuous pilot plant runs were made. Research and development efforts during the pilot plant operations were concentrated on addressing the cost effectiveness of the UOP single-stage slurry catalyzed co-processing concept based on UOP experience gained in the previous DOE contract (DE-AC22-84PC70002). To this end, effect of catalyst metal concentration was studied and a highly-active Mo-based catalyst was developed. This catalyst enabled successful long-term operation (924 hours) of the continuous bench-scale plant at highly severe operating conditions of 3,000 psig, 465°C temperature, and 2:1 resid-to-MAF (moisture- and ash-free) coal ratio with 0.1 wt-% active metal. The metal loading of the catalyst was low enough to consider the catalyst as a disposable slurry catalyst. Also, liquid recycle was incorporated in the pilot plant design to increase the reactor back mixing and to increase the flow of liquid through the reactor (to introduce turbulence in the reactor) and to represent the design of a commercial-scale reactor. This coupled with the highly active catalyst extended the operability of the pilot plant to a much higher temperature of 465°C (compared to 425°C attained under the previous contract with a vanadium catalyst). Under this high-severity conditions, coal conversions greater than 91 weight- (wt)%, heptane insoluble conversions greater than 77 wt-%, and non-distillable (510°C+ material) conversions greater than 83% were achieved.

Alternate catalytic environments were investigated by operating the continuous pilot plant reactor in co-current and counter-current mode with and without packing. Operational problems were encountered during these tests; however, these were believed to be related to the hydrodynamics of the reactor. With further understanding of the hydrodynamics and proper reactor design these problems could be overcome.

The continuous pilot plant data were used to develop conceptual commercial plant designs for the UOP single-stage slurry catalyzed co-processing flow scheme. Basic engineering work was completed to provide the process flows, heat and weight balances, and equipment selection information necessary to generate capital and operating cost estimates for each of the units. The estimated erected cost (EEC) for a co-processing unit handling 25,683 BPSD of Lloydminster vacuum resid and 2,410 MT/d of as-received Illinois No. 6 coal was estimated to be \$183.6 million. This cost includes the coal preparation, catalyst preparation, and solids/resid separation sections.

A refinery economic model was constructed containing the co-processing unit and the downstream upgrading units. This model facilitated the economic analysis and optimization, using linear programming techniques, of the upgrading and blending operations needed to produce salable finished products from intermediate co-processing products. At current (average 1993) crude and finished product prices, the internal rate of return (IRR) on the investment for a refinery including a co-processing unit was estimated to be 15%, compared to 20% for the same refinery without the co-processing unit. At the high 1981 crude and product price levels, the IRR for the co-processing refinery becomes relatively more attractive; however, it is still not high enough for a refiner to justify investment in such an integrated refinery.

1.0 INTRODUCTION

This is the Final Report for the UOP Bench-Scale Co-processing contract which is being funded by the Pittsburgh Energy Technology Center of the U.S. Department of Energy (DOE) under contract number DE-AC22-87PC79818. The overall objectives of this project were to expand on the work carried out by UOP under a previous DOE-supported "Coal Liquefaction Co-processing" contract (DOE Contract Number DE-AC22-84PC70002) by defining process performance at commercially attractive process conditions, by improving catalyst effectiveness, and by updating the economic assessment of the UOP's single-stage slurry-catalyzed co-processing concept. The work performed under the previous contract is summarized in a series of reports¹⁻³.

The present program was divided into four major tasks:

1. Project Management Plan
2. Laboratory Support
3. Continuous Bench-Scale Operations
4. Assessment of New Results

This report is the documentation for the work completed under Task 4, which includes comparison of data obtained under this contract with results from the previous UOP contract³, modification of the conceptual co-processing plant design, cost and utility estimates for the modified co-processing flow scheme, and economic assessment of the modified co-processing design. The work carried out under Task 2 and 3 has been summarized previously in a series of reports⁴⁻⁶.

UOP's co-processing scheme uses a single-stage slurry catalyzed reactor in which petroleum-derived vacuum resid and finely-ground coal are simultaneously upgraded to high-quality synthetic oil. A highly active, well-dispersed UOP proprietary catalyst is

utilized to permit operation at relatively moderate and high temperatures. A simplified diagram of the UOP co-processing flow scheme is shown in Figure 1. Finely-ground coal, petroleum resid, and catalyst are mixed, combined with hydrogen, heated to reaction temperature, and then directed to a single-stage reactor where simultaneous upgrading of the petroleum-derived resid and coal takes place. The reactor products go to a hot separator where the hydrogen, light gases, and light oil are carried overhead. The hot separator overhead is directed to a series of progressively colder separators to recover a hydrogen-rich gas which is then recycled to the reactor. Bottoms from the hot and the series of cold separators are combined and sent to the atmospheric distillation column where they are fractionated to recover the light gases and light oil. Bottoms material from the atmospheric distillation column is sent to a vacuum fractionator from which the vacuum gas oil is recovered as the overhead. The vacuum bottoms material is sent to a solid/resid separation unit to recover additional oil product. The solid material consisting of unconverted coal, coal minerals (ash-forming materials), and used catalyst is disposed off. As will be discussed in greater detail in the following sections, no catalyst recovery is anticipated in the modified flow scheme proposed in this document. This is a major change from the conceptual flow scheme developed under the previous co-processing contract³ and this change was made possible by the highly-active catalyst (at relatively low metals loading) developed under the current program.

It is UOP's opinion that the easiest way to develop and eventually commercialize co-processing will be by integration of the co-processing scheme with a petroleum refinery. Stand-alone co-processing will require very high capital investments and may not be attractive. The objective of UOP experimental work is not to produce the finished products, but rather to maximize the amount of material that can be most efficiently upgraded by conventional refinery operation to high-quality transportation fuels.

2.0 CONCEPTUAL COMMERCIAL CO-PROCESSING UNIT

In this section, the design of the UOP conceptual co-processing plant will be reviewed. Except for a few modifications, the plant flow scheme is similar to that described in the Final Report for the previous co-processing contract³. The modifications are discussed in detail in this section.

2.1 Design Basis

For the conceptual design evaluated in the last contract, Lloydminster Resid and Illinois No. 6 coal were used as the feedstocks. These materials are also used as references in UOP pilot plant studies under the current contract. For the comparison to be meaningful between the two assessments made under the two DOE-UOP co-processing contracts, same feedstocks are used in the design of the conceptual plant discussed in this report.

Since the objective of UOP co-processing is to produce a slate of products that can be more efficiently upgraded in a petroleum refinery than the raw coal and the resid feedstocks, it is important that the design basis for co-processing closely match the size of conventional refineries. The most probable route for co-processing to become viable and acceptable to the refiners would be to use the existing refinery infrastructure and product distribution networks. The size of such an integrated co-processing unit will be dictated by the amount of resid available from the refinery to provide the "solvent" for co-processing the coal. Process variable studies conducted earlier under the UOP-DOE co-processing contracts indicated that a 2:1 resid-to-MAF coal ratio was optimum for the reference feedstocks¹. For an average sized 100,000 barrels per stream day (BPSD) refinery with a conventional distribution of products, the resid portion of the crude to the refinery might typically be 20-30 vol.-%. At a 2:1 resid-to-MAF coal ratio, such a refinery

could handle a maximum of about 2,500 metric tons/day (MT/D) of coal. When compared to a stand-alone coal liquefaction plant, the co-processing unit will be quite smaller, as much as by a factor of ten.

The site for the conceptual plant will be in the midwestern United States to minimize the transportation costs for the coal and the feedstocks to the site. The Lloydminster crude will be delivered to the site by pipeline from Canada. Lloydminster is a fairly heavy crude (API = 20.7) and would probably not be used as the sole crude feedstock in a conventional refinery. To get a more realistic product distribution, a second lighter crude, Interprovincial Mixed Blend Sweet (IPL), is included as feedstock in the design of the refinery complex (with and without integration with the co-processing unit). This light feedstock was chosen, in part, for its access to the same Interprovincial pipeline that would be used to supply the heavy Lloydminster crude. Lloydminster is typically blended with light condensate to permit easier transport by pipeline. Assay for these two crudes are given in Tables 1 and 2. The refinery for this study will be sized to process 50,000 BPSD of Lloydminster crude plus 50,000 BPSD of IPL crude. The co-processing unit will handle the Lloydminster resid (25,683 BPSD) plus enough Illinois No. 6 coal (2,311 MT/D moisture-free (MF) basis) to give the desired resid-to-MAF coal ratio.

The conceptual co-processing design described in the following sections is based on the pilot plant results that were generated in Run No. 37, the "High-Severity" run. Results of this run have been discussed in detail in various technical reports that were prepared under this contract^{5,7}. The high-severity run was carried out with the reference feedstocks, Lloydminster vacuum resid and Illinois No. 6 coal, at a 2:1 resid-to-MAF coal ratio. The operating conditions for this run were: 564 hours-on-stream (HOS) at 455°C temperature and 348 HOS at 465°C temperature, 3000 psig pressure, and with a minimum liquid recycle to fresh feed ratio of 7.5. The catalyst used for this run contained 0.1 wt-%

molybdenum (Mo) based on the coal + resid feed. Daily weight balances were conducted to ensure integrity of the operations and to obtain reliable hydrogen consumption data. Average performance observed at this two process conditions is shown in Table 3.

Comparison of the data in Table 3 shows that at the higher temperature of 465°C conversion of coal, C₇ insolubles, and 510°C+ non-distillable is significantly higher than at 455°C. Also the feed is more thoroughly cleaned as shown by the denitrification and desulfurization at the higher temperature. As would be expected, the yield of the less desirable lighter materials (C₁-C₄) is higher (11.26 wt-% versus 7.49 wt-%) at 465°C; however, this is also accompanied by a higher (more desirable) liquids yield (66.65 wt-% versus 63.72 wt-%). As a result, it was decided to base the economic assessment on the results obtained at the higher temperature of 465°C. The products from three 12-hour periods (No. 51, 62, and 72) at 465°C temperature were combined and detailed physical and chemical analyses were performed on the composite sample. Results of this detailed analysis are presented in Appendix A. This product analysis was used as the basis to estimate the yields and properties of various fractions that might be expected from the conceptual co-processing plant. A summary of the intermediate co-processing products properties is shown in Table 4.

The basis for co-processing economic assessment under the previous contract was Run No. 19³. The operating conditions for Run No. 19 were: 425°C temperature and 3,000 psig pressure with 1.0 wt-% vanadium (V) catalyst. Pilot plant results for Run No. 19 (low-severity run) and Run No. 37 (high-severity run) are compared in Table 5. There are several differences between the two runs and these have considerable impact on the design of the conceptual plants.

- ◆ Liquid Recycle: There was no liquid recycle in Run No. 19, whereas in Run No. 37 liquid from the high-pressure separator (HPS) was recycled to the reactor. The HPS liquid to fresh feed ratio was at least 7.5 and for the three periods discussed

above, this ratio was 8.4. The objective of the liquid recycle was to improve the operation of the pilot plant by pumping larger quantities of liquid through the reactor. This introduced turbulence in the reactor and helped in mixing of the reactor contents and made it possible to operate the reactor at higher temperatures. The pilot plant operation with liquid recycle also represents a more realistic comparison with the commercial reactor operation.

In commercial design of the co-processing plant, the reactor can be designed to assure that the flow through the reactor is in the turbulent regime and that there is proper mixing of the reactor contents. This can be done by adjusting the height-to-diameter ratio of the reactor and/or by providing baffles in the reactor. Also, in the commercial plant the liquid throughput is much higher (75,000 g/sec) compared to the pilot plant (0.15 g/sec fresh feed). As a result, external recycle of the liquids to achieve good mixing may not be necessary in the commercial design.

- ◆ Temperature: Run No. 37 was operated at the high severity conditions of 465°C temperature as compared to Run No. 19, which was operated at the low severity conditions of 425°C. This temperature difference has a significant impact on the metallurgy of the reactor for the new co-processing plant. Whereas 2¼ wt-% chromium (Cr) - 1 wt-% molybdenum (Mo) stainless steel was used in the previous design, the new design required higher levels of Cr (3 wt-% Cr - 1 wt-% Mo) in the stainless steel. This coupled with the high pressure operation also restricted the diameter of the reactor that can be used. In the old design, reactor with a 16' diam. was designed, in the new design reactor diameter was limited to 8'. As a result of these changes, the number of reactors required increased from 2 to 6.
- ◆ Catalyst - In the current design, UOP's proprietary low-metal Mo (0.1 wt-% metal) catalyst is being used. In Run 19, 1.0 wt-% V catalyst was used. The Mo catalyst (together with the liquid recycle) enabled the operation of the plant at the high

severity conditions. Under the previous contract an exhaustive analysis was conducted to determine the effect of metal loading on the need for catalyst recovery and recycle. Cost of catalyst required for a once-through operation was balanced with the cost of recovery, recycle, and makeup catalyst. It was determined that with the relatively lower-activity V catalyst, the metal loading had to be reduced to below 0.08 wt-% to eliminate catalyst recovery and recycle. With this analysis as the background it was determined that with the higher-activity 0.1 wt-% Mo catalyst, recovery and recycle of the catalyst would not significantly improve the process economics. So for the current design, the Mo catalyst was used as a throw-away catalyst.

- ◆ Product Yields - The products resulting from low-severity and high-severity runs are quite different (see tables 6, 7, 8, and 9). Higher proportion of the feed is converted to lighter materials in the high-severity run than in the low-severity run (Figure 2). This should be expected since the high-severity run operated at a much higher temperature of 465°C than the low-severity run (425°C).

2.2 PROCESS FLOW

For simplicity, the co-processing unit design has been divided into several subsections as shown in Figure 3. Overall material balance for base-case co-processing refinery design is presented in Table 10 and illustrated in Figure 3. Also, in the description that follows, some major flow streams for this base case are reported in the parenthesis.

In the first subsection, as received (AR) coal (212,219 lbs/hr - on a moisture-free basis) is unloaded, transported from storage, dried and ground to a particle size of less than 200 mesh (74 microns), and slurried with the vacuum resid (384,217 lbs/hr) and catalyst slurry. This section will be similar to any other coal liquefaction process, and a detailed design for this was not carried out. Capital and utility costs for the coal

preparation section were scaled from a study done by Bechtel for the Electric Power Research Institute (EPRI) for the SRC process⁸. Major process requirement and equipment in this section include:

- ◆ Live storage capacity for 5 days of coal feed
- ◆ Long-term storage for a 30-day inventory of coal
- ◆ Reclaiming and transportation equipment
- ◆ Pulverizing equipment: Feed coal is ground to approximately 80% <200 mesh, and dried to about 2% moisture in roller mills.
- ◆ Cyclone separators and baghouse
- ◆ Weight feeder for coal
- ◆ Eductor to combine the ground coal with resid
- ◆ Slurry mixing tank, including a recirculation pump to maintain even composition. Two trains of pump plus one spare are included.

In the second subsection, the catalyst is converted to its active form by a proprietary process and then slurried with vacuum gas oil (VGO) for transport (Total catalyst-VGO slurry is 2,368 lbs/hr). The catalyst slurry, prepared coal, and vacuum resid are then mixed in the slurry preparation section.

For the other three subsections shown in Figure 3, where applicable the design was kept similar to that for the conceptual co-processing plant of UOP's previous contract³. Where changes were necessary because of the different operating conditions of this design, detailed designs were completed. The engineering specifications for the major equipment present in this three sections are presented in Appendix B. In the following sections, the reactor section, fractionation section, and the solids/resid separation section are described.

2.2.1 Co-Processing Reactor Section

The flow scheme for the co-processing reactor section is shown in Figure 4. The coal/resid/catalyst slurry (598,804 lbs/hr) is combined with a hydrogen recycle stream (337,639 lbs/hr, 69.2 mole-% H_2 , 3.0 mole-% H_2S , 16.3 mole-% methane, 4.2 mole-% ethane, 1.8 mole-% propane, 4.0 mole-% CO_2 , 1.1 mole-% C_4+ , 0.3 mole-% NH_3 , and 0.1 mole-% H_2O) and fed up flow through the reactor. The hydrogen recycle stream comprises of makeup hydrogen (16,804 lbs/hr, 99.9 mole-% H_2) and the cold separator overhead stream (320,835 lbs/hr, 59.6 mole-% H_2 , 3.9 mole-% H_2S , 21.3 mole-% methane, 5.6 mole-% ethane, 2.4 mole-% propane, 5.3 mole-% CO_2 , 1.5 mole-% C_4+ , 0.3 mole-% NH_3 , and 0.1 mole-% H_2O). Heat for the reactor is supplied by preheating 328,799 lbs/hr of the recycle hydrogen stream in the Recycle Gas Heater and heating the rest of the hydrogen recycle stream (8,839 lbs/hr) and the feed slurry by exchange with the hot separator bottoms. The hydrogen recycle rate is sufficient to maintain thorough mixing and intimate contact of the hydrogen with the coal slurry in the reactor.

As discussed above, since the operating temperature of the current design (465°C) is significantly higher than the previous design (425°C), the reactor metallurgy had to be changed. As a result, there was a limitation on the diameter of the reactor that could be manufactured for this high pressure operation. So in the current design the number of reactors required were changed from 2 (in the previous design) to 6. The reactor size is 8' diam. (internal), 45' height, and 15.6" thick. These six reactors will operate in parallel to treat all the feed to the co-processing section.

Another change between the two designs has resulted from the high level of light gases produced in the high severity case (Figure 2). Additional equipment will be required to absorb the C_4 's and C_5 's from the fuel gas. These light gases will be absorbed in an absorber by naphtha and then the naphtha will be recovered from the light gases in a stripper. In the integrated co-processing/petroleum refinery this additional absorber and

stripper will be integrated with the refinery saturated gas recovery system. So for the conceptual design the cost of these additional equipment is included in the co-processing section; however, the absorber and stripper are not shown in the co-processing reactor flow scheme.

The reactor effluent (936,442 lbs/hr) is sent to a series of separators, each at a lower temperature than the last, to recover a high-purity hydrogen stream for recycle. The commercial design differs from the pilot plant design in this aspect since it has multiple separators and flash drums instead of a single vessel (as in the pilot plant). The multiple separators are necessary at the much higher commercial flow rates to ensure maximum recovery of the hydrogen. Every effort has been made in the design to reduce solution losses of hydrogen and thereby minimize the makeup requirements, since these are some of the largest costs of operation. The heat from the hot separator overhead is recovered by exchange with the recycle gas, and heat from the intermediate separator overhead stream is recovered by generation of steam. The combined hot (123,235 lbs/hr) and intermediate separator bottoms (212,943 lbs/hr) are sent to a hot flash drum, where the C₃- material goes overhead to the cold flash drum in combination with the cold separator bottoms. The cold flash overhead stream contains primarily hydrogen, methane, ethane, and hydrogen sulfide (26,579 lbs/hr, 34.7 mole-% H₂, 9.5 mole-% H₂S, 27.2 mole-% methane, 11.0 mole-% ethane, 5.0 mole-% propane, 9.0 mole-% CO₂, 0.8 mole-% NH₃, 0.2 mole-% H₂O and rest C₄+), and is sent to the refinery amine treating and sulfur recovery unit and then to the fuel gas pool. The hot (332,111 lbs/hr) and cold (204,622 lbs/hr) flash bottoms streams pass on to the co-processing fractionation section for further separation. Both the hot separator and the hot flash drum have two VGO washing trays to minimize carry-over of solids to the downstream equipment.

2.2.2 Co-Processing Fractionation Section

The flow scheme for the co-processing fractionation section is shown in Figure 5. The hot and cold flash bottoms product streams from the reactor section are stripped with medium pressure (150 psig, 8,000 lbs/hr) steam to yield an overhead sour fuel gas stream (42,656 lbs/hr, 5.5 mole-% H_2 , 21.9 mole-% H_2S , 12.5 mole-% methane, 17.2 mole-% ethane, 18.5 mole-% propane, 10.6 mole-% C_4s , 3.7 mole-% C_5+ , 8.4 mole-% CO_2 , 1.2 mole-% NH_3 , and 0.5 mole-% H_2O) and a light naphtha (C_5/C_6 - 13,145 lbs/hr, 0.9 mole-% methane, 5.7 mole-% ethane, 18.4 mole-% propane, 27.1 mole-% butanes, 24.6 mole-% pentanes, 9.0 mole-% hexane, 3.5 mole-% C_7+ , 1.3 mole-% CO_2 , 0.8 mole-% NH_3 , 8.7 mole-% H_2S) product. The stripper bottoms (481,879 lbs/hr) is heated in the charge heater and sent to an atmospheric fractionation column. The light gas oil is removed as a side cut from the fractionator. Part of this stream is used to generate steam and is returned to the fractionator as an external reflux and the remainder is stripped with medium-pressure steam in a side-cut stripper. The resulting light gas oil product is cooled and sent to the refinery.

The bottoms (230,653 lbs/hr) from the atmospheric fractionator is heated in a furnace and fed to the vacuum column. Separate light vacuum gas oil (LVGO - 51,825 lbs/hr), heavy vacuum gas oil (HVGO - 65,989 lbs/hr), and slop oil (4,000 lbs/hr) side-cuts are taken from the vacuum column to maximize the heat recovery. These streams would probably be recombined further for upgrading in the refinery. The vacuum column bottoms (108,743 lbs/hr), containing 1050+°F resid and all of the solids is sent to the solids/resid separation section for recovery of additional product and concentration of the solids.

2.2.3 Solids/Resid Separation Section

A simplified schematic of the conceptual solids/resid separation section is shown in Figure 6. In this section, the unconverted coal, ash, and catalyst solids are transferred from the resid phase to an aqueous phase. In a series of mixers, the solids-containing vacuum bottoms is diluted with 1.5 volumes of naphtha and then water washed using 3.5 lbs of water for every pound of the solids. Temperature and pressure of these steps are maintained to minimize the amount of naphtha solvent required to effect the separation of the solids from the resid phase to the aqueous phase and also to maintain the operation in the liquid phase. The resulting mixture of aqueous and the hydrocarbon phase is allowed to separate in a settler.

The hydrocarbon phase (top layer) from the settler is heated in flash drums and the naphtha is steam-stripped and recirculated as a recycle to the naphtha surge drum. A very small amount of makeup naphtha is required to offset the naphtha losses with the stripper bottoms. The makeup naphtha is taken as a slip-stream from the co-processed naphtha product. The stripper bottoms approximates as a vacuum resid stream. This stream is estimated to contain about one percent of the total feed solids to the separation, so there is a small loss of catalyst associated with this unit.

The aqueous slurry of catalyst/coal solids from the bottom of the settler is cooled for the separation of the solids using rotary vacuum filters at atmospheric pressure. Water is recovered overhead from the aqueous slurry flash drum, as the pressure is reduced to less than atmospheric. Water from the flash drums and water recovered from the vacuum filters is recycled to the water surge drum. To prepare the filter cake for solids disposal, the cake is then mechanically expressed to remove additional water from the solids. The expressed solids contain about 30% water which must be replenished by makeup water to the water surge drum.

2.3 HEAT AND WEIGHT BALANCES

Detailed heat and weight balances were prepared for the reactor, fractionation, and solid/resid sections of the co-processing unit using Hyprotech's "HYSIM" process simulation program. A simplified mass balance for these sections for the high-severity case is given in Table 10 and illustrated in Figure 3. In this summary, the makeup naphtha for the solid/resid separation has been netted out of the product. The total MAF feed to the unit is 572,243 lb/hr (384,217 lb/hr resid and 188,026 lb/hr MAF coal, Resid-to-MAF Coal ratio of 2.0) or 260 MT/hr. The makeup hydrogen to the unit is 2.94 wt-% of the MAF feed, compared to 1.80 wt-% for the pilot plant operation. This difference reflects the commercial reality of solution losses during fractionation and recovery of the co-processing products. Total liquid product from the unit is 35,160 BPSD, from which a part of the VGO will be used for catalyst slurry and VGO-washing of the separators to minimize the solids carry-over. Comparing the gross liquid product to the 25,683 BPSD of resid charged to the unit, we can see that the contribution of the coal is 9,477 BPSD (gross) of liquid products, or 4.63 barrels/MAF MT of coal.

2.4 COST AND UTILITY ESTIMATES

Detailed equipment specifications and costs for the co-processing section are summarized in Appendix B and C. All costs are reported in 1st quarter 1994 dollars. In arriving at these numbers varying levels of detail were used for different subsections, with the greatest attention given to the reactor and the fractionation sections. For the coal preparation section, which includes no change in equipment from the previous conceptual design of the low-severity case³, the equipment costs were estimated by applying a cost factor of 1.2 to the equipment costs of the low-severity design. This factor represents the increase in equipment cost from 1988 to 1994. Similar procedure was used to estimate the equipment costs for the solid/resid separation subsection. The FOB costs estimated for these sections are probably good to $\pm 30\%$.

For the reactor/fractionation section, individual material costs were determined by using both the in-house methods and vendor budget quotes. The overall accuracy of these direct material costs are estimated to be $\pm 15\%$. However, two important points should be noted concerning these equipment costs:

- ◆ The average per piece cost of equipment is quite high.
- ◆ A number of pieces of equipment (representing a very significant part of the total cost for this section) are unusual due to metallurgy, design conditions, size or solids handling requirements. As a result, few vendors have ever quoted, let alone manufactured, this type of equipment. Their ability to quote accurately is thus diminished in some situations, and accordingly their numbers may be high or low.

The estimated erected cost (EEC) of the co-processing reactor and fractionation sections, based upon overall factoring, is \$100,000,000. This cost includes associated bulk items, installation, contractors' home office expenses, and profit. The EEC is for battery limits only and is accurate within $\pm 50\%$.

Table 11 summarizes the EEC for the major subsections of the co-processing unit. The reactor and the fractionation section account for about 55% of the total EEC. About 43% of the total capital costs are associated with coal preparation and separation of unconverted coal, ash-forming minerals, and spent catalyst from the liquid product. These capital expenses would be present in any coal liquefaction facility and are not exclusively related to co-processing. Estimate for the coal preparation, handling, transportation, grinding, and drying is based on a previous EPRI report and cost factors to account for the cost increase to 1st quarter 1994 costs. The accuracy of this estimate is not as high as the cost estimate associated with the co-processing and the refining sections.

The major equipment cost items (those in excess of \$1,000,000) are listed in Table 12. The reactors account for 33%, the makeup compressors account for 17% and the intermediate separator overhead air coolers account for 9% of the total cost for the reactor/fractionation section. Together these three major pieces account for 58% of the total cost for the reactor/fractionation section. The reactor costs were estimated from an actual quote from a vendor. Originally the co-processing section was designed as a four reactor system; however, in discussion with Nuovo Pignone of Italy (a manufacturer specializing in heavy thickness pressure vessel designs), it was determined that the reactors will be too thick to manufacture. As a result, the reactor section was designed as a six reactor system. The reactor dimensions are: 2,439 mm (8') internal diameter and 13,760 mm (45.1') height with shell thickness of 396 mm (15.6") and head thickness of 216 mm (8.5"). The reactors will be made out of 3 Cr - 1 Mo stainless steel with two layers of TP 347 weld overlay and will weigh a total of 2,880,000 kg (3,168 tons).

The relatively high costs for these items are due to the high design pressure of the reactor section, and the conservative metallurgy specified to withstand high potential HCl and H₂S levels. Most of the equipment in contact with the solids has been specified with a 1/4" TP 347 weld overlay cladding, and the vessels and exchangers in contact with the overhead gas streams are specified as Incoloy 825. If the operating pressure of the co-processing section could be decreased, significant reduction in the capital costs for this section would be realized.

The major equipment costs in the solids/resid separation section are due to the rotary vacuum filters and belt expressors, which together account for about 48% of the total costs for this section. These items are used to dewater the solids which are subsequently disposed off. Minimization of volume by removing as much water as possible is critical for reducing the overall operating costs for solids handling.

Utility requirements for the co-processing unit are summarized by section in Table 13 and major utility consumption items are broken out in Table 14. Power costs account for 87% of the net utility costs. About 31% of the power costs are associated with coal preparation, the bulk of which goes to pulverization and drying of the coal. Most of the remaining power costs (63%) are associated with the reactor and the fractionation section. In this section, the single largest power consumer is the makeup hydrogen compressor in the reactor section, which uses 16,000 HP to compress the hydrogen from 300 psig hydrogen plant pressure to the 3,000 psig reactor pressure. This single item accounts for 75% of the power used in the reactor/fractionation section and 48% of the total co-processing plant power requirements. Any reductions that could be made in the reactor pressure, will significantly reduce the power cost associated with the hydrogen compressors.

After power costs, most of the remaining utility costs are due to fuel consumption in the recycle hydrogen, fractionator charge, and vacuum column charge heaters. However, costs associated with these are offset by heat recovery as low-pressure steam in the steam generators. Other utility costs are not significant.

2.5 COMPARISON OF HIGH- AND LOW-SEVERITY CO-PROCESSING UNIT COSTS

Comparison of the conceptual co-processing unit costs for the low-severity design and the high-severity design is shown in Table 15 and the major cost item for these two designs are listed in Table 16. The 1988 UOP co-processing cost assessment is the basis for the cost numbers for the low-severity design case.³ These numbers were multiplied by a factor of 1.2 to account for the increase in the equipment cost between 1988 and 1st quarter 1994. The corrected numbers are shown in Table 15 and 16.

There is no difference in the design of the coal preparation and the solids/resid separation section between the two designs and as a result, costs for these two sections are identical. Catalyst preparation plant cost for the high-severity design is a much lower \$3.2 million versus \$17.5 million for the low-severity design. The high-severity design uses 0.1 wt-% Mo catalyst and the low-severity design uses 1.0 wt-% V catalyst. The difference in the capital cost of these two UOP-proprietary catalyst preparation plants is because of the different metal loadings and subsequently different volumes of the equipment required. Table 17 shows a comparison of the operating costs for the catalysts in the two design cases in a throw-away operation. For the high-severity case, catalyst operating costs are reasonable at \$27 million/year to justify a throw-away operation; however, for the low-severity case catalyst operating costs are prohibitively high at \$711 million/year. Additional capital investment for the recovery and recycle of the vanadium catalyst is required for the low-severity design case.

The reactor and fractionation section of the low-severity design case is 20% more expensive than the high-severity design case (\$120 million versus \$100 million). In the low-severity operation, significantly higher quantities of the heavier products are formed (385,416 lb/hr of 650°F+ liquids and 66,667 lb/hr of solids) compared to the high-severity case (189,249 lb/hr of 650°F+ liquids and 41,396 lb/hr of solids). As a result, the volume of equipment required downstream of the reactor to process these heavier materials is significantly greater for the low-severity case than the high-severity case. This difference is reflected in the costs of the two conceptual design cases. For example, in the low severity case very expensive feed/hot separator bottoms heat exchanger were required because the flow rate of the heavier material through these exchangers was much higher than in the high-severity case. The difference in the cost of this one item in the two cases was approximately \$18 million. The differences in the costs of reactor downstream equipment for the high-severity versus the low-severity design case were more than enough to overcome the more expensive reactors required for the high-severity design case (\$17.4 million versus \$9.8 million for low-severity reactors).

3.0 CO-PROCESSING ECONOMICS

As discussed earlier, UOP believes that for co-processing to be commercially viable, integration of co-processing with petroleum refinery is essential. Coal-derived liquid products differ from the corresponding crude cuts in many respect. The intrinsic properties of coal-derived liquid (its nitrogen, sulfur, and aromatic content) require a different level of upgrading than the corresponding boiling range cuts derived from a petroleum crude. In general, the refiner must pay additional costs to upgrade the coal-derived liquid to the same finished product quality to which the crude-derived products are refined. In most cases, the additional upgrading costs include additional hydrogen costs. The cost of additional upgrading steps or refinery modifications must be reflected in the value assigned to the coal or the co-processed products, if a fair assessment is to be made of the potential of the co-processing when compared to petroleum crude-based refineries. Even if the co-processed products are suitable for blending to finished product specs rather than directly upgraded, their value will be low unless the refinery has other streams (of premium quality) necessary to blend with the co-processed products to give a salable product. Many of these uncertainties in economic evaluation of a synfuel unit are avoided if the synfuels unit is evaluated as an integral part of the refinery complex. Hydrogen and utility consumptions can then be calculated over the whole upgrading process. Only the raw materials and the finished products need to be assigned values, and these prices are well established in current markets.

To carry out the economic evaluation of co-processing with the high severity case, the linear programming (LP) model of the refinery, which was developed for the previous economic assessment,³ was used. This model allows for the upgrading and blending of the co-processed products within the refinery to be analyzed and optimized. The basis

for this model is kept similar to the previous economic assessment so that a valid comparison can be made. This basis is re-visited in the following section.³ A description of the model is given in Sections 3.2 and 3.3. A total of four LP runs were made under this contract. A detailed description of these four cases is presented in Section 3.4.

3.1 MODEL BASIS

A number of inputs are required to run the LP model. The feedstock properties and costs must be completely specified. For all the cases discussed in this report, the feed will be 50,000 BPSD of Lloydminster crude, 50,000 BPSD of the lighter Interprovincial crude, and 2,215 MT/day of Illinois No. 6 coal. The criteria for the selection of the feedstocks were discussed in Section 2.1 and crude assays for the Lloydminster and the Interprovincial crude are given in Tables 1 and 2.

The finished product properties and prices must also be specified to run the LP model. The product specifications assumed for this study are listed in Table 18. These specifications are similar to those used for the previous economic assessment.³

It is important that the prices used for crude and finished products all be on a consistent basis. UOP Marketing Services maintains an updated list of current prices which is used for all of the internal economic UOP evaluations. This price structure for average 1993 cost of various feedstocks and products is shown in Table 19. The crude prices were obtained by contacting Petro-Canada's representative. The price of Illinois No. 6 coal was estimated from the average cost of Illinois No. 6 coal delivered to the utilities. The coal price (\$34/ton) represents the contract price paid by utility. This price is higher than the price of coal on the open market. To see the effect of change in coal price a sensitivity analysis was performed, the results of which are discussed under

Section 3.6. Other product and raw materials cost were obtained from Platt's Oilgram and these are average 1993 spot prices for U.S. Gulf Coast. One year averages have been used to eliminate seasonal fluctuations.

It was anticipated that the economics of co-processing would not be favorable at today's relatively low crude prices, so an alternative high crude price case was also examined. Rather than speculate where crude and product prices will be in the future, we used actual historical data from 1981, when crude prices were at about their 15-year peak. Similar approach was also used in UOP's previous co-processing economic assessment.³ This approach of setting high crude prices assures us that the finished products and utility costs are internally consistent with the feedstock prices. The 1981 high crude prices are compared with the 1993 prices in Table 19.

In addition to the given feedstocks, the refinery LP model was allowed to "buy in" butane for use in alkylation or gasoline blending. The butane prices were also obtained from UOP's Marketing Services database and the source for this price was Platt's Oilgram.

The refinery complex was assumed to be hydrogen and fuel self sufficient. All the fuel consumed in the complex came from available refinery streams. An internal hydrogen plant was specified to make up any difference between the hydrogen produced in the reforming unit and the hydrogen demands of the co-processing units and hydrotreaters. All utilities (steam, cooling water, etc.) were produced within the refinery with the exception of power, which was purchased at the prices specified in Table 19.

For financing of the refinery complex which is assumed to be a "grassroots" facility designed to process the crude and coal feedstocks, 100% equity financing was assumed. Offsites are estimated for the support of the process facilities; however, no allowance is made for infrastructure outside of the battery limits of the complex. Other economic assumptions made are listed in Table 20. Ten year straight-line depreciation is assumed

with an estimated ten year plant life. Five years are allowed for construction. Profit after taxes was used as the variable for optimization in the LP model. Simple return-on-investment and payout period were calculated to compare different cases.

3.2 MODEL DESCRIPTION

In addition to the various unit operations described under the co-processing plant flow scheme, the refinery model developed for this study makes provision for the following common refinery units:

Crude Unit(s)	Alkylation Unit
Vacuum Tower(s)	Polymerization Unit
Delayed Coker	Isomerization Unit
Kerosine Hydrotreater	Amine Absorption Unit
Diesel Hydrotreater	Sulfur Plant
Gas Oil Hydrotreater	Hydrogen Plant
Naphtha Hydrotreater	Offsites
Fluid Catalytic Cracker	Utility Conversion Unit
CCR Platforming™ Unit	

The block flow diagram in Figure 7 illustrates how the co-processing unit would fit into the overall refinery. This diagram shows the complex nature of the possible flows between units. Each unit is defined by a series of equations, and the model provides for multiple cases of many of the units, each representing a different feedstock or operating condition. For example, the Platforming™ unit can be run at a number of different severity to make 92, 94, or 98 RON product. For each case, a yield estimate has been made related to the product quantities and properties to the feed quantities and properties of that unit. The on-stream efficiency, base capacity investment costs, and operating and utility costs at base capacity have also been estimated for each unit. For variations from base capacity, the model scales investment costs using a scaling factor selected for the given unit, which is typically in the range of 0.6 to 0.8. Operating and utility costs are scaled directly with unit capacity.

The LP model optimizes the selection of units, flows between units, unit capacities and product blending to maximize profit for the overall refinery complex within given price and product specification constraints. The "optimized" solution often requires iterative modification of the yield or cost estimates if the feed composition or capacities of selected units are significantly different from those specified in setting up the model.

3.3 REFINERY FLOW SCHEMES

Figure 8 shows a simplified version of a conventional refinery flow scheme. The Lloydminster and Interprovincial crude are fractionated into naphtha (IBP-380°F), kerosine (380-480°F), diesel fuel (480-650°F), vacuum gas oil (650-1050°F), and resid cuts in the crude and vacuum towers. The resid is upgraded to lighter products in the delayed coker. The coker gas oil and vacuum gas oil are then hydrotreated and cracked to gasoline and LCO in a Fluid Catalytic Cracker (FCC) unit. The FCC also produces significant quantities of C_3/C_4 olefins that are treated and either alkylated with iso-butane or polymerized to make high octane gasoline blending components. The FCC gasoline is treated before blending, while the FCC light cycle oil is used in fuel oil blending. The FCC slurry oil is generally suitable only for blending in heavy fuel oil or for use as an internal refinery fuel. The naphtha from the coker and crude units is hydrotreated and fractionated to yield light and heavy naphtha cuts. The C_5/C_6 material is isomerized in a PenexTM unit. The heavy naphtha is reformed in a CCR Platforming unit to improve the octane. Light petroleum gases together with saturated gases from most of the refinery's process units are sent to a gas recovery section, where hydrogen sulfide is removed via amine absorption. Sulfur product is then recovered from the hydrogen sulfide in a Claus unit. C_2 and lighter gases are separated from propane and butane by fractionation and utilized as fuel gas. Propane and butane are treated for mercaptan removal and recovered as LPG, with some of the butane being used for gasoline blending and alkylation.

Figure 9 shows how the co-processing unit will be integrated into the overall refinery complex. In this case, the Lloydminster vacuum resid from the fractionation section is sent to the co-processing unit instead of the coker. Various co-processing products are then upgraded in the refinery in substantially the same units that are present in the conventional refinery. One should note that the sizes and utility requirements for these units will differ in many cases from those in the base case conventional refinery. These differences in size and utility requirements are further discussed in the following sections.

Even though the resid product from the co-processing unit has reduced metal levels (54 wt-ppm vanadium (V) and 60 wt-ppm nickel (Ni)) when compared to the resid material (183 wt-ppm V and 82 wt-ppm Ni) fed to the co-processing unit, it has significantly higher levels of asphaltenes (96 wt-% heptane insolubles compared to 14 wt-% heptane insolubles for Lloydminster resid) and Conradson carbon (54 wt-% compared to 17 wt-%) and as such, it is a low-quality material. Also, in comparison to the low severity co-processing resid product, the high severity resid product is of lower quality. This is expected because the high-temperature high-severity operation will more aggressively treat the feed material to relatively lighter product slate, leaving behind a much "dirtier" co-processed resid product. Processing options for the high-severity co-processed resid are limited to use as a refinery fuel or as a coker feedstock. The markets for coke and fuel oil No. 6 would ultimately dictate which option is chosen by the refiner. In this study a delayed coker was selected for the conceptual design because of the added flexibility it provides to the refiner.

Whereas the VGO cut, accounting for 42 vol-% of the total liquid products, was the single largest product from the low-severity co-processing unit, in the high-severity co-processing unit naphtha (25 vol-%), diesel (34 vol-%), and VGO (25 vol-%) were all produced in significant quantities. The co-processed naphtha (C₇-380°F) cut is blended

with the straight run naphtha cuts and coker naphtha. The blended stream is then hydrotreated to produce an acceptable platforming charge stock containing less than 0.5 ppm nitrogen and sulfur.

The middle distillate material (380-650°F diesel) from the co-processing unit along with IPL and Lloydminster diesel and coker distillate are hydrotreated to produce a blending stock for the diesel/fuel oil No. 2 pool. The co-processed VGO (650-1050°F cut) is combined with the straight-run and coker gas oils and then hydrotreated in VGO hydrotreater prior to processing in the FCC unit. The co-processed VGO has a sulfur level of about 1.6 wt-% and a nitrogen level of 6,700 ppm. This compares to about 3.9 wt-% S and 3,500 ppm N for the coker gas oil produced by delayed coking of the Lloydminster resid in a conventional refinery case. The additional nitrogen in co-processing case may be directly attributed to the coal portion of the feed.

3.4 ECONOMICS OF CO-PROCESSING REFINERY

Table 21 summarizes the overall economics of the four cases optimized with the LP refinery model described above. These cases include two refinery configurations, one with the co-processing unit and one without it. Each configuration is examined under two crude/product price structures. The "base case" uses current (1993 average) crude and product prices and "the high crude price case" uses average prices from 1981. Detailed results from each of these four runs are given in Appendix D.

Compared to the conventional refinery case at base price levels, the co-processing case shows a lower internal rate of return (IRR) on capital investment (15% versus 20%, Table 21) and higher payout time (4.3 years versus 3.4 years). This is not surprising considering the low current price of crude and the huge capital requirements that would be needed to add the solids handling facilities for the coal. The total capital required for the base case co-processing refinery (\$895.6 million) is \$214.2 million more than that required for the base case refinery (\$681.4 million). Product/raw material margins are

actually better (\$975,800/day for co-processing versus \$856,900/day for base case refinery, Table 21) by \$118,900/day (or \$43.4 million/year) for the co-processing case; however, this advantage is offset by the high catalyst (\$78,800/day versus \$5,400/day) and operating costs (\$231,100/day versus \$97,500/day) of the co-processing base case. As a result, gross profits would be \$147,700/day (or \$5.37 million/year) less in the co-processing base case compared to the refinery base case. There is clearly no incentive for installation of co-processing capacity under current economic conditions. The economics for co-processing/refinery complex become marginally attractive (IRR of 13%) compared to the conventional refinery (IRR of 11%), when crude and product prices increase as illustrated by 1981 high crude price cases in Table 21. However, the return on investment for both conventional and co-processing refinery cases is still below what would be considered attractive (IRR of at least 20%) by most refineries.

When compared with the low-severity co-processing refinery base case,³ the IRR for high-severity co-processing refinery is significantly higher (8.8% versus 15%). For quick reference, the economic summary table from reference 3 which evaluated UOP's low-severity co-processing concept is reproduced here as Table 22. Though there is no direct comparison between Table 21 and Table 22 because of different cost numbers and the differences explained in the next paragraph, one obvious difference between the low-severity and the high-severity co-processing is the costs associated with catalysts required in the co-processing unit. This better than 6% improvement observed in IRR for the high-severity co-processing refinery can be directly attributed to the improvement in catalyst (Mo-based catalyst versus V-based catalyst) achieved in the current contract.

When comparing Table 21 and Table 22, one would expect similar results for Conventional Refinery - High Crude Price Case because the raw material costs and product prices are same for this case in both analysis. However, such is not the case because of slightly different assumptions were made in the optimization of the LP Model for the two analysis. In the 1981 study, the IPL crude and the Lloydminster crude flow rates were adjusted to optimize the IRR; whereas, in the present study the crude flow

rates were kept constant at 50,000 bpd, each. In the 1981 study the refinery was designed to treat 40,000 bpd of IPL crude and 50,000 bpd of the Lloydminster crude. Details of the present study High Crude Price Conventional Refinery Case are presented in Appendix D.2 and details of the 1981 study High Crude Price Conventional Refinery Case are presented in Appendix C.2 of reference 3.

In the current design, the raw material costs are about \$76,200/day (or \$27.8 million/year) higher for the co-processing refinery base case than in the conventional refinery. Most of this (94%) is accounted for by the purchase of coal. The balance of the difference is due to different requirement for the butane for alkylation and gasoline blending. This additional raw material costs in the co-processing refinery are more than offset by additional product revenues (\$2,222,800/day for co-processing refinery versus \$2,027,700/day for conventional refinery) of \$195,100/day (or \$71.2 million/year).

Blending pool summaries for the four cases studied are shown in Table 23. Total gasoline production for the co-processing base case refinery is increased by 4,218 barrels/stream day (BPSD), accounting for \$36.7 million/year in revenues over the base case refinery. Production of Fuel Oil No. 6 is also significantly higher for the base case co-processing refinery (13,614 BPSD) than the conventional refinery (4,111 BPSD) and accounts for another \$46.8 million/year in revenue. No. 2 fuel oil production is only slightly more (25,512 BPSD versus 24,502 BPSD) and refinery fuel oil is considerably less (79.5 BPSD versus 1,434 BPSD) for the base case co-processing refinery in comparison to the conventional refinery. Jet fuel production comes from the straight-run material, and therefore, does not change in these two cases.

In the co-processing refinery case, the Lloydminster resid is sent to the co-processing unit first rather than directly to the coker. In the high-severity co-processing case conversion of non-distillable is quite high, so very little charge is fed to the delayed coker. This is evident in the feed to the delayed coker, which is 22,938 BPSD for the base case refinery versus 1,568 BPSD for base case co-processing refinery (see

Appendix D). Correspondingly, the coke production is reduced by about 1,025 MT/day (accounting for a difference in revenue of \$11 million/year) for the co-processing refinery. Sulfur entering with coal in the co-processing refinery ends up as product sulfur and contributes to a major portion of the additional \$2.9 million/year in product sales revenue over the base case conventional refinery.

Operating costs for the co-processing case are more than two times higher than the conventional refinery base case (\$231,100/day versus \$97,500/day). The single biggest contribution to operating expenses comes from the catalyst costs for the co-processing unit. Whereas catalyst costs account for only about 5.5% of operating costs in a conventional refinery, these costs are responsible for 50% of the operating costs in the base case co-processing refinery. Any effort made to increase the effectiveness of the catalyst and to reduce catalyst requirements and hence, the catalyst costs will help considerably in improving the IRR of the co-processing refinery.

Another major component of the operating cost is the maintenance and insurance expense. Both of these items are directly related to total capital costs. The erected cost of the co-processing refinery (\$895.6 million) is more than 30% greater than that of a conventional refinery (\$681.4 million). Of the \$214 million of additional capital costs in the co-processing unit, about 85% (\$183.6 million) are due to the co-processing unit itself. Significant capital costs are also associated with many conventional upgrading units. Table 24 gives a capital cost breakdown by unit for each of the four cases discussed. Because of the much larger hydrogen demands, the co-processing cases require a separate hydrogen plant to augment the hydrogen produced by the platforming unit. This adds about \$30 million to the erected cost of the complex. Sulfur is more easily removed from the coal than it is from the resid in a conventional refinery containing a coker unit. Co-processing of the coal thus results in significantly increased H_2S loads on the amine treating and sulfur recovery units of the refinery. Increasing the capacity of these units adds another \$7 million to the erected cost of the complex. Additional offsite costs (included in the cost of the co-processing unit) are required for storing and handling the

coal. Other offsite costs are also increased by about \$25 million to handle the additional utility conversion requirements and product storage capacity needed as a result of the coal unit. As discussed above, feed to the delayed coker is reduced drastically in the co-processing refinery. Reduction in the size of the coker will save about \$41 million. Much of these savings are distributed over the hydrotreaters and other process units, which are of somewhat larger size in the co-processing refinery than the conventional refinery.

Utility costs for the co-processing refinery case are about three times higher than those for the conventional refinery. Utility requirements broken down by the processing units are summarized in Table 25. Major contributor to the increased utility requirement for the co-processing refinery are the co-processing unit (power requirements for coal grinding, drying, and transport) and the hydrogen plant. Delayed coker because of its smaller capacity in the co-processing refinery uses considerably less utilities.

Table 26 compares the hydrogen usage, by units in the base case conventional refinery with that for the base case co-processing refinery. In the conventional refinery the platforming unit supplies more hydrogen than the demand and about 48% of the hydrogen (38.2 MT/day) is burned as fuel. In comparison, the demand for hydrogen in the co-processing unit of the base case co-processing refinery is quite high (165.6 MT/day or 83% of the total hydrogen demand) and a hydrogen plant is required to supply this excess hydrogen. Propane and saturated gas produced in the co-processing refinery are used as feed to this hydrogen plant.

3.5 HIGH CRUDE PRICE CASE

If crude prices were to increase from their current low levels back to 1981 levels of about \$35/barrel, the IRR from a co-processing refinery (13%) would exceed that from a conventional refinery (11%). As crude and product prices rise in tandem, the refiner's margin per unit feed increases, if many operating and capital costs either remain fixed or

increase at a much smaller rate. This makes it easier for the refiner to justify capital investment. Because of the relative abundance of coal relative to petroleum, it is further assumed that the price of coal would not increase as much as that of the petroleum. This adds a further economic advantage for the co-processing refinery, which derives about 14 wt-% of its total feed from coal rather than crude. The co-processing refinery is also able to shield a greater percentage of its increased profits from taxes, by virtue of the greater depreciation allowance that arises from its higher capital investment cost. Thus, while the conventional refinery shows a sharp drop in the IRR (from 20% to 11%) as the price of the crude increases from about \$12 to \$35/barrel, the IRR for the co-processing refinery case drops modestly from 15% to 13%. This trend is opposite to the trend observed in UOP's previous economic assessment of low-severity co-processing³. As shown in Table 22, in the previous analysis the IRR increased from 8.8% to 13.8% when the crude prices increased from \$17 to \$35/barrel. An explanation for this reversal in trend can be found by comparing the relative increase in the price of the products versus the price of the raw materials or by comparing the gross profits. One notices that for the previous economic assessment the product prices rose in tandem with the crude prices as shown in Table 22. When raw material costs rose by about 200%, the product prices also rose by 180%. However, in the present assessment the raw material costs rose at a much higher rate of about 300%, whereas the product prices rose only by 200% (Table 18). This difference in the rate of increase in the raw material costs and product sales will adversely affect the gross margins for the current analysis.

For the previous economic assessment, the gross profit for the conventional refinery case increased by 4% (base case gross profit: \$399,100/day, high crude prices gross profit: \$416,200/day; Table 19) and for the co-processing refinery case by 45% (base case gross profit: \$289,500/day, high crude prices gross profit: \$421,400/day; Table 19). In comparison, for the present case the gross profit for the conventional refinery decreased by 42% (base case gross profit: \$759,400/day, high crude prices gross profit: \$442,100/day; Table 21) and for the co-processing refinery by 7% (base case gross profit: \$744,700/day, high crude prices gross profit: \$694,500/day; Table 21). Since the capital

investment in all cases is comparable to each other, an increase in gross profit results in an increase in IRR (Table 22) and a decrease in the gross profit translates in a decrease in the IRR for the current assessment (Table 21). These differences between the two assessments illustrate that high crude prices may not necessarily make co-processing attractive and it will be the relative difference between the product sales and the raw material costs (or gross profits), which will determine the attractiveness of co-processing to a refiner.

3.6 Sensitivity of IRR to the Price of Coal

The price of the coal assumed for the base case co-processing refinery is an average of the contract coal price for coal delivered to utilities in midwest region for the year 1993. These contract prices are relatively higher than the coal price on the open market. To see the effect of a change in the coal price on the overall economics of the co-processing refinery a sensitivity analysis was performed by changing the coal price from \$15/ton to \$30/ton. Results of this analysis for the base case co-processing refinery are summarized in Table 27. The effect of coal price on the IRR is not very drastic. Even when the coal price changes to as low as \$15/ton the IRR increases by less than 1% from 14.9% (for coal at \$34/ton) to 15.8% (for coal priced at \$15/ton). At \$34/ton, coal costs amount to 5.8% of the total raw material costs and at \$15/ton, the coal costs are only 2.6% of the total raw material costs. Since the impact of coal costs on the total raw material costs is so small in the coal price range of \$15/ton to \$34/ton, overall effect of coal price on the IRR is not very drastic.

While the IRR of the co-processing refinery is higher at lower coal prices and also it is higher than that for a conventional refinery at higher crude price levels, the absolute rate of return on the capital investment is still low. It is unlikely that a refiner would make the large capital investment required to build a unit without additional incentives, since profits are not large enough to justify the capital risks involved. Also, with the present

environment of relatively low crude prices and relatively low inflation rate (which will translate in very modest increases in product sales), it is highly unlikely that building a co-processing or even a gross-roots petroleum crude refinery will become attractive to the refiner. Very little inflation or upward pressure is expected on the crude prices for the foreseeable future (10-15 years in the future). Another unknown that the refiners face today is the impact of the present environmental laws and regulations on the economics of building a refinery. As the refiners are asked to comply to tougher environmental regulations, the economic incentive to build a grassroots refinery decreases considerably.

In the past the U. S. Federal Government has supported large synfuels projects to increase the Nation's energy security and to promote the use of domestic coal resources. However, with the end of the cold war, it will become increasingly difficult to justify large-scale federal investment in a synfuels project. One reason to consider such an investment would be to balance the trade deficit which is adversely impacted by the large quantities of crude which is imported to support the Nation's demand for transportation fuel. In co-processing significant quantities of the transportation fuel can be obtained by converting coal and thus reducing the amount of crude required for the production of the fuels.

4.0 OBSERVATIONS AND SUMMARY

4.1 CONTRACT ACCOMPLISHMENTS

All the contractual tasks were completed. Autoclave experiments were carried out to evaluate dispersed vanadium catalysts, molybdenum catalysts, and a less costly UOP-proprietary catalyst preparation technique. Autoclave experiments were also carried out in support of the continuous pilot plant unit operation and to study the effects of the process variables (pressure, temperature, and metal loading on the catalyst).

A total of 24 continuous pilot plant runs were made. Research and development efforts during the pilot plant operations were concentrated on addressing the cost effectiveness of the UOP single-stage slurry catalyzed co-processing concept based on UOP experience gained in the previous DOE contract (DE-AC22-84PC70002). To this end, effect of catalyst metal concentration was studied and a highly-active Mo-based catalyst was developed. This catalyst enabled successful long-term operation (924 hours) of the continuous bench-scale plant at highly severe operating conditions of 3,000 psig, 465°C temperature, and 2:1 resid-to-MAF coal ratio with 0.1 wt-% active metal. The metal loading of the catalyst was low enough to consider the catalyst as a disposable slurry catalyst. Also, liquid recycle was incorporated in the pilot plant design to increase the reactor back mixing and to increase the flow of liquid through the reactor (to introduce turbulence in the reactor). This coupled with the highly-active catalyst extended the operability of the pilot plant to a much higher temperature of 465°C (compared to 425°C attained under the previous contract with a vanadium catalyst). Under this high-severity conditions, coal conversions greater than 91 wt-%, heptane insoluble conversions greater than 77 wt-%, and non-distillable (510°C+ material) conversions greater than 83% were achieved.

Alternate catalytic environments were investigated by operating the continuous pilot plant reactor in co-current and counter-current mode with and without packing. Operational problems were encountered during these tests; however, these were believed to be related to the hydrodynamics of the reactor. With further understanding of the hydrodynamics and proper reactor design these problems could be overcome.

The continuous pilot plant data were used to develop conceptual commercial plant designs for the UOP single-stage slurry catalyzed co-processing flow scheme. Basic engineering work was completed to provide the process flows, heat and weight balances, and equipment selection information necessary to generate capital and operating cost estimates for each of the units. The estimated erected cost for a co-processing unit handling 25,683 BPSD of Lloydminster vacuum resid and 2,410 MT/d of as-received Illinois No. 6 coal would be \$183.6 million. This cost includes the coal preparation, catalyst preparation, and solids/resid separation sections.

A refinery economic model was constructed containing the co-processing unit and downstream upgrading units. This model facilitated the economic analysis and optimization, using linear programming techniques, of the upgrading and blending operations needed to produce salable finished products from intermediate co-processing products. At current (average 1993) crude and finished product prices, the internal rate of return on the investment for a refinery including a co-processing unit would be 14%, compared to 20% for the same refinery without the co-processing unit. At the high 1981 crude and product price levels, the IRR for the co-processing refinery becomes relatively more attractive; however, it is still not high enough for a refiner to justify investment in such an integrated refinery.

4.2 ASSESSMENT OF CO-PROCESSING

Co-processing technology makes possible the simultaneous upgrading of heavy petroleum resid and coal to higher-quality transportation fuels. Both of these feedstocks will become increasingly important as and when petroleum resources are exhausted. Each feedstock presents a number of technical challenges. Coal contains ash-forming mineral materials that must be removed, complex aromatic structures that must be converted or removed to get hydrocarbons that meet finished product fuel specifications. Most importantly, the hydrogen/carbon ratio must be increased from less than 0.9 to about 2.0. The heavy petroleum resid contains high levels of metal contaminants, asphaltenes, hetero-atoms and carbon residue, which make the resid economically unattractive to upgrade. These contaminants lead to catalyst deactivation, coke formation, high gas make and reactor fouling in downstream upgrading processes.

Coal liquefaction requires a solvent to provide transfer of hydrogen to the fragments of the coal structure as it is broken apart by thermal reactions. If the effective hydrogen transfer is not accomplished, coke is formed, and the yield of useful liquid products is reduced. In most traditional coal liquefaction processes, considerable expense is incurred in separating a stream from the products, hydrotreating it, and recycling it back to the liquefaction reactor to provide the solvent for hydrogen transfer. In co-processing, the petroleum resid is used directly as the solvent. There are significant advantages to this arrangement. Solvent recovery and hydrotreating expenses are eliminated. The properties of the solvent are decoupled from the coal. The resid "solvent" simply leaves the reactor as a product. At the same time, the quality of the resid is improved by reduction of hetero-atom, asphaltene and metals levels, making subsequent upgrading operations much simpler. Metals removal has been shown to be greater in the presence of coal than if the resids were upgraded alone. Since petroleum refineries already have an infrastructure in place for the upgrading and distribution of fuel products, considerable capital expenses can be saved by tying the coal processing unit into an existing refinery.

The UOP single-stage slurry catalyzed co-processing concept has a number of attractive features. Its single-stage operation and highly-active throw-away catalyst allows for process simplicity and reduced capital outlay. Once-through operation eliminates the expense of any solvent recycle. The catalyst is present in slurry form, which provides continuous contact between the hydrogen, resid, and coal throughout the process. This helps reduce coking in the preheater and separator. The finely-divided catalyst eliminates any diffusion problems that might be associated with a supported catalyst. The highly-active UOP catalyst coupled with properly designed commercial reactor will make the process operable at relatively higher temperature (465°C) and will give high conversions of coal to liquid products. Despite the high operating temperature, conversion to light ends is controlled and increased amounts of usable liquid products are obtained. The objective of UOP conceptual process is not to produce end-use products from the co-processing operation but to take advantage of the existing refinery processes, which have been optimized over many years of operating experience and can be used to most efficiently upgrade the co-processed materials to finished products.

Over the last decade of UOP involvement in the DOE-sponsored development of the co-processing concept, the technology has progressed considerably and coal/oil co-processing has been demonstrated in a relatively long, continuous bench-scale operation. However, for commercialization of this technology much need to be learned in optimizing the process and in further reducing the overall cost of processing. Besides, there are several other factors that need to be addressed for the co-processing concept to be commercially viable. A very important and critical factor that needs to be addressed is to make co-processing *acceptable* to the petroleum refiners as a viable and practical extension of petroleum resid processing.

Under current regulatory environment in the U.S. construction of a new grass-roots refinery to process coal-based liquid feeds is highly unlikely and since refiners have huge capital invested in petroleum refineries it is logical to assume that co-processing is a

reasonable means to introduce coal into the refineries. However, refiners have been very slow to make drastic changes to the feed entering into the refinery because in most instances feed changes require considerable and often, quite expensive revamp of the existing processing schemes in a refinery. UOP believes that integration of co-processing in the U.S. refineries will be a gradual process and will only follow the introduction and a wider acceptance of resid (low-value heavy petroleum material) processing in the refineries.

Following the Clean Air Act of 1990, which places a strong emphasis on reducing automotive emissions, gasoline re-formulation has become a priority throughout the refining industry. Conventional gasoline has been linked to the emission of carbon monoxide, nitrogen oxides, photo-reactive unburned hydrocarbons, and air-borne toxic chemicals. With heavier and "dirtier" feeds like resid or coal/resid co-processing products, complying with the environmental regulations, while still meeting the octane requirements of a gasoline pool, will be a much more challenging problem.

4.3 REQUIREMENTS FOR COMMERCIALIZATION OF CO-PROCESSING

UOP has proved the viability of the co-processing concept by completing a long-term operability run in the co-processing mode under the previous DOE contract. A good database on the properties of the co-processing products has been created and some understanding on the refining needs of this products has been gained. Several research areas that will help improve the technology, and eventually, the economics of co-processing have been identified by UOP and these include: study of reactor and process configurations, evaluation of catalysts, enhancement of co-processing reaction, examination of solvent reactivity, and optimization of process variables. Additionally, to increase the yield of the usable products from resid and coal/resid processing, the mechanism of asphaltene conversion, differences between resid and coal asphaltenes, and the effect of process variables on asphaltene conversion need to be understood.

UOP believes that for the successful and cost-effective integration of the co-processing concept in the U.S. refineries, one has to first optimize the processing of resid and then extend this knowledge to the optimization of the processing of coal/resid mixtures. Understanding of the differences between the processing of these two feedstocks would be of vital importance to design and cost the revamp of the resid-refining processes to co-processing-derived feedstock-refining processes.

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Tables

TABLE 1
Lloydminster Crude Assay

Crude		Naphtha	
Gravity, °API	20.7	Boiling Range, °C	C ₅ -190
Sulfur, wt-%	3.15	Yield, vol-%	19.5
Reid Vapor Pressure, kPa	30.3	Sulfur, wt-%	0.079
Pour Point, °C	-32.0	Paraffins, vol-%	55.1
Con. Carbon, wt-%	9.18	Olefins, vol-%	1.5
Viscosity, cSt @40°C	101.0	Naphthenes, vol-%	31.6
Neutralization Value, mg	0.78	Aromatics, vol-%	11.8
Metal, wt-ppm			
Nickel	52.7		
Vanadium	10.5		
C ₄ - Yield, vol-%	0.48		
Kerosine		Distillate	
Boiling Range, °C	190-277	Boiling Range, °C	277-343
Yield, vol-%	9.8	Yield, vol-%	9.5
Gravity, °API	34.9	Gravity, °API	26.5
Sulfur, wt-%	0.505	Sulfur, wt-%	0.845
Aromatics, vol-%	22.4	Aromatics, vol-%	38.9
Aniline Point, °C	52.4	Aniline Point, °C	56.5
Pour Point, °C	-63.3	Pour Point, °C	-31.1
Cetane Number	35.0	Cetane Number	40.0
Smoke Point, mm	19.0	Cetane Index	41.5
		Viscosity, cST @38°C	6.32
Light Gas Oil		Resid	
Boiling Range, °C	343-427	Boiling Range, °C	427+
Yield, vol-%	9.5	Yield, vol-%	51.7
Gravity, °API	20.9	Gravity, °API	6.5
Sulfur, wt-%	1.5	Sulfur, wt-%	5.1
Conradson Carbon, wt-%	0	Pour Point, °C	120
Metal, wt-ppm		Conradson Carbon, wt-%	18.19
Nickel	0	Furol Viscosity, sec	266
Vanadium	0	@121°C	
		Metal, wt-ppm	87.9
		Nickel	165.2
		Vanadium	

This assay (based on one from the Oil and Gas Journal, 12/12/83, pp 122-3) has been modified by UOP to give cuts that are representative of normal refinery operations. The properties shown are for a blend of the crude with pipeline condensate. Typically, 20-25% condensate is added to the crude to facilitate transportation, as it would otherwise be too viscous for normal pumping through a pipeline.

TABLE 2
Federated Light and Medium Alberta Crude Assay
(Interprovincial)

Crude		Naphtha	
Gravity, °API	39.7	Boiling Range, °C	C ₅ -190
Sulfur, wt-%	0.201	Yield, vol-%	30.9
Total Nitrogen, wt-%	0.11	Sulfur, wt-%	0.005
Reid Vapor Pressure, kPa	31.0	Paraffins, vol-%	28.3
Pour Point, °C	-10.0	Naphthenes, vol-%	54.9
Conradson Carbon, wt-%	1.46	Aromatics, vol-%	16.8
Viscosity, cSt @40°C	3.30	RON Clear	56.4
Wax, wt-%	9.12		
Neutralization Value, mg	0		
Metal, wt-ppm			
Nickel	0.07		
Vanadium	0		
C ₄ - Yield, vol-%	3.43		
Kerosine		Distillate	
Boiling Range, °C	190-277	Boiling Range, °C	277-343
Yield, vol-%	18.25	Yield, vol-%	13.73
Gravity, °API	41.0	Gravity, °API	34.6
Sulfur, wt-%	0.068	Sulfur, wt-%	0.197
Aromatics, vol-%	23.0	Aniline Point, °C	80.7
Aniline Point, °C	142.2	Pour Point, °C	-7.5
Pour Point, °C	-40.0	Cetane Number	55.0
Freezing Point, °C	-39.0	Cetane Index	49.0
Cetane Number	47.0	Viscosity, cST @38°C	4.43
Gas Oil		Resid	
Boiling Range, °C	343-566	Boiling Range, °C	566+
Yield, vol-%	28.29	Yield, vol-%	8.83
Gravity, °API	25.6	Gravity, °API	15.2
Sulfur, wt-%	0.31	Sulfur, wt-%	0.54
Total Nitrogen, wt-%	0.16	Pour Point, °C	26.1
Pour Point, °C	40	Viscosity, cSt @135°C	30.64
Conradson Carbon, wt-%	0.45	Metal, wt-ppm	
Viscosity, cSt @50°C	29.32	Nickel	9.5
Metal, wt-ppm		Vanadium	5.7
Nickel	1.24		
Vanadium	0		

This assay is from the Oil and Gas Journal, 12/12/83, pp 122.

TABLE 3
Pilot Plant Results from Run No. 37

Operating Conditions:

Lloydminster Resid to MAF Illinois No. 6 Coal Ratio	2:1
Reactor Outlet Pressure, psig	3,000
Recycle Liquid to Fresh Feed Ratio	8.4
Catalyst, wt-% molybdenum	0.1

Average of Periods	13, 20, 35	51, 62, 72
Temperature, EC	455	465
Normalized Conversion, wt-%		
Coal	90.2	91.4
C ₇ insolubles	74.0	77.3
Non-distillable (510°C+)	76.2	84.0
Product Distribution, wt-% on total feed basis		
H ₂ O + CO _x	4.65	5.64
H ₂ S	4.75	4.14
C ₁ -C ₄	7.49	11.26
C ₅ -510°C+	63.72	66.65
510°C+	17.74	11.23
Unconverted MAF Coal	3.27	2.88
H ₂ Consumption, wt-%	1.62	1.80
Desulfurization, %	75.7	83.7
Denitrification, %	63.0	69.3

For a detailed description of definition and calculation procedure, see Appendix E.

Normalized Coal Conversion, wt.% = 100% (1-(Unconverted coal/MAF feed coal)/(wt-% MAF recovery/100%)), where unconverted coal is measured as toluene insolubles less ash.

Heptane Insoluble (HI) Conversion = 100% (1- (HI in liquid product + unconverted coal)/(HI in feed resid + MAF feed coal))

Non-distillable (ND) Conversion = 100% (1-(ND in liquid product + unconverted coal)/(ND in feed resid + MAF feed coal))

TABLE 4
Average Product Properties from the High Severity Run
(Run No. 37, 465°C, Period Nos. 51, 62, and 72)

Co-Processed Naphtha		Co-Processed Diesel	
Boiling Range, °C	C ₇ -177	Boiling Range, °C	177-343
Gravity, °API	50.1	Gravity, °API	27.3
Sulfur, wt-%	0.13	Sulfur, wt-%	0.5
Nitrogen, wt-%	0.12	Nitrogen, wt-%	0.45
Bromine Number	22	Bromine Number	16
Composition, vol-%		Composition, vol-%	
Paraffins	47.2	Paraffins +	40.8
Olefins	2.6	Naphthenes	51.3
Naphthenes	31.0	Olefins	7.9
Aromatics	19.2	Aromatics	-42
		Pour Point, °C	27
		Cetane Number	8.6
		Smoke Point, mm	-23
		Freeze Point, °C	
Co-Processed Vacuum Gas Oil		Co-Processed Resid	
Boiling Range, °C	343-570	Boiling Range, °C	570+
Gravity, °API	5.4	Gravity, °API	-19
Carbon, wt-%	86.4	Sulfur, wt-%	1.9*
Hydrogen, wt-%	8.6	Hydrogen, wt-%	6.2
Sulfur, wt-%	1.6	Nitrogen, wt-%	0.15*
Nitrogen, wt-%	0.67	Metals, ppm	
MCRT, wt-%	2.6	Nickel	60*
Metals, ppm		Vanadium	54*
Nickel	< 1	Heptane Insolubles, wt-%	96
Vanadium	< 1	%	54
		Conradson Carbon, wt-%	19.7
		Engler to coke, wt-%	23.7
		Gas	56.6
		Overhead	
		Coke	

* These analyses were obtained from a separate sample.

TABLE 5
Comparison of Pilot Plant Results

	High-Severity Run	Low-Severity Run
Run Number	37	19
Operating Conditions		
Ave. Reactor Temp., °C	465	425
Reactor Pressure, psig	3,000	3,000
H ₂ to Reactor, SCFB	11,800	8,100
H ₂ to Feed, SCFB	6,800	5,500
Liquid Recycle/Feed Ratio	8.4	0
Catalyst	Mo	V
Catalyst (metal) in Feed, wt-%	0.1	1.0
Feed		
Coal	Illinois No.6	
Wt.% MAF Coal	33.3	34.25
Resid Type	Lloydminster Vacuum	
Wt. % Resid	66.7	65.75
Normalized Conversion, wt-%		
Coal	91.3	89.3
C ₇ insolubles	77.3	79.7
Non-distillable (510°C+)	84.0	55.8
H ₂ Consumption, wt-%	1.80	2.35
Desulfurization, %	83.7	63.0
Liquid Product (C ₅ -510°C) Yield, wt-%	66.0	54.0

TABLE 6
Comparison of Co-Processed Naphtha Properties

	High-Severity (Run No. 37)	Low-Severity (Run No. 19)
Naphtha (C₇-177°C) Yield		
BPD	6,770	2,820
vol-% Fresh Feed	18.7	7.7
Gravity, °API	50.1	51.3
Distillation, °C		
IBP	91	70
10	106	91
50	135	138
90	162	179
EP	187	195
Sulfur, wt-%	0.1	0.64
Nitrogen, wt-%	0.1	0.09
Paraffins, vol-%	47	50
Olefins, vol-%	3	10
Naphthenes, vol-%	31	30
Aromatics, vol-%	19	10

TABLE 7
Comparison of Co-Processed Diesel Properties

	High-Severity (Run No. 37)	Low-Severity (Run No. 19)
Diesel (177-343°C) Yield		
BPD	12,300	7,050
vol-% Fresh Feed	34.0	19.3
Gravity, °API	27.3	24.3
Sulfur, wt-%	0.50	1.50
Nitrogen, wt-%	0.45	0.32
Olefins, vol-%	51.3	40.3
Aromatics, vol-%	7.9	5.1

TABLE 8
Comparison of Co-Processed Gas Oil Properties

	High-Severity (Run No. 37)	Low-Severity (Run No. 19)
Gas Oil (343-570°C) Yield		
BPD		
vol-% Fresh Feed	8,470	14,500
	23.4	39.8
Gravity, °API	5.4	13.5
Sulfur, wt-%	1.6	1.7
Nitrogen, wt-%	0.7	0.4
Residual Carbon, wt-%	2.6	1.8

TABLE 9
Comparison of Co-Processed Resid Properties

	High-Severity (Run No. 37)	Low-Severity (Run No. 19)
Resid (Non-distillable, 570+°C) Yield		
BPD		
vol-% Fresh Feed	2,960	9,560
	8.2	26.2
Gravity, °API	-19.0	-4.8
Sulfur, wt-%	1.9	2.7
Nitrogen, wt-%	0.15	0.89
C ₇ Insolubles, wt-%	96	22
Residual Carbon, wt-%	54	28

TABLE 10
Simplified Material Balance for
High Severity Co-Processing Unit

<u>Feed Streams</u>	<u>lb/hr</u>	<u>BPD</u>
Lloydminster Vacuum Resid	384,217	25,683
MF Illinois No. 6 Coal	212,219	(2,311 MT/D)
Catalyst Slurry (1:1 Catalyst/VGO)	2,368	
Makeup Hydrogen	16,804	
Water Injection	28,245	
Stripping Steam	<u>23,000</u>	
	666,853	
 <u>Product Streams</u>		
Waste Water	71,575	
Vent Gas Streams to Scrubber and Saturated		
Gas Condensate	57,172	
Stripper Offgas (C ₄ -)	42,656	
Stripper Overhead Liquid (C ₅ /C ₆ Light Naphtha)	13,145	1,508
C ₇ -380°F Naphtha	96,089	8,816
380-650°F Diesel	155,571	11,960
650-1050°F LVGO + HVGO + Slop Oil	121,906	8,644
1050°F + Vacuum Resid	67,343	4,232
Granulated Catalyst/Coal Solids	<u>41,396</u>	
	666,853	

TABLE 11
Estimated Erected Costs for Major Sections

	<u>EEC in million \$</u>	<u>%</u>
Coal Preparation (Storage, Handling, Grinding, Drying, Slurry Preparation)	59.8	32.6
Catalyst Preparation	3.2	1.6
Reactor/Fractionation	100.0	54.5
Solids/Resid Separation	19.4	10.6
Additional Offsites	<u>1.2</u>	<u>0.7</u>
TOTAL EEC	183.6	100.0

TABLE 12
Major Cost Items in the Co-Processing Plant
(Over \$1 Million)

<u>Reactor/Fractionation Section (Quantity)</u>	<u>FOB Equipment million \$</u>	<u>% of Section Total</u>
Feed Charge Pumps (4)	2.6	4.9
Reactors (6)	17.4	32.7
Cold Separator	1.5	2.8
Recycle Gas/Hot Separator Overhead Exchangers	1.1	2.1
Intermediate Separator Overhead Cooler	5.0	9.4
Recycle Compressor (2)	1.5	2.8
Makeup Compressors (3 @ 50%)	9.1	17.1
Fractionator Gcharge Heater	1.7	3.2
Recycle Hydrogen Heater	2.3	4.3
 Solids/Resid Separator Section		
Rotary Vacuum Filters (5)	1.3	20.3
Belt Expressors (5)	1.8	27.2

TABLE 13
Summary of Utility Requirements for
High Severity Co-Processing

	<u>Power</u>		<u>Fuel</u>		<u>Steam, 10³ lb/hr</u>		<u>CW</u>
	<u>HP</u>	<u>%</u>	<u>10⁶ Btu/hr</u>	<u>%</u>	<u>150 lb</u>	<u>600 lb</u>	<u>gpm</u>
Coal Preparation	10,335	30.8	20.27	9.7	----	----	22
Catalyst Preparation	123	0.4	0.46	2.2	(1.5)	----	145
Reactor/Fractionation	21,275	63.2	202.00	88.1	(146.5)	----	4,420
Solids/Resid Separation	1,856	5.5	----	----	4.5	27.6	178
Additional Offsites	<u>20</u>	<u>0.1</u>	<u>----</u>	<u>----</u>	<u>----</u>	<u>----</u>	<u>----</u>
Total	33,609	100.0	222.73	100.0	(143.5)	27.6	4,765
Unit Cost	\$0.05/kWh		\$2.25/10 ⁶ Btu		\$3.2/10 ³ lb		\$0.05/10 ³ gal
Yearly Cost, million \$	9.92		3.97		(3.63)		0.11

TABLE 14
Major Utility Consumption Items

Power	<u>HP</u>	% of Section <u>Total</u>
<u>Coal Preparation Section</u>		
Storage and Reclaiming	775	7.5
Pulverization and Drying	5,900	57.1
Coal Slurry and Pumping	3,660	35.4
<u>Reactor/Fractionation Section</u>		
Slurry Charge Pumps	1,750	8.2
Recycle Compressor	2,000	9.4
Makeup Compressor	16,000	75.2
<u>Solids/Resid Separation Section</u>		
Rotary Vacuum Filters	600	32.3
Fuel	<u>MM</u> <u>Btu/hr</u>	% of Section <u>Total</u>
<u>Reactor/Fractionation Section</u>		
Recycle Hydrogen Heater	76.4	37.8
Fractionator Charge Heater	92.8	46.0
Vacuum Column Charge Heater	<u>32.8</u>	<u>16.2</u>
	202	100

TABLE 15
Comparison of Estimated Erected Costs for Major Sections

Note: The costs for the low-severity case are from Reference 3. These 1988 costs are multiplied by a factor of 1.2.

	High-Severity Case <u>million \$</u>	Low-Severity Case <u>million \$</u>
Coal Preparation (Storage, Handling, Grinding, Drying, Slurry Preparation)	59.8	59.8
Catalyst Preparation	3.2	17.5
Reactor/Fractionation	100.0	120.0
Solids/Resid Separation	19.4	19.4
Additional Offsites	<u>1.2</u>	<u>1.2</u>
TOTAL EEC	183.6	217.9

TABLE 16
Comparison of Major Cost Items in the Co-Processing Plant
(Over \$1 Million)

<u>Reactor/Fractionation Section</u>	High-Severity FOB Equip. <u>million \$</u>	Low-Severity FOB Equip. <u>million \$</u>
Feed Charge Pumps	2.6	4.0
Reactors	17.4	9.8
Cold Separator	1.5	
Recycle Gas/Hot Separator Overhead Exchangers	1.1	1.4
Intermediate Separator Overhead Cooler	5.0	10.6
Recycle Compressor	1.5	1.4
Makeup Compressors	9.1	4.8
Fractionator Charge Heater	1.7	2.0
Recycle Hydrogen Heater	2.3	
Feed/Hot Separator Bottoms Exchangers		18.0
Hot Separator		1.6
Intermediate Separator		1.3
Vacuum Column Charge Heater		1.4

TABLE 17
Catalyst Operating Costs
for a Throw-Away Operation

	<u>High-Severity Case</u>	<u>Low-Severity Case¹</u>
Catalyst Mixture (MT/D)	12.9	219
Feed (MT/D)	6,495	6,538
Catalyst Metal	Mo	V
Wt-% Metal	54.3	28.4
Catalyst Metal (MT/D)	7.00	62.20
Feed Wt-% Metal	0.11	0.95
Metal Oxide (\$/lb)	2.65	6.65
Metal Price \$/lb	3.98	11.87
Metal Price \$/MT	8,747	26,118
Catalyst Cost, \$/D	61,000	1,624,000
Final Cost ² , \$/D	73,000	1,948,000
Operating Costs, million \$/yr	26.6	711.0

¹ In the low-severity case (Reference 3, 1988), the catalyst was recycled to avoid these higher operating costs.

² The Final cost includes an estimation of costs required for the step of catalyst precursor activation. This cost is estimated at 20% higher than the catalyst cost.

TABLE 18
Product Specifications for the LP Model

Ratio of Premium/Regular in the Gasoline Pool: 1:2		
Unleaded Premium Gasoline	Research Octane Motor Octane RVP	95.5 minimum 86.5 minimum 9.0 maximum
Unleaded Regular Gasoline	Research Octane Motor Octane RVP	91.5 minimum 82.5 minimum 9.0 maximum
Jet Fuel (Kerosine)	Sulfur Smoke Point	0.1 wt-% maximum 21 minimum
No. 2 Fuel Oil	Sulfur Cetane Index	0.05 wt-% maximum 40 minimum
No. 6 Fuel Oil	Sulfur Viscosity Index Specific Gravity	1.0 wt-% maximum 37.5 maximum 0.985 maximum
Refinery Fuel Oil	Sulfur Viscosity Index	1.5 wt-% maximum 45.0 maximum

TABLE 19
Feed and Product Pricing

<u>Feeds</u>	Base Case Ave. of 1993 <u>Midwest Prices</u>	High Crude Price Case Avg. of 1981 <u>Prices</u>
Lloydminster Crude, \$/bbl	10.48	33.00
IPC Crude, \$/bbl	13.47	38.36
Illinois No. 6 Coal (AR), \$/ton	34.00	35.00
n-C ₄ , \$/bbl	15.41	24.70
i-C ₄ , \$/bbl	16.81	28.70
 <u>Products</u>		
LPG, \$/bbl	13.50	31.10
Unleaded Premium Gasoline, \$/bbl	24.76	51.95
Unleaded Regular Gasoline, \$/bbl	23.33	44.10
Kerosine/Jet Fuel, \$/bbl	22.26	42.92
Fuel Oil No. 2, \$/bbl	21.22	40.74
Fuel Oil No. 6, \$/bbl	13.49	33.64
High Sulfur Coke, \$/MT	30.00	30.00
Sulfur, \$/MT	100.00	110.00
Power, \$/kWh	0.05	0.06

TABLE 20
Economic Assumptions for the LP Model

Financing:	100% equity
Depreciation:	Straight Line
Tax Life:	10 years
Project Life:	10 years
Federal Income Tax:	34%
Construction time:	5 years
Interest Rate:	8%
Working Capital:	20% of EEC
Maintenance:	3% of EEC
Taxes and Insurance:	2% of EEC

TABLE 21
Economic Summary from LP Refinery Models

	<u>Conventional Refinery</u>		<u>Refinery with Co-Processing Unit</u>	
		High Crude		High Crude
	<u>Base Case</u>	<u>Price Case</u>	<u>Base Case</u>	<u>Price Case</u>
<u>Raw Material Costs (10³\$/D)</u>				
IPL Crude	639.9	1822.1	639.8	1822.1
Lloydminster Crude	497.8	1567.5	497.8	1567.5
Illinois Coal No. 6	--	--	71.6	77.4
Isobutane	15.3	114.4	14.6	114.7
Normal Butane	<u>17.9</u>	<u>75.8</u>	<u>23.1</u>	<u>43.2</u>
TOTAL	1170.8	3579.8	1247.0	3625.0
<u>Product Sales (10³\$/D)</u>				
LPG	32.2	72.8	--	--
Unleaded Premium Gasoline	407.7	872.4	442.4	989.1
Unleaded Regular Gasoline	768.2	1481.2	833.8	1679.3
Jet Fuel	190.0	366.4	190.0	366.4
0.05% S Fuel Oil No. 2	519.9	924.9	541.4	1076.2
1.0% S Fuel Oil No. 6	55.5	351.5	183.6	414.8
Coker	39.9	32.9	9.1	9.6
Sulfur	<u>14.4</u>	<u>14.1</u>	<u>22.2</u>	<u>25.7</u>
TOTAL	2027.7	4116.3	2222.8	4561.1
<u>Operating Costs (10³\$/D)</u>				
Catalyst Costs	5.4	5.1	78.8	82.4
Labor, Maint. & Ins. Costs	77.0	74.8	115.5	120.7
Purchased Power Costs	14.3	13.8	36.0	37.7
Utility Conversion Costs	<u>0.7</u>	<u>0.7</u>	<u>0.8</u>	<u>0.8</u>
TOTAL	97.5	94.4	231.1	241.6
Gross Margin (10 ³ \$/D)	856.9	536.5	975.8	936.1
Gross Profit (10 ³ \$/D)	759.4	442.1	744.7	694.5
EEC (million \$)	681.4	675.6	895.6	910.6
Internal Rate of Return*	20%	11%	15%	13%
Simple Payout Time, years	3.4	5.4	4.3	4.7

*IRR assumptions - 5 year construction, 10 year project life

TABLE 22
Economic Summary from LP Refinery Models
Low-Severity Co-Processing Plant*

	<u>Conventional Refinery</u>		<u>Refinery with Co-Processing Unit</u>	
		<u>High Crude</u>		<u>High Crude</u>
	<u>Base Case</u>	<u>Price Case</u>	<u>Base Case</u>	<u>Price Case</u>
<u>Raw Material Costs (10³\$/D)</u>				
IPL Crude	751.2	1534.4	751.2	1534.4
Lloydminster Crude	797.4	1650.0	799.0	1650.0
Illinois Coal No. 6	--	--	85.7	85.7
Isobutane	64.3	109.8	85.3	145.9
Normal Butane	<u>21.1</u>	<u>32.7</u>	<u>28.7</u>	<u>44.4</u>
TOTAL	1634.0	3326.8	1749.9	3460.3
<u>Product Sales (10³\$/D)</u>				
LPG	--	69.3	--	--
Unleaded Premium Gasoline	458.0	823.1	917.5	973.5
Unleaded Regular Gasoline	776.4	1397.4	917.5	1653.9
Jet Fuel	178.1	336.1	178.3	336.1
0.05% S Fuel Oil No. 2	496.4	897.2	482.0	872.3
1.0% S Fuel Oil No. 6	165.7	268.7	174.9	322.1
Coker	34.7	34.6	28.6	28.6
Sulfur	<u>14.6</u>	<u>14.6</u>	<u>41.8</u>	<u>41.7</u>
TOTAL	2123.9	3841.1	2364.1	4228.3
<u>Operating Costs (10³\$/D)</u>				
Catalyst Costs	4.7	4.7	149.0	149.0
Labor, Maint. & Ins. Costs	72.2	72.5	134.4	135.3
Purchased Power Costs	13.3	20.2	40.7	61.5
Utility Conversion Costs	<u>0.7</u>	<u>0.7</u>	<u>0.8</u>	<u>0.8</u>
TOTAL	90.9	98.1	324.8	346.6
Gross Margin (10 ³ \$/D)	489.9	514.2	614.3	768.0
Gross Profit (10 ³ \$/D)	399.1	416.2	289.5	421.4
EEC (million \$)	545.6	546.1	825.8	827.3
Internal Rate of Return*	20.0%	20.8%	8.83%	13.8%

* This table is reproduced from reference no. 3.

TABLE 23
Blending Pool Summaries

	<u>Conventional Refinery</u>		<u>Refinery with Co-Processing Unit</u>	
	<u>High Crude</u>		<u>High Crude</u>	
	<u>Base Case</u>	<u>Price Case</u>	<u>Base Case</u>	<u>Price Case</u>
(All numbers are in BPD)				
<u>Unleaded Premium Gasoline</u>	16,464	16,793	17,870	19,039
n-C ₄	1,465	1,355	1,504	1,652
Alkylate from Butylene	4,017	--	2,536	4,363
98 RONC Reformate	7,596	9,897	8,108	8,084
Penexate	--	2,474	--	--
FCC Gasoline	3,385	--	5,722	4,817
94 RONC Reformate	--	3,068	--	122
<u>Unleaded Regular Gasoline</u>	32,928	33,588	35,740	38,078
n-C ₄	1,994	2,164	1,947	2,060
Alkylate	584	8,423	1,752	4,020
FCC Gasoline	11,752	14,581	9,534	10,674
98 RONC Reformate	11,392	4,228	12,709	12,923
94 RONC Reformate	--	1,200	--	--
SP Light Naphtha	3,431	2,531	2,542	2,213
Co-Proc. Light Naphtha	--	--	2,736	2,876
Penexate	2,074	461	2,983	3,311
Motor Fuel Polymer	1,699	--	1,537	--
<u>0.05% S Fuel Oil No. 2</u>	24,502	22,702	25,512	26,416
Hydrotreated SR Diesel	13,931	13,931	13,517	13,808
Hydrotreated Coker Diesel	10,572	8,771	331	348
Hydrotreated Co-Proc. Diesel	--	--	11,664	12,260
<u>1.0% S Fuel Oil No. 6</u>	4,111	10,450	13,614	12,332
SR IPL Vac. Bottoms	--	4,227	4,227	4,227
SR Lloyd Vac. Bottoms	--	881	1,179	--
FCC LCO	3,576	3,445	4,634	4,707
FCC Slurry Oil	535	1,896	3,154	3,274
IPL Diesel	--	--	--	124
<u>1.5% S Refinery Fuel Oil</u>	1,434	--	79.5	--
SR Lloyd. Vac. Bottoms	--	--	9.3	--
FCC Slurry Oil	1,434	--	70.2	--

TABLE 24
Capital Cost Breakdown for the Four Cases

<u>Estimated Erected Cost, million \$</u>	<u>Conventional</u> <u>Refinery</u>		<u>Refinery With</u> <u>Co-Processing Unit</u>	
	<u>Base Case</u>	<u>High Crude</u>	<u>Base Case</u>	<u>High Crude</u>
		<u>Price Case</u>		<u>Price Case</u>
Crude Unit	40.7	40.7	40.7	40.7
Vacuum Unit	14.3	14.3	14.3	14.3
Co-processing Unit	---	---	183.6	183.6
Delayed Coker	63.3	55.7	22.5	23.2
Kerosine Hydrotreater	6.2	6.2	6.2	6.2
Diesel Hydrotreater	17.8	16.9	18.2	18.7
Gas Oil Hydrotreater	13.3	12.6	15.4	15.7
FCC Unit	62.8	61.2	67.3	68.0
Catalytic Condensation	7.2	---	6.7	---
Alkylation Unit	22.6	36.0	21.4	35.9
Naphtha Hydrotreater	7.1	7.1	7.7	7.8
C ₅ /C ₆ Penex Unit	5.5	7.0	7.1	7.7
CCR Platforming Unit	34.0	33.0	36.4	36.8
Hydrogen Plant	---	---	29.7	31.2
Amine Tr./Sulfur Plant	21.0	19.3	28.6	29.5
Offsites	<u>365.4</u>	<u>365.4</u>	<u>390.1</u>	<u>391.3</u>
TOTAL	681.4	675.6	895.6	910.6

TABLE 25
Comparison of Utility Consumption by Units

	<u>Conventional Refinery</u>			<u>Refinery with Co-Processing</u>		
	Capacity, BPD	Power, kWh	Fuel, 10 ⁶ Btu	Capacity, BPD	Power, kWh	Fuel, 10 ⁶ Btu
Crude Unit	95,000	58,952	6,558	95,000	58,952	6,558
Vacuum Unit	48,600	13,029	4,000	48,600	13,029	4,000
Co-Processing Unit				23,300	557,115	4,496
				Resid*		
Delayed Coker	28,700	66,332	5,557	3,000	7,671	643
Kerosine Hydrotreater	8,500	7,821	298	8,500	7,821	298
Diesel Hydrotreater	24,200	32,581	900	25,000	30,289	933
Gas Oil Hydrotreater	11,300	21,388	456	13,800	23,249	568
FCC Unit	25,600	30,381		28,200	33,922	
Cat. Cond. Unit	2,600	7,776	664	2,300	7,035	600
Alkylation Unit	4,600	9,090	1,029	4,300	8,471	959
Naphtha Hydrotreater	24,000	20,177	1,883	26,800	17,345	2,114
Penex Unit	2,100	3,199		3,100	4,601	
CCR Platforming Unit	22,100	32,093	5,112	24,000	35,186	5,605
Hydrogen Plant				111.6 ⁺	26,450 ^{**}	6,267
Amine Tr./Sulfur Plant	145.3 ^{**}	15,013	427	225.8 ^{**}	23,336	664
Offsites	95,000	21,713		102,800	25,163	
Internal Utility Conv.		17,463	4,335		21,014	4,923
Total		357,011	31,217		900,652	38,626

* +2447.5 MT/D Coal
+ MT/day hydrogen

** Uses Sat. Gas. and Propane for Feed
** MT/day hydrogen sulfide

TABLE 26
Comparison of Hydrogen Usage by Various Units

	<u>Conventional Refinery</u>	<u>Refinery with Co-Processing</u>	<u>Delta</u>
Co-Processing Unit	---	165.62	+165.62
Kerosine Hydrotreater	1.33	1.33	
Diesel Hydrotreater	18.27	12.52	-5.75
Gas Oil Hydrotreater	12.12	11.50	-0.62
Naphtha Hydrotreater	7.95	5.30	+2.65
Penex	1.94	2.80	+0.86
Excess used for Fuel	38.17	--	-38.17
Total Consumption	79.78	199.07	+119.29
Platforming Unit	79.78	87.47	+7.69
Hydrogen Plant	---	111.60	+111.60
Total Production	79.78	199.07	+119.29

TABLE 27
Sensitivity of Internal Rate of Return to the Coal Price

Coal Price, \$/ton	34	15	20	25	30
	(All costs are in \$/day)				
Coal Costs	71,567	31,574	42,098	52,623	63,147
(as % of Total Raw Materials)	5.74	2.62	3.46	4.29	5.10
Total Raw Material Costs	1,246,994	1,207,001	1,217,526	1,228,050	1,238,575
Total Product Sales	2,222,795	2,222,795	2,222,795	2,222,795	2,222,795
Total Operating Costs	231,195	231,195	231,195	231,195	231,195
Gross Margin	975,801	1,015,794	1,005,270	994,745	984,221
Gross Profit	744,606	784,599	774,075	763,550	753,026
EEC (million \$)	895.6	895.6	895.6	895.6	895.6
IRR	14.88%	15.76%	15.53%	15.30%	15.07%

Figures

FIGURE 1

Simplified Diagram of UOP's Co-Processing Plant

FIGURE 2

Comparison of Pilot Plant Product Yields

FIGURE 3

Block Flow Diagram of Co-Processing Sections

- Feed Streams
- Product Streams

All flow rates are in lbs/hr.
 Numbers are for High-Severity Co-Processing
 unit material balance (see table 10, App. D.3).

FIGURE 4
Co-Processing Reactor Section

FIGURE 5

Co-Processing Fractionation Section

FIGURE 6

Co-Processing Solids/Resid Separation Section

FIGURE 7

Block Representation of Refinery LP Model

FIGURE 7 (Continued)

Key to the Refinery Model Block Diagram

Unit Names

ACU "A" Crude Unit
 AK3 C₃ Alkylation Unit
 AK4 C₄ Alkylation Unit
 ASP Amine Treating & Sulfur Plant
 BCU "B" Crude Unit
 CCU Catalytic Cond. (Poly) Unit
 COK Delayed Coker
 COP Co-processing Unit
 DHT Diesel Hydrotreater
 FCC Fluid Cat. Cracker
 GHT Gas Oil Hydrotreater
 H2P Hydrogen Plant
 KHT Kerosine Hydrotreater
 NHT Naphtha Hydrotreater
 OFF Offsites
 PEN Penex (C₅/C₆ Isom) Unit
 PLT Platforming Unit
 UTY Utility Conversions
 VTA "A" Vacuum Tower
 VTB "B" Vacuum Tower

PRODUCT NAMES

CO Coke
 F2 No. 2 Fuel Oil
 F6 No. 6 Fuel Oil
 JF Jet Fuel
 SU Sulfur
 UP Unleaded Premium Gasoline
 UR Unleaded Regular Gasoline

Stream Names

ADF IPL Diesel
 AHN IPL Heavy Naphtha
 AK3 C₃ Alkylate
 AK4 C₄ Alkylate
 AKE IPL Kerosine
 ALN IPL Light Naphtha
 ARC IPL Reduced Crude
 AVB IPL Vac. Bottoms
 AVG IPL Vac. Gas Oil
 BDF Lloyd. Diesel
 BHN Lloyd. Heavy Naphtha
 BKE Lloyd. Kerosine
 BLN Lloyd. Light Naphtha
 BRC Lloyd. Reduced Crude
 BTU Fuel (MMBTU)
 BVB Lloyd. Vac. Bottoms
 BVG Lloyd. Vac. Gas Oil
 C30 Propylene
 C40 Butene
 CDL Coker Diesel
 CGO Coker Gas Oil
 CNA Coker Naphtha
 COK FCC Coke
 COL Ill. No. 6 Coal
 CWA Cooling Water
 FGA Unsat. Fuel Gas
 GAS Sat. Fuel Gas
 H2S Hydrogen Sulfide
 HCD Hydrotreated Coker Diesel
 HCG Hydrotreated Coker Gas Oil
 HDL Hydrotreated SR Diesel
 HGO Hydrotreated SR Gas Oil
 HHN Hydrotreated Heavy Naphtha
 HKE Hydrotreated Kerosine
 HLN Hydrotreated Light Naphtha
 HPS High Pressure Steam
 HSC High Sulfur Coke
 HVD Hydrotreated Co-Processed Diesel
 HVG Hydrotreated Co-Processed Gas Oil
 HYD Hydrogen
 IC4 Isobutane
 IPL Interprovincial Crude
 LCO Light Crude Oil
 LLD Lloydminster Crude
 LPS Low Pressure Steam
 MPS Medium Pressure Steam
 NC4 Normal Butane
 P92 92 Octane Reformate
 P94 94 Octane Reformate
 P98 98 Octane Reformate
 PEX Penexate
 PLG Poly Gasoline
 PRO Propane
 PWR Power (kWh)
 RFO Refinery Fuel Oil
 SLO FCC Slurry Oil
 SOL Waste Solids
 SOX SO₂/SO₃
 SUL Sulfur
 TAR Alky. Tar
 VHN Co-Processed Heavy Naphtha
 VLN Co-Processed Light Naphtha
 VSD Co-Processed Diesel
 VSG Co-Processed Gas Oil
 VSP Co-Processing Pitch

FIGURE 8

Flow Diagram of a Conventional Refinery

Appendix A
Detailed Analysis of Composite Sample from Run No. 37

Composite Sample Analysis for Run No. 37

Test Periods	51, 62, 72
WHSV, g/hr/cc	0.86
Temperature, C	464.6

MAF Feed, g	6566.88
Product: (MASS BALANCE PROGRAM)	
H2O+COX, wt% MAF Feed	5.64
C1-C4 Gases	
wt% of MAF Feed	11.26
C5+ Gases	
wt% of MAF Feed	3.78
H2S Gas	
wt% of MAF Feed	4.14
H2 Consumed	
wt% of MAF Feed	1.80
Vacuum OVHD, g	3389.00
wt% of MAF Feed	50.49
Vacuum Bottom, g	2103.00
wt% of MAF Feed	31.33
Toluene Soluble, wt% of Btm.	75.35
Toluene Insoluble, wt% of Btm.	24.65
Ash, wt% of Btm. TI	62.70
Ash, wt% of MAF Feed	4.84

TOTAL, wt%	101.80
------------	--------

Normalized Coal Conv., wt%	91.4
Normalized C7 Insol. Conv., wt%	77.3
Normalized 510 C+ Conv., wt%	84.0

Product Fractionation Summary

Vacuum OVHD	
Start with, g	2204
Cut 1 (IBP-177 C), g	646.5
wt% of OVHD	29.3
wt% of MAF Feed	14.8
Cut 2 (177-343 C), g	1212.2
wt% of OVHD	55.0
wt% of MAF Feed	27.8
Cut 3 (343 C +), g	344.8
wt% of OVHD	15.6
wt% of MAF Feed	7.9
Weight Recovery, %	100.0

Composite Sample Analysis for Run No. 37

Toluene Soluble Bottom

Start with, g	943.2	425.0
Cut 1 (IBP-510 C), g	497.8	201.0
wt% of TSB	52.8	47.3
wt% of MAF Feed	12.5	11.4
Cut 2 (510-531 C), g	147.1	55.0 (510-565
wt% of TSB	15.6	12.9
wt% of MAF Feed	3.7	3.1
Cut 3 (531 C+), g	298.3	69.0 (565+)
wt% of TSB	31.6	39.8
wt% of MAF Feed	7.5	9.6
Weight Recovery, %	100.0	100.0

REANALYSIS

Vacuum OVHD, Total Sample

Distillation D1160, C

IBP	69
5 vol%	118
10 vol%	130
20 vol%	152
30 vol%	173
40 vol%	193
50 vol%	213
60 vol%	241
70 vol%	265
80 vol%	300
90 vol%	366
EP @ 98.0 vol%	400

Composite Sample Analysis for Run No. 37

Relative Density D4052RD, g/ml	0.8644
API Gravity D4052	32.2
Carbon Residue MCRT D4530, wt%	0.016
Heptane Insoluble U614, wt%	0.037
Viscosity D445, cSt	
@100 F	1.784
@122 F	1.435
Viscosity Index D2270	Beyond Scope
Pour Point D97, C	-38
Reid Vapor Pressure D323,psig	Cancelled
Petroleum Ash D482, wt%	0
UOP K Factor U375	11.12
C U866, wt%	86.5
H U866, wt%	7.611
O LECO, wt%	0.75
S U864, wt%	0.54
P in Gasoline, D3231, g/US Gal	0.0603
N D4629, wt%	0.3925
Basic N U269, wt%	0.271
Heat of Combustion, BTU	Cancelled
Acid No U565	0.04
Bromine No U304N	17.4
Metals AASAG	
Mo ppm	<0.1
Ni ppm	<0.1
V ppm	0.13
Fe ppm	1.4
Phenol U262H, ppm	Undetermined
GS Simulated Dist D2887H	
IBP	35
5 vol%	97
10 vol%	121
20 vol%	156
30 vol%	181
40 vol%	205
50 vol%	229
60 vol%	255
70 vol%	285
80 vol%	317
90 vol%	352
EP @99.5 vol%	417

Composite Sample Analysis for Run No. 37

Vacuum OVHD Cut 1, IBP-177 c

Relative Density D4052RD, g/ml	0.7672
API Gravity D4052	52.9
C U866, wt%	80.73
H U866, wt%	12.83
O LECO, wt%	0.57
S U864	0.13
N D4629, wt%	0.1207
Phenol U262PH, %	2.2
Thiophenol U262TPH, %	0.09
Micro Motor Octane No D2700M	61.1
Micro Research Octane No D2699M	64.2
Bromine No U304	22.2
PONA UOP880, wt%	
Paraffins	46.8
Olefins	2.3
Naphthenes	31.1
Aromatics	19.8
Total	100
Viscosity D445, cSt	
@100 F	0.667
@122 F	0.592
Distillation D86A	
IBP	55.7
5 vol%	90.9
10 vol%	100.3
20 vol%	110.7
30 vol%	118.5
40 vol%	125.6
50 vol%	133
60 vol%	138.7
70 vol%	145
80 vol%	150.7
90 vol%	158.6
EP @98 vol%	187.4

Composite Sample Analysis for Run No. 37

Vacuum OVHD Cut 2, 177-343 C

Relative Density D4052RD, g/ml	0.8912
API Gravity D4052	27.3
C U866, wt%	84.22
H U866, wt%	11.12
O LECO, wt%	0.77
S U864, wt%	0.5
N D4629, wt%	0.4541
Phenol U262PH, %	2.4
Thiophenol U262TPH, %	0.01
Aniline Point D611, C	31.6
Freezing Point D2386, C	-23.5
Pour Point D97, C	-42
Smoke Point D1322, mm	8.6
Bromine No U304N	16.4
Cetane No D613	26.9
FIA U501, vol%	
Parafins/Naphthenes	40.8
Olefins	7.9
Aromatics	51.3
Total	100
Viscosity D445, cSt	
@100 F	2.431
@122 F	1.924
Distillation D86A, C	
IBP	212.8
5 vol%	215.4
10 vol%	216.1
20 vol%	217
30 vol%	218
40 vol%	225.8
50 vol%	237.5
60 vol%	249.8
70 vol%	264.4
80 vol%	279.6
90 vol%	296.7
EP @ 98 vol%	315.2

Composite Sample Analysis for Run No. 37

Vacuum OVHD Cut 3, 343 C+

Relative Density D4052RD, g/ml	0.9882
API Gravity D4052	11.7
C U866, wt%	85.61
H U866, wt%	9.38
O LECO, wt%	0.8
S U864, wt%	1.27
N D4629, wt%	0.705
Phenol & Thiophenol U262	NA
Pour Point D97, C	2
Viscosity D445, cSt	
@140 F	11.12
@210 F	3.809
Ash D482, wt%	0.002
Metals AASAG, wt%	
Mo, ppm	1.4
Ni, ppm	<1
V, ppm	1.1
Fe, ppm	6.2
Carbone Residue MCRT D4530, wt%	0.09
Distillation D1160, C	
IBP	315
5 vol%	330
10 vol%	337
20 vol%	347
30 vol%	354
40 vol%	361
50 vol%	368
60 vol%	374
70 vol%	381
80 vol%	390
90 vol%	407
EP @ 99.5 vol%	533

Composite Sample Analysis for Run No. 37

GS Simulated Dist D2887H

IBP	315
5 vol%	330
10 vol%	337
20 vol%	347
30 vol%	354
40 vol%	361
50 vol%	368
60 vol%	374
70 vol%	381
80 vol%	390
90 vol%	407
EP @99.5 vol%	533

Toluene Soluble Bottom, Total Sample

Relative Density D70RD, g/ml	1.1135
API Gravity D70	-3.3
Carbon Residue MCRT D4530, wt%	23.95
Heptane Insoluble U614, wt%	32.76
Toluene Insoluble U614M, wt%	0.74
Road Oil Vis. @210F, sec	25.9
Saybolt Viscosity @250 F, secs	54.1
Viscosity Index D2270	Beyond Scope
Pour Point D97, C	41
Softening Point D36, C	48
Petroleum Ash D482, wt%	0.028
UOP K Factor U375	Beyond Scope
C U866, wt%	87.29
H U866, wt%	7.53
O LECO, wt%	0.92
S U864, wt%	1.76
N D4629, wt%	0.47
P in Gasoline, D3231, g/US Gal	<0.0008
Metals AASAG	
Mo ppm	2.7
Ni ppm	15
V ppm	11
Fe ppm	15

REANALYSIS

1.115
-4.6
24.37

Composite Sample Analysis for Run No. 37

Distillation D1160, C	
IBP	387
5 vol%	423
10 vol%	431
20 vol%	446
30 vol%	465
40 vol%	490
50 vol%	516
EP @ 57 vol%	542
D1160 OVHD API	3.6
D1160 OVHD Spec. Grav.	1.0474
D1160 OVHD C, wt%	86.32
D1160 OVHD H, wt%	8.92
D1160 OVHD S, wt%	1.63
D1160 OVHD N, wt%	1.1036
D1160 OVHD Sim Distillation D2887 Mod	
IBP	308
5 wt%	360
10 wt%	377
20 wt%	395
30 wt%	407
40 wt%	419
50 wt%	431
60 wt%	442
70 wt%	455
80 wt%	472
90 wt%	501
EP @96 %	554
D1160 Bottom API	-16.4
D1160 Bottom Sepc. Grav.	1.2294
D1160 Bottom C, wt%	86.62
D1160 Bottom H, wt%	6.53
D1160 Bottom S, wt%	2.07
D1160 Bottom N, wt%	0.246

Composite Sample Analysis for Run No. 37

Toluene Soluble Bottom Cut 1, IBP-510 C

		REANALYSIS
Relative Density D4052RD, g/ml	1.0543	1.0486
API DM	2.7	3.4
Carbon Residue MCRT D4530, wt%	0.454	0.24
Heptane Insoluble U614, wt%	0.247	
Viscosity D445, cst		
@140 F	154	111.3
@210 F	17.93	15.05
Pour Point D97, C	26	17
Petroleum Ash D482, wt%	0	
C U866, wt%	83.31	87.02
H U866, wt%	7.52	8.115
O LECO, wt%	NA	
S U864, wt%	1.67	
P D1091, wt%	NA	
N D4629, wt%	NA	
Metals AASAG		
Mo	<0.1	
Ni	0.1	
V	<0.1	
Fe	3.1	
Distillation D1160, C		D2887 Sim Dist
IBP	373	307
5 vol%	411	362
10 vol%	420	380
20 vol%	431	400
30 vol%	438	414
40 vol%	446	425
50 vol%	453	436
60 vol%	464	447
70 vol%	475	460
80 vol%	488	473
90 vol%	502	492
95 vol%	516	513
EP @ 98 vol%	522	

Composite Sample Analysis for Run No. 37

D1160 OVHD Rel. Dens., g/ml	1.051
D1160 OVHD API Gravity	3.1
D1160 OVHD GLC	
IBP	321
5 wt%	375
10 wt%	390
20 wt%	409
30 wt%	421
40 wt%	432
50 wt%	444
60 wt%	456
70 wt%	469
80 wt%	480
90 wt%	496
95 wt%	507
EP @ 100 wt%	539

Toluene Soluble Bottom Cut 2, 510-531 C

REANALYSIS
(510-565 C)

Relative Density D70RD, g/ml	1.1045	1.105
API D70	-3.4	-3.4
Carbon Residue MCRT D4530, wt%	16.71	11.735
Heptane Insoluble U614, wt%	1.47	
Viscosity D445, cSt		
@210 F	160.7	348.6
@250 F	43.8	73.45
Pour Point D97, C	40	
Petroleum Ash D482, wt%	0	
C U866, wt%	86.33	87.93
H U866, wt%	8.12	7.685
O LECO, wt%	NA	
S U864, wt%	1.89	
N D4629, wt%	0.44	
Metals AASAG ppm		
Mo	6	
Ni	3	
V	0.43	
Fe	19	

Composite Sample Analysis for Run No. 37

D2887 Sim Dist

IBP	275	422
5 wt%	372	470
10 wt%	436	481
20 wt%	500	497
30 wt%	518	508
40 wt%	529	518
50 wt%	537	526
60 wt%	544	533
70 wt%	553	541
80 wt%	563	551
EP @ 90 wt%	575	569

Toluene Soluble Bottom Cut 3, 531 C+

REANALYSIS
(565+C)

Relative Density D70RD, g/ml	1.6044	1.255
API D70	-43.3	-18.8
Carbon Residue MCRT D4530, wt%	74.66	54.45
Heptane Insoluble U614, wt%	96.61	96.2
Toluene Insol., U614M, wt%	32.1	
Road Viscosity, sec		
@500 F	19.5	
Softening Point D36, F	362	
Petroleum Ash D482, wt%	0.134	
C U866, wt%	90.01	88.72
H U866, wt%	5.74	6.22
O LECO, wt%	NA	
S U864, wt%	1.86	
N D4629, wt%	0.149	
Metals AASAG ppm		
Mo	20	
Ni	60	
V	54	
Fe	132	
P	18	
U1 Coke		
Gas, wt%	22.14	19.7
OVHD, wt%	5.76	23.7
Btm (Coke), wt%	72.1	56.6
OVHD Rel Density, g/ml		1.0299
API		5.9

Composite Sample Analysis for Run No. 37

U1 Coke OVHD Sim Dist

IBP	96	89
5 wt%	164	190
10 wt%	197	245
20 wt%	251	330
30 wt%	320	394
40 wt%	391	448
50 wt%	459	495
60 wt%	516	527
70 wt%	546	545
80 wt%	567	564
EP @ 83 wt%	574	570

Toluene Insoluble Bottoms

C U866, wt%	31.24
H U866, wt%	1.73
S U864, wt%	5.55
Metals AASAG	
Mo wt%	2.21
Ni wt%	0.058
V wt%	0.14
Fe wt%	0.0153
TGA ash, wt%	62.7

Composite Sample Analysis for Run No. 37

Plant 558 Run 37
NORMALIZED CONVERSIONS

Coal	91.3
Heptane Insolubles	77.3
Nondistillable (510C+)	84.0

COMPOSITE PRODUCT DISTRIBUTION

Product	wt% MAF Feed*
Gas	
C1-C4	11.26
C5+	3.78
H2S	4.14
Liquid	
Vacuum Column Overhead	50.49
Toluene Soluble	
Vacuum Column Bottom	23.61
H2O + COx	5.64
Solid	
Unconverted Coal/Coke	2.88
H2 Consumption	-1.80
Total	<hr/> 100.00

Ash	wt% MAF Feed*
Ash	4.84

* Normalized to 100 wt% MAF Recovery.

Appendix B
Engineering Specifications for Major Equipment

UOP

VESSELS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal Co ProcessingPROJECT NO 728005-4000PROCESS UNIT Resid/Coal Co ProcessingDATE March 7, 1994

ITEM NO.	V1a/b/c/d/e/f	V2	V3	V4
SERVICE (on drums give Horiz. or Vert.)	Reactor A/B/C/D/E/F (Vertical)	Hot Separator (Vertical)	Hot-Flash (Vertical)	Intermediate Separator (Vertical)
DIAMETER, ft - in	8'	6'	8'	7'
LENGTH, TAN - TAN, ft - in	45'	12'	16'	20'
DESIGN TEMP, °F	920°F	920°F	625°F	570°F
DESIGN PRESSURE, psig	3150	3150	800	3140
MATERIAL OF CONST.				
CARBON STEEL				
OTHER	3' CR-1 MO	2 1/2 Cr-1Mo	1 1/2 Cr - 1/2 Mo	1 1/2 Cr - 1/2 Mo
X-RAY, SPOT OR FULL				
LINING	1/4" TP347 weld only	1/4" TP347 weld overlay	1/4" TP347 or TP321	1/4" TP347 or TP321
SKIRT LENGTH OR ELEV. ABOVE GRADE, ft	8'0"	8'0"	8'0"	8'0"
INSULATION REQ'D?	Yes	Yes	Yes	No
FOR FRACTIONATORS				
NO. OF TRAYS		2	2	
TYPE OF TRAYS		sieve	sieve	
NO. LIQUID PASSES		1	1	
TRAY MATERIAL		TP410SS	TP410SS	
FOR REACTORS				
ASME SECT VIII	✓	✓	✓	
DIV. 1 OR 2	2	2	1	
IF NOT, GIVE DETAILS				
DOWN FLOW	Upflow			
INTERNAL DISTRIBUTOR	✓			
RADIAL FLOW				
OTHER	1/4" 347 weld overlay	1/4" 347 weld overlay	1/4" 347 weld overlay	
REMARKS				

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VESSELS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal CoProcessingPROJECT NO 728005-4000PROCESS UNIT Resid/Coal CoProcessingDATE March 7, 1994

ITEM NO.	V-5	V-6	V-7	V-8
SERVICE (on drums give Horiz. or Vert.)	Cold Separator Horizontal	Cold Flash Horizontal	1st Stage Makeup Comp-Suction Drum (Vertical)	2nd Stage Makeup Comp-Suction Drum (Vertical)
DIAMETER, ft - in	9'	9'	4'	3'6"
LENGTH, TAN - TAN, ft - in	30'	28'	8'	8'
DESIGN TEMP, °F	250	250	250	250
DESIGN PRESSURE, psig	3130	770	325	655
MATERIAL OF CONST. CARBON STEEL	KCS Lined with Incoloy 825	KCS Lined with Incoloy 825	KCS	KCS
OTHER	PWHT	PWHT	TP304 Mesh BLK	TP 304 Mesh
X-RAY, SPOT OR FULL				
LINING				
SKIRT LENGTH OR ELEV.ABOVE GRADE, ft	8'0"	8'0"	4'0"	4'0"
INSULATION REQ'D?	NO	NO	NO	NO
FOR FRACTIONATORS				
NO. OF TRAYS				
TYPE OF TRAYS				
NO. LIQUID PASSES				
TRAY MATERIAL				
FOR REACTORS				
ASME SECT VIII				
DIV. 1 OR 2				
IF NOT, GIVE DETAILS				
DOWN FLOW				
INTERNAL DISTRIBUTOR				
RADIAL FLOW				
OTHER				
REMARKS				

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VESSELS

CUSTOMER DOE

PROJECT ENGR Terry Marker

COMPLEX Resid/Coal CoProcessor

PROJECT NO 728005-4000

PROCESS UNIT Resid/Coal CoProcessor

DATE March 7, 1994

ITEM NO.	V9	V10	V11	V12
SERVICE (on drums give Horiz. or Vert.)	3rd Stage Makeup Comp Suction Drum (Vertical)	4th Stage Makeup Comp Suction Drum (Vertical)	Stripper (Vertical)	Stripper Receiver (Horizontal)
DIAMETER, ft - in	3'0"	2'6"	7'6"	3'6"
LENGTH, TAN - TAN, ft - in	8'0"	6'0"	75'0"	14'0"
DESIGN TEMP, °F	250	250	475	250
DESIGN PRESSURE, psig	1325	2310	165	150
MATERIAL OF CONST.				
CARBON STEEL	KCS	KCS	KCS lined w/ 1/8" TP316L	KCS
OTHER	TP304 Mesh Blk	TP304 Mesh Blk		PWHT
X-RAY, SPOT OR FULL				
LINING				Incoloy 325
SKIRT LENGTH OR ELEV.ABOVE GRADE, ft	4'0"	4'0"	20'0"	
INSULATION REQ'D?	NO	NO	YES	NO
FOR FRACTIONATORS				
NO. OF TRAYS			20	
TYPE OF TRAYS			Sieve	
NO. LIQUID PASSES			2	
TRAY MATERIAL			TP410	
FOR REACTORS				
ASME SECT VIII				
DIV. 1 OR 2				
IF NOT, GIVE DETAILS				
DOWN FLOW				
INTERNAL DISTRIBUTOR				
RADIAL FLOW				
OTHER				
REMARKS				

UOP

VESSELS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal CoProcessingPROJECT NO 728005-4000PROCESS UNIT Resid/Coal CoProcessingDATE March 7, 1994

ITEM NO.	V-13	V-14	V-15	V-16
SERVICE (on drums give Horiz. or Vert.)	Product Fractionator (Vertical)	Product Fractionator Receiver (Horizontal)	Light Gas Oil Stripper (Vertical)	Vacuum Column (Vertical) <i>top Mid Bot</i>
DIAMETER, ft - in	13'6"	10'	7'	<i>10' / 13' / 5'</i>
LENGTH, TAN - TAN, ft - in	90'0"	32'	23'0"	<i>88'</i>
DESIGN TEMP, °F	550/700°F	250	540	550/760
DESIGN PRESSURE, psig	50	50	50	50
MATERIAL OF CONST. CARBON STEEL OTHER X-RAY, SPOT OR FULL LINING	KCS 1/8" CA in unlined area	KCS 3/16" CA & PWHT 1/4" CA on Boot	✓	KCS 1/8" Ca in unlined areas
SKIRT LENGTH OR ELEV.ABOVE GRADE, ft	20'0"	20'0"	20'	30'
INSULATION REQ'D?	Yes	No	Yes	Yes
FOR FRACTIONATORS				
NO. OF TRAYS	27		6	Packing
TYPE OF TRAYS	Sieve		Sieve	Glitsch Grid
NO. LIQUID PASSES	4		1	
TRAY MATERIAL				
ASME SECT VIII DIV. 1 OR 2 IF NOT, GIVE DETAILS DOWN FLOW INTERNAL DISTRIBUTOR RADIAL FLOW OTHER				
REMARKS				

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VESSELS

CUSTOMER DOE

PROJECT ENGR Terry Marker

COMPLEX Resid/Coal CoProcessing

PROJECT NO 728005-4000

PROCESS UNIT Resid/Coal CoProcessing

DATE March 7, 1994

ITEM NO.	V-16	V-17	V-18	V-19
SERVICE (on drums give Horiz. or Vert.)	Condensate Receiver (Horizontal)	Absorber (Vertical)	Solvent Recovery (Vertical)	Solvent Recovery Receiver (Horizontal)
DIAMETER, ft - in	3'	9'	4'116'	3'
LENGTH, TAN - TAN, ft - in	9'	40'	150	8'
DESIGN TEMP, °F	250	250°F	603	250
DESIGN PRESSURE, psig	50	205	280	275
MATERIAL OF CONST. CARBON STEEL OTHER X-RAY, SPOT OR FULL LINING	✓	✓	✓	✓
SKIRT LENGTH OR ELEV.ABOVE GRADE, ft	25'	8'0"	8'	25'
INSULATION REQ'D?	No	No	Yes	No
FOR FRACTIONATORS NO. OF TRAYS TYPE OF TRAYS NO. LIQUID PASSES TRAY MATERIAL		Packing Raschig Ring	Packing Flexipak Upper ½, 4' diameter Lower ½, 16' diameter	
ASME SECT VIII DIV. 1 OR 2 IF NOT, GIVE DETAILS DOWN FLOW INTERNAL DISTRIBUTOR RADIAL FLOW OTHER				
REMARKS				

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EXCHANGERS

CUSTOMER DOE

PROJECT ENGR Terry Marker

COMPLEX Resid/Coal CoProcessing

PROJECT NO. 728005-4000

PROCESS UNIT Resid/Coal CoProcessing

DATE March 7, 1994

ITEM NO.		E1	E2	E3	E4
SERVICE		Hot Flash Vapor Steam Gen	Hot Flash Vapor Cooler	1st Stage Makeup Comp Cooler	2nd Stage Makeup Comp Cooler
DUTY, 10 ⁶ Btu/hr		3.2	3.2	8.9	9.4
HOT SIDE	IN °F	590	350	219	224
	OUT °F	350	120	100	100
HOT SIDE THRU SHELL OR TUBE		T	T	S	S
COLD SIDE	IN °F	300	90	80	80
	OUT °F	300	--	100	100
DESIGN PRESSURE psig	TUBE	800	790	440	890
	SHELL	535	--	660	1335
MATERIAL	TUBE	Incoloy 825	Incoloy 825	Killed C Steel	Killed C Steel
	SHELL	C. Steel	Header - Incoloy 825	Killed C. Steel	Killed C. Steel
TEMA TYPE		BKU	--	AEU	AEU
TUBE DIAMETER & PITCH		¾" x 14BWG x 1" sq	--	¾" x 14BWG x 1" sq	¾" 14BWG x 1" sq
BUNDLE LENGTH, in		20'	30'	20'	20'
MAX SHELL DIAMETER, in		--	--	--	--
TYPE SHELL & TUBE, AIR COOLED, HAIR PIN		S&T (Kettle)	Air Fin	S&T	S&T
NOTES: U, Btu/ft ² hr°F 7.71 ED6 4.54 ED6 1.40		U=90	U=3.5 Finned Basis	U=80	U=80

UOP

EXCHANGERS

CUSTOMER DOE

PROJECT ENGR Terry Marker

COMPLEX Resid/Coal CoProcessing

PROJECT NO. 728005-4000

PROCESS UNIT Resid/Coal CoProcessing

DATE March 7, 1994

ITEM NO.		E5	E6	E7	E8
SERVICE		Feed/Hot Sep Liquid	Recycle Gas/Hot Sepn Vapor	Int Separator Vapor Steam Gen	Int Sept Vapor Cooler
DUTY, 10 ⁶ Btu/hr		17.5	224	93	124
HOT SIDE	IN °F	870	870	540	350
	OUT °F	665	540	350	120
HOT SIDE THRU SHELL OR TUBE		Tube (slurry)	S	T	T
COLD SIDE	IN °F	389	140	300	90
	OUT °F	442	810	300	--
DESIGN PRESSURE psig	TUBE	3150	3360	3140	3130
	SHELL	3355	3150	2095	--
MATERIAL	TUBE	TP304	TP304	Incoloy 825	Incoloy 825
	SHELL	2 1/4 Cr-1Mo lined with 1/4" TP347 Weld Overlay	2 1/4 Cr-1Mo lined with 1/4" TP347 Weld Overlay	C. Steel	Header Incoloy 82S
TEMA TYPE		DEU	DEU	CKU	--
TUBE DIAMETER & PITCH		1" x 12BWG x 1 1/2" sq.	3/4" x 14BWG x 1" sq	3/4" x 14BWG x 1" sq	1" x 12BWG
BUNDLE LENGTH, in		20'	20'	20'	20'
MAX SHELL DIAMETER, in					
TYPE SHELL & TUBE, AIR COOLED, HAIR PIN		S&T	S&T	S&T Kettle	Air Fin
NOTES: U, Btu/ft ² hr°F		U = 25	U = 80	U = 90	U = 3.6 <u>Fin Tubes</u> Pipe Header V-tubes Ratio Fin/Bare = 23

UOP

EXCHANGERS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal CoProcessingPROJECT NO. 728005-4000PROCESS UNIT Resid/Coal CoProcessingDATE March 7, 1994

ITEM NO.		E9	E10	E11	E12
SERVICE		3rd Stage Makeup Comp Cooler	Stripper Condenser	Product Fract. Condenser	Diesel Pump-Around Steam Generator
DUTY, 10 ⁶ Btu/hr		8.0	8.5	82.4	15.0
HOT SIDE	IN °F	205	151	296	500
	OUT °F	100	100	140	433
THOT SIDE THRU SHELL OR TUBE		S	T	T	T
COLD SIDE	IN °F	80	80	90	300
	OUT °F	100	100	--	300
DESIGN PRESSURE psig	TUBE	1545	140	50	50
	SHELL	2315	95	--	75
MATERIAL	TUBE	K.C. Steel	Incoloy 825	KC Stl	CS
	SHELL	K.C. Steel	KCS	--	C Steel
TEMA TYPE		AEU	AES/T	--	AKU
TUBE DIAMETER & PITCH		¾" x 14BWG, 1"	¾" x 14BWG, 1" sq	1" x 12 BWG	¾" x 14 BWG x 1"
BUNDLE LENGTH, in		20'	20'	30'	20'
MAX SHELL DIAMETER, in					
TYPE SHELL & TUBE, AIR COOLED, HAIR PIN		S&T	S&T	Air Fin	S&T Kettles
NOTES: U, Btu/ft ² hr°F --		U = 80	U = 85	U = 3.9 (Finned)	U = 70

UOP

EXCHANGERS

CUSTOMER DOE

PROJECT ENGR Terry Marker

COMPLEX Resid/Coal CoProcessing

PROJECT NO. 728005-4000

PROCESS UNIT Resid/Coal CoProcessing

DATE March 7, 1994

ITEM NO.		E13	E14	E15	E16
SERVICE		Diesel Product Cooler	Ejector Condivision	LVGO Cooler	HVGO Cooler
DUTY, 10 ⁶ Btu/hr		26.7	9.0	22.0	15.0
HOT SIDE	IN °F	460	250	270	553
	OUT °F	150	140	150	516
HOT SIDE THRU SHELL OR TUBE		T	S	T	T
COLD SIDE	IN °F	90	80	90	300
	OUT °F	--	110	--	300
DESIGN PRESSURE psig	TUBE	90	75	50	100
	SHELL	--	50	--	75
MATERIAL	TUBE	CS		CS	TP410
	SHELL		CS		C. Steel
TEMA TYPE			AEU		AKU
TUBE DIAMETER & PITCH		1" x 12 BWG	¾" - 14 - 1" sq	1" - 12 BWG	¾" x 14BWG x 1"
BUNDLE LENGTH, in		30'	20'	30'	20'
MAX SHELL DIAMETER, in					
TYPE SHELL & TUBE, AIR COOLED, HAIR PIN		Air Fin	S&T	Air Finned	S&T (Kettle)
NOTES: U, Btu/ft ² hr°F		U = 2.6 Finned	U = 150	U = 3.0 Finned Ratio 23 = Fin/Bare	U = 60

UOP

EXCHANGERS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal CoProcessingPROJECT NO. 728005-4000PROCESS UNIT Resid/Coal CoProcessingDATE March 7, 1994

ITEM NO.		E17	E18	E19	
SERVICE		HVGO Product Cooler	Vac Btms Cooler	Recovery Condenser	
DUTY, 10 ⁶ Btu/hr		13.9	23.3	6.9	
HOT SIDE	IN °F	516	710	181	
	OUT °F	150	350	120	
HOT SIDE THRU SHELL OR TUBE		T	T	T	
COLD SIDE	IN °F	90	300	90	
	OUT °F	--	300	--	
DESIGN PRESSURE psig	TUBE	75	100	275	
	SHELL	--	75	--	
MATERIAL	TUBE	CS	TP410		
	SHELL	--	CS Lined with TP405/410S	Killed C Steel	
TEMA TYPE		--	AKU	Killed C Steel	
TUBE DIAMETER & PITCH		1" x 12 BWG	¾" x 14 BWG x 1" sq		
BUNDLE LENGTH, in		30'	20'	30'	
MAX SHELL DIAMETER, in					
TYPE SHELL & TUBE, AIR COOLED, HAIR PIN		Air Fin	S&T (Kettle)	Air Fin	
NOTES: U, Btu/ft ² hr°F		U = 2.5 Finned	U = 30	U = 3.5 Finned Basis	

UOP

FIRED HEATERS

CUSTOMER DOE

PROJECT ENGR Terry Marker

COMPLEX Resid/Coal CoProcessing

PROJECT NO. 728005-4000

PROCESS UNIT Resid/Coal CoProcessing

DATE March 7, 1994

ITEM NO.	F1	F2	F3	F4
SERVICE	Recycle Hydrogen	Product Fract.	Vacuum Column Charge	Recovery Column
ABSORPTION DUTY, 10 ⁶ Btu/hr				
PROCESS	77.5	94.4	31.4	116
STEAM				
DESIGN RADIANT FLUX RATE, Btu/ft ² .hr	15000	8000	7000	8000
DESIGN PRESSURE, psig	3465	200	200	275
MAX TUBE WALL TEMP, °F	1200	1000	1000	1000
PROCESS FLOW RATE, lb/hr	328867	476989	227714	1561421
TYPE		✓	✓	✓
UOP HORIZONTAL TUBE				
NON UOP HORIZONTAL TUBE				
WICKET				
CYLINDRICAL VERTICAL				
OTHER	UOP - Vertical Double Fired Cruciform			
CONVECTION SECTION NONE, PROCESS OR STEAM (psig)	Process	Process	Process	Process
NUMBER OF PASSES (See note below)	Four	Four	Four	Four
TYPE OF RETURN BENDS WELDED, REMOVABLE PLUGS OR MANIFOLDS	Welded	Removeable plugs	Removable plugs	Removeable plugs
TUBE MATERIAL	TP347 HSS	9 Cr-1Mo	9 Cr-1Mo	9 Cr-1Mo
FUEL TO BE FIRED	Gas	Gas	Gas	Gas
SOOT BLOWERS				
STACK (if other than 40' stub)				
HIGH INTENSITY BURNERS				
COMB. AIR PREHEAT				
NOTES:				

UOP

PUMPS AND DRIVERS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal CoProcessingPROJECT NO 728005-4000PROCESS UNIT Resid/Coal CoProcessingDATE March 7, 1994

ITEM NO.	P1	P2	P3	P4
SERVICE	Charge	Charge	Stripper OVH	Fractionation OVH
CONSTRUCTION API or ANSI	API	API	API	API
SEALS: SINGLE TANDEM DOUBLE	Single	Single	Single	Tandem
API SEAL FLUSH PLANS REQUIRED (yes/no)	No	No	No	52
TYPE OF PUMP				
RECIPROCATING		✓		
CENTRIFUGAL	✓		✓	✓
PROPORTIONING				
OTHER				
RATED FLOW, gpm	2376	150	145	933
FLOWING TEMPERATURE, °F	395	140	100	140
FLOWING SP. GR.	.60	.98	.70	.77
SUCTION PRESSURE, psig	35	0	115	10
DISCHARGE PRESSURE, psig	3300	2985	165	75
RATED HEAD, ft	9940	77	165	195
CASING MATERIAL	Steel*	Steel	Steel	Steel
IMPELLER MATERIAL	11-13% Cr *	SS PLNGRS	C.Iron	C. Iron
TYPE DRIVER (MOTOR, TURBINE, etc.)	Motor	Motor	Motor	Motor
SPARE	✓	✓	✓	✓
TYPE DRIVER (MOTOR, TURBINE, etc.)	Motor	Motor	Motor	Motor
REMARKS	33 wt% solids 200 mesh or smaller		w/3 mole % H ₂ S	

* Hard Coated

UOP

PUMPS AND DRIVERS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal CoProcessingPROJECT NO 728005-4000PROCESS UNIT Resid/Coal CoProcessingDATE March 7, 1994

ITEM NO.	P5	P6	P7	P8
SERVICE	Diesel Pump Around	Net Diesel	Product Fractionator Btms	LUGO
API CONSTRUCTION API or ANSI	API	API	API	API
SEALS: SINGLE TANDEM DOUBLE	Single	Single	Single	Single
API SEAL FLUSH PLANS REQUIRED (yes/no)	No	No	No	No
TYPE OF PUMP				
RECIPROCATING				
CENTRIFUGAL	✓	✓	✓	✓
PROPORTIONING				
OTHER				
RATED FLOW, gpm	1590	388	494	888
FLOWING TEMPERATURE, °F	464	416	641	280
FLOWING SP. GR.	.83	.76	.88	.88
SUCTION PRESSURE, psig	17	18	20	0
DISCHARGE PRESSURE, psig	60	75	125	75
RATED HEAD, ft	120	173	275	197
CASING MATERIAL	Steel	Steel	11-13% Cr	Steel
IMPELLER MATERIAL	Steel	Steel	11-13% Cr	C. Iron
TYPE DRIVER (MOTOR, TURBINE, etc.)	Motor	Motor	Motor	Motor
SPARE	✓	✓	✓	✓
TYPE DRIVER (MOTOR, TURBINE, etc.)	Motor	Motor	Motor	Motor
REMARKS	1.5 wt% sulfur	1.5 wt% sulfur	18% solids 2.5 wt% S	1.8 wt% S

UOP

PUMPS AND DRIVERS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal CoProcessingPROJECT NO 728005-4000PROCESS UNIT Resid/Coal CoProcessingDATE March 7, 1994

ITEM NO.	P9	P10	P11	P12
SERVICE	HVGO	Slop Oil	Vacuum Btms	Condensate
CONSTRUCTION API or ANSI	API	API	API	API
SEALS: SINGLE TANDEM DOUBLE	Single	Single	Single	Single
API SEAL FLUSH PLANS REQUIRED (yes/no)	No	No	No	No
TYPE OF PUMP				
RECIPROCATING				
CENTRIFUGAL	✓	✓	✓	✓
PROPORTIONING				
OTHER				
RATED FLOW, gpm	1600	110	334	18
FLOWING TEMPERATURE, °F	572	697	725	140
FLOWING SP. GR.	.83	.83	1.0	.98
SUCTION PRESSURE, psig	0	0	0	0
DISCHARGE PRESSURE, psig	75	75	75	75
RATED HEAD, ft	210	210	175	177
CASING MATERIAL	11-13% CR	11-13% CR	11-13% CR*	Steel
IMPELLER MATERIAL	11-13% CR	11-13% CR	11-13% CR*	C. Iron
TYPE DRIVER (MOTOR, TURBINE, etc.)	Motor	Motor	Motor	Motor
SPARE	✓	✓	✓	✓
TYPE DRIVER (MOTOR, TURBINE, etc.)	Motor	Motor	Motor	Motor
REMARKS	2 wt% Sulfur	3 wt% Sulfur	38 wt% Solids 3 wt% Sulfur • Hard Coated	

UOP

PUMPS AND DRIVERS

CUSTOMER DOEPROJECT ENGR Terry MarkerCOMPLEX Resid/Coal CoProcessingPROJECT NO 728005-4000PROCESS UNIT Resid/Coal CoProcessingDATE March 7, 1994

ITEM NO.	P13			
SERVICE	Solvent Rec			
CONSTRUCTION API or ANSI	API			
SEALS: SINGLE TANDEM DOUBLE	Tandem			
API SEAL FLUSH PLANS REQUIRED (yes/no)	52			
TYPE OF PUMP				
RECIPROCATING				
CENTRIFUGAL	✓			
PROPORTIONING				
OTHER				
RATED FLOW, gpm	1128			
FLOWING TEMPERATURE, °F	120			
FLOWING SP. GR.	.44			
SUCTION PRESSURE, psig	195			
DISCHARGE PRESSURE, psig	275			
RATED HEAD, ft	182			
CASING MATERIAL	Steel			
IMPELLER MATERIAL	C. Iron			
TYPE DRIVER (MOTOR, TURBINE, etc.)	Motor			
SPARE	✓			
TYPE DRIVER (MOTOR, TURBINE, etc.)	Motor			
REMARKS				

RECIPROCATING COMPRESSORS

Engineering Data Required for UOP Cost Estimating Department

PROJECT ENGINEER

ECN

PROJECT NUMBER

25956

CUSTOMER DOE	UNIT CO-PROCESSING	DATE 1/5/88
ITEM NO. SERVICE	MAKE-UP HYDROGEN 1ST STAGE	→ 2ND STAGE
*CAPACITY - RATED (EACH MACHINE) MMSCFD OR SCFM, MOLES/HR., NM ³ /HR., LBS./HR.	97.8 MMSCFD	97.8 MMSCFD
*CAPACITY - ACFM AT SUCTION CONDITION	3439	1789
NUMBER OF STAGES	1	1
*LUBE OR NONLUBE	Lube	→
PISTON TRAVEL, FPM	700 MAX	→
*SUCTION TEMPERATURE, °F	100	→
*SUCTION PRESSURE, PSIA	315	610
*DISCHARGE PRESSURE, PSIA	615	1225
*MOLECULAR WEIGHT	2.03	→
MOLECULAR % HYDROGEN	99.9	→
CORROSIVE - YES OR NO DUE TO H ₂ O, H ₂ S, CL - LIST OTHERS	NO	→
*"K" VALUE OF GAS	1.4	→
ESTIMATED RATED BHP	3640 2500	3851
COMPRESSOR RPM	300 MAX	→
GEAR - YES OR NO	NO	→
*DRIVER TYPE MOTOR NONCONDENSING STEAM TURBINE CONDENSING STEAM TURBINE LIST OTHER TYPES	MOTOR	MOTOR
POWER AND STEAM CONDITIONS DRIVER RPM VOLTS, FREQUENCY INDUCTION OR SYNCHRONOUS STEAM PRESSURE, PSIG NONCONDENSING EXHAUST PRESSURE, PSIG CONDENSING EXHAUST PRESSURE, PSIA		
REMARKS	ASSUME 50% EFFICIENCY, 11.5% HP 4 stage - THREE MACHINES	
*REQUIRED INFORMATION		

RECIPROCATING COMPRESSORS

Engineering Data Required for UOP Cost Estimating Department

PROJECT ENGINEER

ECN

PROJECT NUMBER

25956

CUSTOMER DOE	UNIT Co-Processing	DATE 1/5/88
ITEM NO. SERVICE	MAKE-UP Hydrogen 3RD STAGE	MAKE-UP Hydrogen 4TH STAGE
*CAPACITY - RATED (EACH MACHINE) MMSCFD OR SCFM, MOLES/HR., NM ³ /HR., LBS./HR.	97.8 MMSCFD	→
*CAPACITY - ACFM AT SUCTION CONDITION	909	515
NUMBER OF STAGES	1	→
*LUBE OR NONLUBE	Lube	→
PISTON TRAVEL, FPM	700 MAX	→
*SUCTION TEMPERATURE, °F	100	→
*SUCTION PRESSURE, PSIA	1220	2215
*DISCHARGE PRESSURE, PSIA	2220	3215
*MOLECULAR WEIGHT	2.03	→
MOLECULAR % HYDROGEN	99.9	→
CORROSIVE - YES OR NO DUE TO H ₂ O, H ₂ S, CL - LIST OTHERS	NO	→
*"k" VALUE OF GAS	1.40	→
ESTIMATED RATED BHP	3326	2060
COMPRESSOR RPM	300 MAX	→
GEAR - YES OR NO	NO	→
*DRIVER TYPE MOTOR NONCONDENSING STEAM TURBINE CONDENSING STEAM TURBINE LIST OTHER TYPES	MOTOR	→
POWER AND STEAM CONDITIONS DRIVER RPM VOLTS, FREQUENCY INDUCTION OR SYNCHRONOUS STEAM PRESSURE, PSIG NONCONDENSING EXHAUST PRESSURE, PSIG CONDENSING EXHAUST PRESSURE, PSIA		
REMARKS		
*REQUIRED INFORMATION		

CENTRIFUGAL COMPRESSORS

Engineering Data Required for UOP Cost Estimating Department

PROJECT ENGINEER

ECN

PROJECT NUMBER

25956

CUSTOMER <i>DOE</i>	UNIT <i>CO-Processing</i>	DATE <i>1/23/88</i>
ITEM NO. SERVICE	<i>Recycle</i>	
*CAPACITY - RATED (EACH MACHINE) MMSCFD OR SCFM, MILES/HR., NM ³ /HR., LBS./HR.	<i>357 MMSCFD</i>	
*CAPACITY - ACFM AT SUCTION CONDITION	<i>1427</i>	
*SUCTION TEMPERATURE, °F	<i>120</i>	
*SUCTION PRESSURE, PSIA	<i>2995</i>	
*DISCHARGE PRESSURE, PSIA	<i>3215</i>	
*MOLECULAR WEIGHT	<i>8.4</i>	
MOLECULAR % HYDROGEN	<i>76 %</i>	
*CORROSIVE - YES OR NO DUE TO H ₂ O, H ₂ S, CL - LIST OTHERS	<i>H₂S 9.5 % NH₃ 0.2 % H₂O 0.1 %</i>	
*"K" VALUE OF GAS	<i>1.31</i>	
*ESTIMATED RATED BHP	<i>1352 1740</i>	
COMPRESSOR RPM	<i>11000</i>	
GEAR - YES OR NO	<i>NO</i>	
*DRIVER TYPE MOTOR NONCONDENSING STEAM TURBINE CONDENSING STEAM TURBINE LIST OTHER TYPES	<i>COND. STEAM TURBINE</i>	
POWER AND STEAM CONDITIONS DRIVER RPM VOLTS, FREQUENCY INDUCTION OR SYNCHRONOUS STEAM PRESSURE, PSIG NONCONDENSING EXHAUST PRESSURE, PSIG CONDENSING EXHAUST PRESSURE, PSIA	<i>600 PSIG, 750°F HP STEAM</i>	
REMARKS	<i>- SINGLE STAGE BARREL - OIL FILM SEALS - SEPARATE LUBE AND SEAL OIL SYSTEMS</i>	
*REQUIRED INFORMATION		

Appendix C

Equipment Costs

uop *Memorandum*

Date: April 20, 1994
To: Terry Marker
From: Cost Engineering
Subject:

RESID/COAL COPROCESSING ERECTED COST (E1300)
DOE
RE: YOUR REQUEST DATED MARCH 18, 1994
728005-4000

We are noting estimated selected equipment cost and estimated erected cost for the equipment listed. The equipment costs were based on EFCEST type data provided. These costs represent U.S.A. Gulf Coast erection to UOP standards for new equipment on a First Quarter 1994, open shop (non-union) labor basis.

The equipment cost has an accuracy of $\pm 30\%$. The estimated erected cost is a factored cost which includes associated bulk items, installation, indirects, contractors' home office expenses and profit. This erected cost is Class E ($\pm 50\%$). The estimated erected cost is for battery limits only.

FIRED HEATERS	COST
F1 Recycle Hydrogen (Cruciform Type w/ TP 347H SS Tubes)	\$2,302,000
F2 Product Fractionator (Cabin Type w/9 Cr- 1 Mo Tubes)	1,783,200
F3 Vacuum Column Charge (Cabin Type w/9 Cr- 1 Mo Tubes)	700,800
F4 Recovery Column (Cabin Type w/9 Cr- 1 Mo Tubes)	2,154,000
TOTAL FIRED HEATERS	\$6,940,000

Date: April 20, 1994

To: Terry Marker

Subject: Resid/Coal Coprocessing Erected Cost (E1300)

Page 2 of 5

VESSELS & INTERNALS	COST
V1 A,B,C,D,E&F Reactors (Six - 8' x 45' 3 Cr- 1Mo w/Two Layers of TP 347 Weld Overlay)	\$17,400,000
V2 Hot Separator (6'x 12' 3 Cr- 1Mo w/Two Layers of TP 347 Weld Overlay)	760,000
V3 Hot Flash (8'x 16' 1 1/4 Cr- 1/2 Mo w/TP 321 Weld Overlay)	253,400
V4 Intermediate Separator (7'x 20' 1 1/4 Cr- 1/2 Mo w/TP 347 Weld Overlay)	870,000
V5 Cold Separator (9'x 30' KCS w/Incoloy 825 Weld Overlay)	1,510,000
V6 Cold Flash (9'x 28' KCS w/Incoloy 825 lining)	225,900
V7 1st Stage Makeup Comp-Suction Drum (4'x 8' KCS)	22,700
V8 2nd Stage Makeup Comp-Suction Drum (3'- 6" x 8' KCS)	24,400
V9 3rd Stage Makeup Comp-Suction Drum (3'- 6" x 8' KCS)	27,300
V10 4th Stage Makeup Comp-Suction Drum (2'- 6" x 6' KCS)	38,600
V11 Stripper (7'- 6" x 75' KCS w/TP 316L Lining & w/20 TP 410 Sieve Trays)	259,800
V12 Stripper Receiver (3'- 6" x 14' KCS w/Incoloy 825 lining)	39,600
V13 Product Fractionator (13'- 6" x 90' KCS w/TP 410 Lining in Bottom Section & w/27 TP 410 Sieve Trays)	443,700
V14 Product Fractionator Receiver (10' x 32' KCS)	46,300
V15 Light Gas Oil Stripper (7'x 23' KCS w/TP 410S Lining & w/6 TP 410 Sieve Trays)	94,000
V16 Vacuum Column (10' & 13' & 5' x 87' KCS w/TP 410S Lining in Bottom Section & w/Grid & Pall Type Packing)	410,000
V17 Absorber (9'x 40' KCS w/Raschig Ring Packing)	303,100
V18 Solvent Recovery (4' & 16' x 140' KCS w/Flexipak Packing)	1,155,100
V19 Solvent Recovery Receiver (3'x 8' KCS)	18,900
V20 Condensate Receiver (3'x 9' KCS)	18,200
TOTAL VESSELS & INTERNALS	\$23,921,000

Date: April 20, 1994
 To: Terry Marker
 Subject: Resid/Coal Coprocessing Erected Cost (E1300)
 Page 3 of 5

HEAT EXCHANGERS	COST
E1 Hot Flash Vapor Steam Generator (Kettle Type Unit w/CS Shell & Incoloy 825 Tubes)	\$48,600
E2 Hot Flash Vapor Cooler (AC Unit w/Incoloy 825 Tubes)	137,100
E3 1st Stage Makeup Compressor Cooler (KCS S&T Type Unit)	44,600
E4 2nd Stage Makeup Compressor Cooler (KCS S&T Type Unit)	62,800
E5 Feed/Hot Separator Liquid (S&T Type Unit w/2 1/4 Cr-1Mo Shell Including TP 347 Weld Overlay & TP 304 Tubes)	231,700
E6 Recycle Gas/Hot Separator Vapor (Three S&T Type Units w/2 1/4 Cr-1Mo Shells Including TP 347 Weld Overlay & TP 304 Tubes)	1,109,500
E7 Intermediate Separator Vapor Steam Generator (Kettle Type Unit w/CS Shell & Incoloy 825 Tubes)	591,000
E8 Intermediate Separator Vapor Cooler (AC Unit w/Incoloy 825 Tubes)	5,003,000
E9 3rd Stage Makeup Compressor Cooler (KCS S&T Type Unit)	84,600
E10 Stripper Condenser (S&T Type Unit w/KCS Shell & Incoloy 825 Tubes)	207,100
E11 Product Fractionator Condenser (AC Unit w/KCS Tubes)	340,500
E12 Diesel Pump-around Steam Generator (CS Kettle Type Unit)	28,200
E13 Diesel Product Cooler (AC Unit w/CS Tubes)	102,500
E14 Ejector Condensor (CS S&T Type Unit)	15,900
E15 LVGO Cooler (AC Unit w/CS Tubes)	117,500
E16 HVGO Cooler (Kettle Type Unit w/CS Shell TP 410 Tubes)	44,900
E17 HVGO Product Cooler (AC Unit w/CS Tubes)	114,000
E18 Vacuum Bottoms Cooler (Kettle Type Unit w/CS Shell Including TP 410 Lining & TP 410 Tubes)	118,100
E19 Recovery Condenser (AC Unit w/CS Tubes)	58,400
TOTAL HEAT EXCHANGERS	\$8,460,000

Date: April 20, 1994

To: Terry Marker

Subject: Resid/Coal Coprocessing Erected Cost (E1300)

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PUMPS & DRIVERS	COST
P1A&B Charge (Four 2,000 HP Motor Driven Reciprocating Type)	\$2,590,000
P2A&B Water Injection (Two 400 HP Motor Driven Reciprocating Type)	276,500
P3A&B Stripper Overhead (Two 10 HP Motor Driven Centrifugal Type)	24,900
P4A&B Fractionator Overhead (Two 60 HP Motor Driven Centrifugal Type)	34,500
P5A&B Diesel Pump Around (Two 60 HP Motor Driven Centrifugal Type)	36,000
P6A&B Net Diesel (Two 25 HP Motor Driven Centrifugal Type)	27,900
P7A&B Product Fractionator Bottoms (Two 60 HP Motor Driven Centrifugal Type)	42,200
P8A&B LVGO (Two 60 HP Motor Driven Centrifugal Type)	29,700
P9A&B HVGO (Two 125 HP Motor Driven Centrifugal Type)	69,800
P10A&B Slop Oil (Two 15 HP Motor Driven Centrifugal Type)	31,500
P11A&B Vacuum Bottoms (Two 30 HP Motor Driven Centrifugal Type)	39,000
P12A&B Condensate (Two 7 1/2 HP Motor Driven Centrifugal Type)	24,800
P13A&B Solvent Recovery (Two 40 HP Motor Driven Centrifugal Type)	35,200
TOTAL PUMPS & DRIVERS	\$3,268,000

COMPRESSORS & DRIVERS	COST
C1A,B&C Makeup Hydrogen (Three 50% - 8,000 HP Motor Driven Reciprocating Type)	\$9,060,000
C2 Recycle (2,000 HP Motor Driven Centrifugal Type)	1,516,000
TOTAL COMPRESSORS & DRIVERS	\$10,576,000

Date: April 20, 1994
To: Terry Marker
Subject: Resid/Coal Coprocessing Erected Cost (E1300)
Page 5 of 5

EQUIPMENT COST SUMMARY	COST
FIRED HEATERS	\$6,940,000
VESSELS & INTERNALS	\$23,921,000
HEAT EXCHANGERS	8,460,000
PUMPS & DRIVERS	3,268,000
COMPRESSORS & DRIVERS	10,576,000
TOTAL EQUIPMENT COST (USD)	\$53,159,000
TOTAL ESTIMATED ERECTED EQUIPMENT COST	\$100,000,000

Boris Hitalenko
UOP Cost Engineering

Appendix D

Refinery LP Model Results

Appendix D.1
Conventional Refinery Base Case

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY

OVERALL ECONOMIC SUMMARY

TIME 1 BASE MODEL

	DOLLARS PER DAY	\$MM PER YEAR
<u>GROSS REVENUES AT MARKET</u>		
CRACKING REFINERY	2027740.30	740.13
TOTAL...	2027740.30	740.13

RAW MATERIAL PURCHASES

CRACKING REFINERY	1170787.69	427.34
TOTAL...	1170787.69	427.34

PROCESS UNIT OPERATING COSTS

CRACKING REFINERY	82492.66	30.11
TOTAL...	82492.66	30.11

UTILITY POOL COSTS

CRACKING REFINERY	14974.03	5.47
TOTAL...	14974.03	5.47

TRANSPORTATION COSTS

CRACKING REFINERY	9.04	0.00
TOTAL...	9.04	0.00

CAPITAL CHARGES

CRACKING REFINERY	532018.73	194.19
TOTAL...	532018.73	194.19

<u>OPTIMIZED PROFIT</u>	227458.15	83.02
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY

MARKET CENTER SUMMARY

PRODUCT SALESBASE MODEL

PRODUCT SALES

		MAXIMUM	MINIMUM	QUANTITY SOLD	PRICE \$	TOTAL \$ PER DAY
LIQUID PROPANE GAS	BBLS	.	.	2385.01	13.50	32197.63
UNLEADED PRE. GASOLINE	BBLS	.	.	16464.07	24.76	407650.47
UNLEADED REG. GASOLINE	BBLS	.	.	32928.15	23.33	768213.76
JET FUEL	BBLS	.	.	8537.38	22.26	190041.98
.05% S NO.2 FUEL	BBLS	.	.	24502.57	21.22	519944.51
1.0% S NO.6 FUEL	BBLS	.	.	4110.86	13.49	55455.46
COKE	TONS	.	.	1328.50	30.00	39854.99
SULFUR	TONS	.	.	143.82	100.00	14381.52

TOTAL SALES REVENUE2027740.30

MARKET RESTRICTIONS:

SALES	PRODUCT SALES	FIXED	UNLEADED REG TO UNLEADED PRM
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

REFINERY ECONOMIC SUMMARY

	DOLLARS PER DAY	\$MM PER YEAR
SALES REVENUE	2027731.26	740.1219
COST OF RAW MATERIALS	1170787.69	427.3375
PROCESS UNIT OPERATING COST	82492.66	30.1098
UTILITY POOL COSTS	14974.03	5.4655
CAPITAL CHARGES	532018.73	194.1868
OPTIMIZED PROFIT	227458.15	83.0222

PROFITABILITY ANALYSIS

GROSS PROFIT	759476.88	277.2091
DEPRECIATION	149654.00	54.6237

PROFIT BEFORE TAXES		222.5854
ALLOWANCE FOR FEDRL INCOME TAX		75.6790
NET PROFIT AFTER TAXES		146.9063

<u>CASH FLOW</u>	201.5300
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INVESTMENT

	\$MM
PLANT INVESTMENT	681.3573
OFFSITE INVESTMENT	.
TOTAL INVESTMENT	681.3573

<u>SIMPLE PAYOUT TIME (IN YEARS)</u>	3.4
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

GROSS MARGIN SUMMARY

MARKET	PRODUCTS		MAXIMUM	MINIMUM	QUANTITY SOLD	PRICE	TOTAL \$ PER DAY
SALES	LIQUID PROPANE GAS	BBLS	.	.	2385.01	13.50	32197.39
SALES	UNLEADED PRE. GASOLINE	BBLS	.	.	16464.07	24.76	407648.82
SALES	UNLEADED REG. GASOLINE	BBLS	.	.	32928.15	23.33	768210.46
SALES	JET FUEL	BBLS	.	.	8537.38	22.26	190041.12
SALES	.05% S NO.2 FUEL	BBLS	.	.	24502.57	21.22	519942.06
SALES	1.0% S NO.6 FUEL	BBLS	.	.	4110.86	13.49	55455.05
SALES	COKE	TONS	.	.	1328.50	30.00	39854.85
SALES	SULFUR	TONS	.	.	143.82	100.00	14381.50
<u>TOTAL SALES REVENUE</u>							2027731.26
<u>RAW MATERIAL PURCHASED</u>							
			MAXIMUM	MINIMUM	QUANTITY PURCHASE	PRICE	TOTAL \$ PER DAY
D-5	IPL CRUDE	BBLS	47500.00	47500.00	47500.00	13.47	639825.00
	LLOYDMINISTER CRUDE	BBLS	47500.00	47500.00	47500.00	10.48	497800.00
	ISOBUTANE	BBLS	.	.	909.58	16.81	15290.01
	NORMAL BUTANE	BBLS	.	.	1159.81	15.41	17872.68
<u>TOTAL RAW MATERIAL COST</u>							1170787.69
<u>GROSS MARGIN</u>							856943.57

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

OPERATING COST SUMMARY

PROCESS UNIT BY COMBINED MODE

	ACTIVITY TONS/DAY	OPERATING COST/DAY
CRUDE UNIT	13247.71	7471.71
VACUUM UNIT	7505.35	4225.51
DELAYED COKER	4630.54	11576.34
KEROSENE HYDROTREATER	1106.28	2780.09
DIESEL HYDROTREATER	3343.97	5601.15
GAS OIL HYDROTREATER	1695.01	1703.48
FLUID CATALYTIC CRACKER	3683.03	14069.16
CATALYTIC CONDENSATION	207.37	2198.17
ALKYLATION UNIT	514.16	3451.05
NAPHTHA HYDROTREATER	2805.19	3534.54
C5/C6 PENEX UNIT	221.02	1616.74
PLATFORMING UNIT	2581.93	4518.39
AMINE TR- SULFUR PLANT	145.27	6737.09
OFFSITE ALLOWANCE	13247.71	13009.25

TOTAL PROCESS UNIT OPERATING COST

82492.66

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY PURCHASES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
POWER CONSUMED	KWH		357011.88	0.04	14280.48
<u>TOTAL PURCHASED UTILITY COST</u>					14280.48

UTILITY SALES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
LOW PRESSURESTEAM	M LBS		361.47	0.00	0.04
<u>TOTAL UTILITY SALES CREDIT</u>					0.04

UNPOOLED UTILITIES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
CATALYSTCONSUMED	DOLLARS		5457.52	1.00	5457.52
LABOR MAINTNCEINSURNCEDOLLARS			77035.14	1.00	77035.14

TOTAL UNPOOLED UTILITY COST 82492.66

TOTAL UTILITY CONVERSION COST 693.59

TOTAL OPERATING COST 97466.69

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

MATERIAL BALANCE SUMMARY

RAW MATERIALS

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE	0.8265	47500.0	48.93	6234.62	46.40	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	48.93	7013.09	52.20	2559779.2
ISOBUTANE	0.5626	909.6	0.94	81.27	0.61	29662.4
NORMAL BUTANE	0.5835	1159.8	1.20	107.47	0.80	39227.8
TOTAL CHARGE	.	97069.4	100.00	13436.45	100.00	4904304.0

REFINERY PRODUCTS

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
LIQUID PROPANE GAS	0.5077	2385.0	2.46	192.30	1.43	70188.1
UNLEADED PRE. GASOLINE	0.7510	16464.1	16.96	1963.59	14.61	716712.0
UNLEADED REG. GASOLINE	0.7436	32928.2	33.92	3888.74	28.94	1419389.9
JET FUEL	0.8149	8537.4	8.80	1104.85	8.22	403268.9
.05% S NO.2 FUEL	0.8491	24502.6	25.24	3303.99	24.59	1205957.3
1.0% S NO.6 FUEL	0.9644	4110.9	4.24	629.59	4.69	229801.2
COKE	.	.	.	1328.50	9.89	484902.3
SULFUR	.	.	.	143.82	1.07	52492.5
TOTAL PRODUCTS	.	88928.0	91.61	12555.38	93.44	4582712.2

STREAMS CONVERTED TO UTILITIES

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
UNSATURATE REFINERY GAS	.	.	.	332.16	2.47	121236.6
SATURATE REFINERY GAS	.	.	.	61.83	0.46	22566.2
HYDROGEN	.	.	.	38.17	0.28	13931.4
REFINERY FUEL OIL	.	.	.	239.48	1.78	87411.1
TOTAL STREAMS CONVERTED	.	.	.	671.63	5.00	245145.3

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

MATERIAL BALANCE SUMMARY CONTINUED

STREAMS NOT UTILIZED

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
FCC COKE	.	.	.	207.35	1.54	75684.4
SOX EMISSIONS	.	.	.	1.57	0.01	574.0
LLOYDMINISTER CRUDE	0.9297	0.0	0.00	0.00	0.00	0.6
TAR FROM HF ALKY UNIT	.	.	.	0.51	0.00	187.7
TOTAL NOT UTILIZED	.	0.0	0.00	209.44	1.56	76446.6
<u>TOTAL PRODUCTS MADE</u>		88928.0	91.613	13436.45	100.000	4904304.0

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

PLANT CAPACITY SUMMARY

PROCESS UNIT	UNITS OF CAPACITY	MAXIMUM	MINIMUM	CALENDAR DAY BASIS	STREAM DAY BASIS	INVSTMNT
CRUDE UNIT	BBLS CAPACITY	100000.0	.	95000.0	100000.0	40700000.
VACUUM UNIT	BBLS CAPACITY	.	.	48570.1	51126.4	14322398.
DELAYED COKER	BBLS CAPACITY	.	.	28689.3	30199.3	63291540.
KEROSENE HYDROTREATER	BBLS CAPACITY	.	.	8511.3	8959.3	6214472.
DIESEL HYDROTREATER	BBLS CAPACITY	.	.	24203.6	25477.5	17818083.
GAS OIL HYDROTREATER	BBLS CAPACITY	.	.	11275.0	11868.4	13349110.
FLUID CATALYTIC CRACKER	BBLS CAPACITY	.	.	25650.0	27000.0	62836927.
CATALYTIC CONDENSATION	BBLS CAPACITY	.	.	2571.4	2706.8	7218950.
ALKYLATION UNIT	BBLS CAPACITY	.	.	4601.5	4843.7	22557501.
NAPHTHA HYDROTREATER	BBLS CAPACITY	.	.	24045.5	25311.1	7093180.
C5/C6 PENEX UNIT	BBLS CAPACITY	.	.	2128.0	2240.0	5524031.
PLATFORMING UNIT	BBLS CAPACITY	.	.	22062.9	23224.2	34034621.
AMINE TR- SULFUR PLANT	TONS CAPACITY	.	.	145.3	152.9	20994445.
OFFSITE ALLOWANCE	BBLS CAPACITY	.	.	95000.0	95000.0	365402064.
<u>TOTAL INVESTMENT</u>						681357322.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

PRODUCT BLENDING

LIQUID PROPANE GAS

	BARRELS PER DAY	TONS PER DAY	SG 60/60
PROPYLENE	95.2	7.67	0.5077
PROPANE	2289.8	184.62	0.5077
BLEND	2385.0	192.30	0.5077
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

UNLEADED PRE. GASOLINE

	BARRELS PER DAY	TONS PER DAY	SG 60/60	RESEARCH OCTANE NUMBER	MOTOR OCTANE NUMBER	VAPOR PRESSURE INDEX
ALKYLATE FROM BUTYLENE	4017.8	448.94	0.7036	94.50	92.50	40.10
FCC GASOLINE	3385.5	406.89	0.7568	91.90	80.60	87.50
NORMAL BUTANE	1465.1	135.76	0.5835	93.60	90.10	1120.00
98 RON PLATFORMATE	7595.7	972.01	0.8058	98.00	87.00	25.05
BLEND	16464.1	1963.59	0.7510	95.50	87.30	139.00
MAXIMUM SPECIFICATION	.	.	.	.	.	139.00
MINIMUM SPECIFICATION	.	.	.	95.50	86.50	.

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

PRODUCT BLENDING

UNLEADED REG. GASOLINE

	BARRELS PER DAY	TONS PER DAY	SG 60/60	RESEARCH OCTANE NUMBER	MOTOR OCTANE NUMBER	VAPOR PRESSURE INDEX
ALKYLATE FROM BUTYLENE	583.7	65.23	0.7036	94.50	92.50	40.10
LT ST RUN IPL	900.1	93.52	0.6542	60.30	58.80	159.00
LT ST RUN LLOYD	2531.4	262.99	0.6542	70.00	68.00	160.00
FCC GASOLINE	11752.8	1412.52	0.7568	91.90	80.60	87.50
NORMAL BUTANE	1994.2	184.80	0.5835	93.60	90.10	1120.00
C5/C6 PENEXATE	2074.2	212.13	0.6440	88.40	86.00	200.00
MOTOR FUEL POLYMER	1699.3	199.70	0.7400	94.00	83.00	25.50
98 RON PLATFORMATE	11392.3	1457.85	0.8058	98.00	87.00	25.05

BLEND	32928.2	3888.74	0.7436	91.50	82.50	139.00
MAXIMUM SPECIFICATION	.	.	.	.	.	139.00
MINIMUM SPECIFICATION	.	.	.	91.50	82.50	.

JET FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	SMOKE POINT
HTD KEROSENE - HIGH SP	8537.4	1104.85	0.8149	0.01	21.00
BLEND	8537.4	1104.85	0.8149	0.01	21.00
MAXIMUM SPECIFICATION	.	.	.	0.10	.
MINIMUM SPECIFICATION	.	.	.	.	21.00

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

PRODUCT BLENDING

.05% S NO.2 FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	CETANE INDEX
HYDROTREATED SR. DIESEL	13930.7	1890.86	0.8547	0.05	51.00
HYDROTREATED COKER DIESEL	10571.9	1413.14	0.8417	0.05	46.00
BLEND	24502.6	3303.99	0.8491	0.05	48.84
MAXIMUM SPECIFICATION	.	.	.	0.05	.
MINIMUM SPECIFICATION	.	.	.	.	40.00

1.0% S NO.6 FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	VISCOSTY INDEX @ 122 F
LIGHT CYCLE OIL	3576.4	540.30	0.9513	0.35	15.40
SLURRY OIL	534.5	89.29	1.0520	1.15	32.00
BLEND	4110.9	629.59	0.9644	0.46	17.75
MAXIMUM SPECIFICATION	.	.	0.9850	1.00	37.50
MINIMUM SPECIFICATION	.	.	.	.	.

COKE

	BARRELS PER DAY	TONS PER DAY	SG 60/60
HIGH SULFUR COKE	.	1328.50	.
BLEND	.	1328.50	.
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

 PRODUCT BLENDING

SULFUR

	BARRELS PER DAY	TONS PER DAY	SG 60/60
SULFUR	.	143.82	.
BLEND	.	143.82	.
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

 1.5% S REFINERY FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	VISCOSTY INDEX @ 122 F
SLURRY OIL	1434.2	239.60	1.0520	1.15	32.00
BLEND	1434.2	239.60	1.0520	1.15	32.00
MAXIMUM SPECIFICATION	.	.	.	1.50	45.00
MINIMUM SPECIFICATION	.	.	.	.	.

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY BLENDING SUMMARY

MM BTU FUEL CONSUMED

	TONS PER DAY	UTILITY UNITS/D	CONVERSN COST \$/D
UNSATURATE REFINERY GAS	332.16	14282.67	.
SATURATE REFINERY GAS	61.83	2720.31	.
HYDROGEN	38.17	5114.54	.
REFINERY FUEL OIL	239.48	9100.33	.
TOTAL...	671.63	31217.85	.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY CONVERSION SUMMARY

M LBS HIGH PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
FUEL CONSUMED	4334.92	3684.68	693.59
TOTAL...	4334.92	3684.68	693.59

M LBS MEDIUM PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
HIGH PRESSURESTEAM	1958.07	1958.07	.
TOTAL...	1958.07	1958.07	.

M GAL COOLING WATER

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
POWER CONSUMED	17463.65	22353.48	.
TOTAL...	17463.65	22353.48	.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

CRUDE UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE		0.8265	47500.0	50.00	6234.62	47.06	2275634.6
LLOYDMINISTER CRUDE		0.9297	47500.0	50.00	7013.09	52.94	2559778.6
TOTAL...		.	95000.0	100.00	13247.71	100.00	4835413.3

PRODUCTS

SATURATE REFINERY GAS		.	.	.	3.12	0.02	1137.8
PROPANE		0.5077	409.8	0.43	33.04	0.25	12060.9
ISO BUTANE		0.5626	263.4	0.28	23.54	0.18	8590.3
NORMAL BUTANE		0.5835	1130.3	1.19	104.74	0.79	38229.9
LT ST RUN	IPL	0.6542	2928.5	3.08	304.25	2.30	111051.0
LT ST RUN	LLOYD	0.6542	2531.4	2.66	262.99	1.99	95991.7
NAPHTHA	IPL	0.7389	10238.4	10.78	1201.41	9.07	438514.8
NAPHTHA	LLOYD	0.7459	6814.5	7.17	807.21	6.09	294630.5
KEROSENE	IPL	0.8109	5519.2	5.81	710.75	5.37	259422.3
KEROSENE	LLOYD	0.8324	2992.2	3.15	395.54	2.99	144371.5
DIESEL	IPL	0.8468	8929.2	9.40	1200.79	9.06	438287.2
DIESEL	LLOYD	0.8899	4917.8	5.18	695.00	5.25	253674.1
REDUCED CRUDE	IPL	0.9144	18414.3	19.38	2674.03	20.18	976019.7
REDUCED CRUDE	LLOYD	1.0086	30162.9	31.75	4831.32	36.47	1763431.5
TOTAL...		.	95251.9	100.27	13247.71	100.00	4835413.3

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

VACUUM UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
REDUCED CRUDE	IPL	0.9144	18414.3	37.91	2674.03	35.63	976019.7
REDUCED CRUDE	LLOYD	1.0086	30162.9	62.09	4831.32	64.37	1763431.5
TOTAL...		.	48577.3	100.00	7505.35	100.00	2739451.2

PRODUCTS

VACUUM GAS OIL	IPL	0.8996	14187.7	29.21	2026.91	27.01	739822.9
VACUUM PITCH	IPL	0.9639	4227.4	8.70	647.11	8.62	236196.8
VACUUM GAS OIL	LLOYD	0.9365	5701.1	11.74	847.90	11.30	309482.2
VACUUM PITCH	LLOYD	1.0254	24461.9	50.36	3983.42	53.07	1453949.3
TOTAL...		.	48578.2	100.00	7505.35	100.00	2739451.2

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

DELAYED COKER	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
VACUUM PITCH	IPL	0.9639	4227.4	14.74	647.11	13.97	236196.8
VACUUM PITCH	LLOYD	1.0254	24461.9	85.26	3983.42	86.03	1453949.3
TOTAL...		.	28689.3	100.00	4630.54	100.00	1690146.0

PRODUCTS

HYDROGEN SULFIDE	.	.	.	41.74	0.90	15235.5
UNSATURATE REFINERY GAS	.	.	.	189.62	4.10	69212.0
PROPYLENE	0.5220	444.5	1.55	36.85	0.80	13450.3
PROPANE	0.5077	726.0	2.53	58.54	1.26	21366.7
BUTYLENE	0.6100	434.8	1.52	42.12	0.91	15373.0
ISO BUTANE	0.5626	95.5	0.33	8.53	0.18	3113.1
NORMAL BUTANE	0.5835	469.5	1.64	43.51	0.94	15880.0
COKER NAPHTHA	0.7440	4958.3	17.28	585.84	12.65	213830.8
COKER DISTILLATE	0.8871	10279.6	35.83	1448.18	31.27	528586.9
COKER GAS OIL	0.9646	5529.9	19.28	847.11	18.29	309195.5
HIGH SULFUR COKE	.	.	.	1328.50	28.69	484902.3
TOTAL...	.	22938.2	79.95	4630.54	100.00	1690146.0

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

KEROSENE HYDROTREATER CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
KEROSENE	IPL	0.8109	5519.2	64.85	710.75	64.17	259422.3
KEROSENE	LLOYD	0.8324	2992.2	35.15	395.54	35.71	144371.5
HYDROGEN		.	.	.	1.33	0.12	484.6
TOTAL...		.	8511.3	100.00	1107.61	100.00	404278.4

PRODUCTS

HYDROGEN SULFIDE	.	.	.	2.10	0.19	767.2
SATURATE REFINERY GAS	.	.	.	0.66	0.06	242.3
HTD KEROSENE - HIGH SP	0.8149	8537.4	100.31	1104.85	99.75	403268.9
TOTAL...	.	8537.4	100.31	1107.61	100.00	404278.4

DIESEL HYDROTREATER CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
DIESEL	IPL	0.8468	8929.2	37.01	1200.79	35.71	438287.2
DIESEL	LLOYD	0.8899	4917.8	20.38	695.00	20.67	253674.1
COKER DISTILLATE		0.8871	10279.6	42.61	1448.18	43.07	528586.9
HYDROGEN		.	.	.	18.27	0.54	6669.8
TOTAL...		.	24126.6	100.00	3362.24	100.00	1227218.0

PRODUCTS

HYDROGEN SULFIDE	.	.	.	49.30	1.47	17995.0
SATURATE REFINERY GAS	.	.	.	8.95	0.27	3265.7
HYDROTREATED COKER DIESE	0.8417	10571.9	43.82	1413.14	42.03	515795.1
HYDROTREATED SR. DIESEL	0.8547	13930.7	57.74	1890.86	56.24	690162.2
TOTAL...	.	24502.6	101.56	3362.24	100.00	1227218.0

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

GAS OIL HYDROTREATER CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
VACUUM GAS OIL LLOYD	0.9365	5701.1	50.76	847.90	49.67	309482.2
COKER GAS OIL	0.9646	5529.9	49.24	847.11	49.62	309195.5
HYDROGEN	.	.	.	12.12	0.71	4422.4
TOTAL...	.	11231.1	100.00	1707.12	100.00	623100.1

PRODUCTS

HYDROGEN SULFIDE	.	.	.	40.33	2.36	14720.4
SATURATE REFINERY GAS	.	.	.	10.68	0.63	3897.6
HTD MIXED GAS OILS	0.9202	5730.2	51.02	837.38	49.05	305644.6
HTD COKER DIESEL	0.8994	5732.1	51.04	818.73	47.96	298837.4
TOTAL...	.	11462.3	102.06	1707.12	100.00	623100.1

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

FLUID CATALYTIC CRACKER CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HTD MIXED GAS OILS	0.9202	5730.2	22.34	837.38	22.74	305644.6
VACUUM GAS OIL IPL	0.8996	14187.7	55.31	2026.91	55.03	739822.9
HTD COKER DIESEL	0.8994	5732.1	22.35	818.73	22.23	298837.4
TOTAL...	.	25650.0	100.00	3683.03	100.00	1344305.0

PRODUCTS

HYDROGEN SULFIDE	.	.	.	3.31	0.09	1209.9
UNSATURATE REFINERY GAS	.	.	.	142.53	3.87	52024.6
PROPYLENE	0.5220	2057.0	8.02	170.52	4.63	62241.3
PROPANE	0.5077	895.3	3.49	72.19	1.96	26348.4
BUTYLENE	0.6100	2163.3	8.43	209.56	5.69	76491.0
ISO BUTANE	0.5626	1615.9	6.30	144.37	3.92	52696.8
NORMAL BUTANE	0.5835	480.9	1.87	44.56	1.21	16266.1
FCC GASOLINE	0.7568	15138.3	59.02	1819.42	49.40	664086.7
LIGHT CYCLE OIL	0.9513	3576.4	13.94	540.30	14.67	197209.5
SLURRY OIL	1.0520	1968.6	7.68	328.89	8.93	120046.4
FCC COKE	.	.	.	207.35	5.63	75684.4
TOTAL...	.	27895.8	108.76	3683.03	100.00	1344305.0

CATALYTIC CONDENSATION CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
PROPYLENE	0.5220	2501.6	100.00	207.37	100.00	75691.6
TOTAL...	.	2501.6	100.00	207.37	100.00	75691.6

PRODUCTS

PROPYLENE	0.5077	95.2	3.80	7.67	3.70	2800.6
MOTOR FUEL POLYMER	0.7400	1699.3	67.93	199.70	96.30	72891.0
TOTAL...	.	1794.5	71.73	207.37	100.00	75691.6

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

REFINERY FUEL BLENDING CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
1.5% S REFINERY FUEL	1.0520	1434.2	100.00	239.60	100.00	87454.8
TOTAL...	.	1434.2	100.00	239.60	100.00	87454.8

PRODUCTS

REFINERY FUEL OIL	.	.	.	239.48	99.95	87411.1
SOX EMISSIONS	.	.	.	0.12	0.05	43.7
TOTAL...	.	.	.	239.60	100.00	87454.8

ALKYLATION UNIT CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
BUTYLENE	0.6100	2598.1	46.43	251.68	48.45	91863.9
ISO BUTANE	0.5626	2997.1	53.57	267.78	51.55	97738.0
TOTAL...	.	5595.1	100.00	519.46	100.00	189601.9

PRODUCTS

NORMAL BUTANE	0.5835	51.6	0.92	4.78	0.92	1745.3
ALKYLATE FROM BUTYLENE	0.7036	4601.5	82.24	514.16	98.98	187668.9
TAR FROM HF ALKY UNIT	.	.	.	0.51	0.10	187.7
TOTAL...	.	4653.1	83.16	519.46	100.00	189601.9

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

NAPHTHA HYDROTREATER CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
NAPHTHA	IPL	0.7389	10238.4	42.59	1201.41	42.71	438514.8
NAPHTHA	LLOYD	0.7459	6814.5	28.35	807.21	28.69	294630.5
LT ST RUN	IPL	0.6542	2028.4	8.44	210.73	7.49	76917.2
COKER NAPHTHA		0.7440	4958.3	20.63	585.84	20.83	213830.8
HYDROGEN		.	.	.	7.95	0.28	2902.3
TOTAL...		.	24039.5	100.00	2813.14	100.00	1026795.6

PRODUCTS

HYDROGEN SULFIDE	.	.	.	8.48	0.30	3094.8
SATURATE REFINERY GAS	.	.	.	0.74	0.03	268.9
PROPANE	0.5077	5.7	0.02	0.46	0.02	166.5
NORMAL BUTANE	0.5835	5.6	0.02	0.51	0.02	187.9
HYDROTREATED LT NAPHTHA	0.6540	2128.0	8.85	221.02	7.86	80671.2
HYDROTREATED NAPHTHA	0.7270	22363.4	93.03	2581.93	91.78	942406.2
TOTAL...	.	24502.6	101.93	2813.14	100.00	1026795.6

C5/C6 PENEX UNIT CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HYDROTREATED LT NAPHTHA	0.6540	2128.0	100.00	221.02	99.13	80671.2
HYDROGEN	.	.	.	1.94	0.87	709.9
TOTAL...	.	2128.0	100.00	222.96	100.00	81381.1

PRODUCTS

SATURATE REFINERY GAS	.	.	.	10.83	4.86	3952.9
C5/C6 PENEXATE	0.6440	2074.2	97.47	212.13	95.14	77428.3
TOTAL...	.	2074.2	97.47	222.96	100.00	81381.1

DETAILED MATERIAL BALANCE

PRODUCTS

AMINE TR- SULFUR PLANT CHARGE

PRODUCTS

SULFUR	.	.	.	143.82	99.00	52492.5
SOX EMISSIONS	.	.	.	1.45	1.00	530.2
TOTAL...	.	.	.	145.27	100.00	53022.8

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

OFFSITE ALLOWANCE	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
	IPL CRUDE	0.8265	47500.0	50.00	6234.62	47.06	2275634.6
	LLOYDMINISTER CRUDE	0.9297	47500.0	50.00	7013.09	52.94	2559779.2
	TOTAL,,	.	95000.0	100.00	13247.71	100.00	4835413.9

PRODUCTS

IPL CRUDE	0.8265	47500.0	50.00	6234.62	47.06	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	50.00	7013.09	52.94	2559779.2
TOTAL...	.	95000.0	100.00	13247.71	100.00	4835413.9

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY CONSUMPTION SUMMARY

PROCESS UNITS	COMBINED MODE	LABOR MAINTNCE INSURANCE DOLLARS	CATALYST CONSUMED DOLLARS	POWER CONSUMED KWH	FUEL CONSUMED MM BTU	HIGH PRESSURE STEAM M LBS	MEDIUM PRESSURE STEAM M LBS	LOW PRESSURE STEAM M LBS	COOLING WATER M GAL
CRUDE UNIT		7471.71	.	58952.30	6557.62	.	543.16	.	168.25
VACUUM UNIT		4225.51	.	13029.28	4000.35	.	825.59	112.58	5208.71
DELAYED COKER		11576.34	.	66332.44	5556.64	.	-208.37	79.65	2155.52
KEROSENE HYDROTREATER		2780.09	.	7821.43	297.59	66.38	.	.	27.66
DIESEL HYDROTREATER		5601.15	.	32581.37	899.53	200.64	.	.	83.60
GAS OIL HYDROTREATER		1703.48	.	21387.60	455.96	101.70	.	.	42.38
FLUID CATALYTIC CRACKER		10713.93	3355.24	30381.29	.	1974.10	-1042.30	-1974.10	2533.92
CATALYTIC CONDENSATION		1658.99	539.17	7776.54	663.60	91.24	.	.	466.59
ALKYLATION UNIT		3159.01	292.04	9089.86	1029.35	.	.	411.33	8519.65
NAPHTHA HYDROTREATER		3534.54	.	20177.23	1883.12	.	.	.	1468.80
C5/C6 PENEX UNIT		1394.62	222.12	3199.44	.	.	.	137.47	149.41
PLATFORMING UNIT		3614.71	903.68	32093.45	5112.23	-707.45	.	.	1272.89
AMINE TR- SULFUR PLANT		6591.82	145.27	15013.00	426.94	.	-213.40	871.61	256.11
OFFSITE ALLOWANCE		13009.25	.	21713.00	.	.	2053.39	.	.
TOTAL...		77035.14	5457.82	339548.23	26882.93	1726.61	1958.07	-361.47	22353.48

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY COST

COST

LABOR	MAINTNCE	INSURANCE	DOLLARS	1.00
CATALYST	CONSUMED		DOLLARS	1.00
POWER	CONSUMED		KWH	0.05
FUEL	CONSUMED		MM BTU	2.25
HIGH	PRESSURE	STEAM	M LBS	3.65
MEDIUM	PRESSURE	STEAM	M LBS	3.20
LOW	PRESSURE	STEAM	M LBS	3.15
COOLING	WATER		M GAL	0.10

UTILITY CONSUMPTION COST SUMMARY

	LABOR MAINTNCE INSURANCE DOLLARS	CATALYST CONSUMED DOLLARS	POWER CONSUMED DOLLARS	FUEL CONSUMED DOLLARS	HIGH PRESSURE STEAM DOLLARS	MEDIUM PRESSURE STEAM DOLLARS	LOW PRESSURE STEAM DOLLARS	COOLING WATER DOLLARS
CRUDE UNIT	7471.71	.	2947.61	14754.63	.	1738.10	.	16.82
VACUUM UNIT	4225.51	.	651.46	9000.79	.	2641.88	354.63	520.87
DELAYED COKER	11576.34	.	3316.62	12502.45	.	-666.80	250.88	215.55
KEROSENE HYDROTREATER	2780.09	.	391.07	669.58	242.28	.	.	2.77
DIESEL HYDROTREATER	5601.15	.	1629.07	2023.94	732.33	.	.	8.36
GAS OIL HYDROTREATER	1703.48	.	1069.38	1025.90	371.21	.	.	4.24
FLUID CATALYTIC CRACKER	10713.93	3355.24	1519.06	.	7205.47	-3335.35	-6218.42	253.39
CATALYTIC CONDENSATION	1658.99	539.17	388.83	1493.10	333.04	.	.	46.66
ALKYLATION UNIT	3159.01	292.04	454.49	2316.04	.	.	1295.69	851.97
NAPHTHA HYDROTREATER	3534.54	.	1008.86	4237.02	.	.	.	146.88
C5/C6 PENEX UNIT	1394.62	222.12	159.97	.	.	.	433.04	14.94
PLATFORMING UNIT	3614.71	903.68	1604.67	11502.52	-2582.19	.	.	127.29
AMINE TR- SULFUR PLANT	6591.82	145.27	750.65	960.62	.	-682.88	2745.56	25.61
OFFSITE ALLOWANCE	13009.25	.	1085.65	.	.	6570.86	.	.
TOTAL COST (BY UTILITY)	77035.14	5457.52	16977.41	60486.59	6302.14	6265.82	-1138.63	2235.35

UTILITY CONSUMPTION COST SUMMARY

	TOTAL COST (BY UNIT)
CRUDE UNIT	26928.88
VACUUM UNIT	17395.14
DELAYED COKER	27195.05
KEROSENE HYDROTREATER	4085.79
DIESEL HYDROTREATER	9994.84
GAS OIL HYDROTREATER	4174.21
FLUID CATALYTIC CRACKER	13493.32
CATALYTIC CONDENSATION	4459.79
ALKYLATION UNIT	8369.24
NAPHTHA HYDROTREATER	8927.30
C5/C6 PENEX UNIT	2224.69
PLATFORMING UNIT	15170.67
AMINE TR- SULFUR PLANT	10536.66
OFFSITE ALLOWANCE	20665.76
TOTAL COST (BY UTILITY)	173621.35

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

PLANT CAPACITY SUMMARY II
--- BY CAPACITY TYPES ---

PROCESS UNIT	UNITS OF CAPACITY	MAXIMUM	MINIMUM	CALENDAR DAY BASIS	STREAM DAY BASIS	INVSTMNT	PER UNIT OPERATIN COST
CRUDE UNIT	BBLS CAPACITY	100000.0	.	95000.0	100000.0	40700000.	0.28
VACUUM UNIT	BBLS CAPACITY	.	.	48570.1	51126.4	14322398.	0.36
DELAYED COKER	BBLS CAPACITY	.	.	28689.3	30199.3	63291540.	0.95
KEROSENE HYDROTREATER	BBLS CAPACITY	.	.	8511.3	8959.3	6214472.	0.48
DIESEL HYDROTREATER	BBLS CAPACITY	.	.	24203.6	25477.5	17818083.	0.41
GAS OIL HYDROTREATER	BBLS CAPACITY	.	.	11275.0	11868.4	13349110.	0.37
FLUID CATALYTIC CRACKER	BBLS CAPACITY	.	.	25650.0	27000.0	62836927.	0.53
CATALYTIC CONDENSATION	BBLS CAPACITY	.	.	2571.4	2706.8	7218950.	1.73
ALKYLATION UNIT	BBLS CAPACITY	.	.	4601.5	4843.7	22557501.	1.82
NAPHTHA HYDROTREATER	BBLS CAPACITY	.	.	24045.5	25311.1	7093180.	0.37
C5/C6 PENEX UNIT	BBLS CAPACITY	.	.	2128.0	2240.0	5524031.	1.05
PLATFORMING UNIT	BBLS CAPACITY	.	.	22062.9	23224.2	34034621.	0.69
OFFSITE ALLOWANCE	BBLS CAPACITY	.	.	95000.0	95000.0	365402064.	0.22
AMINE TR- SULFUR PLANT	TONS CAPACITY	.	.	145.3	152.9	20994445.	72.53
<u>TOTAL INVESTMENT</u>						681357322.	

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

RAW MATERIAL PURCHASE SUMMARY

TIME 1 BASE MODEL

PURCHASED MATERIALS

		MAXIMUM	MINIMUM	QUANTITY BOUGHT	PRICE	TOTAL \$ PER DAY
IPL CRUDE	BBLS	47500.00	47500.00	47500.00	13.47	639825.00
LLOYDMINISTER CRUDE	BBLS	47500.00	47500.00	47500.00	10.48	497800.00
ISOBUTANE	BBLS	.	.	909.58	16.81	15290.01
NORMAL BUTANE	BBLS	.	.	1159.81	15.41	17872.68

TOTAL RAW MATERIAL PURCHASE

1170787.69

RAW MATERIAL PURCHASE RESTRICTIONS:

Appendix D.2
Conventional Refinery High Crude Price Case

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY

OVERALL ECONOMIC SUMMARY

TIME 1 BASE MODEL

<u>GROSS REVENUES AT MARKET</u>	DOLLARS PER DAY	\$MM PER YEAR
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CRACKING REFINERY	4116280.27	1502.44
TOTAL...	4116280.27	1502.44

RAW MATERIAL PURCHASES

CRACKING REFINERY	3579763.19	1306.61
TOTAL...	3579763.19	1306.61

PROCESS UNIT OPERATING COSTS

CRACKING REFINERY	79833.39	29.14
TOTAL...	79833.39	29.14

UTILITY POOL COSTS

CRACKING REFINERY	14525.24	5.30
TOTAL...	14525.24	5.30

TRANSPORTATION COSTS

CRACKING REFINERY	9.56	0.00
TOTAL...	9.56	0.00

CAPITAL CHARGES

CRACKING REFINERY	527489.11	192.53
TOTAL...	527489.11	192.53

OPTIMIZED PROFIT

	-85340.23	-31.15
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY

MARKET CENTER SUMMARY

PRODUCT SALESBASE MODEL

PRODUCT SALES

		MAXIMUM	MINIMUM	QUANTITY SOLD	PRICE \$	TOTAL \$ PER DAY
LIQUID PROPANE GAS	BBLS	.	.	2341.54	31.10	72821.89
UNLEADED PRE. GASOLINE	BBLS	.	.	16793.83	51.95	872439.62
UNLEADED REG. GASOLINE	BBLS	.	.	33587.67	44.10	1481216.19
JET FUEL	BBLS	.	.	8537.38	42.92	366424.15
.05% S NO.2 FUEL	BBLS	.	.	22701.79	40.74	924871.02
1.0% S NO.6 FUEL	BBLS	.	.	10449.81	33.64	351531.60
COKE	TONS	.	.	1096.68	30.00	32900.29
SULFUR	TONS	.	.	127.96	110.00	14075.49
<u>TOTAL SALES REVENUE</u>						4116280.27

MARKET RESTRICTIONS:

SALES	PRODUCT SALES	FIXED	UNLEADED REG TO UNLEADED PRM
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

REFINERY ECONOMIC SUMMARY

	DOLLARS PER DAY	\$MM PER YEAR
SALES REVENUE	4116270.70	1502.4388
COST OF RAW MATERIALS	3579763.19	1306.6136
PROCESS UNIT OPERATING COST	79833.39	29.1392
UTILITY POOL COSTS	14525.24	5.3017
CAPITAL CHARGES	527489.11	192.5335
OPTIMIZED PROFIT	-85340.23	-31.1492

PROFITABILITY ANALYSIS

GROSS PROFIT	442148.88	161.3843
DEPRECIATION	149654.00	54.6237

PROFIT BEFORE TAXES		106.7606
ALLOWANCE FOR FEDRL INCOME TAX		36.2986
NET PROFIT AFTER TAXES		70.4620

<u>CASH FLOW</u>	125.0857
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INVESTMENT

	\$MM
PLANT INVESTMENT	675.5562
OFFSITE INVESTMENT	.
TOTAL INVESTMENT	675.5562

<u>SIMPLE PAYOUT TIME (IN YEARS)</u>	5.4
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

GROSS MARGIN SUMMARY

MARKET	PRODUCTS		MAXIMUM	MINIMUM	QUANTITY SOLD	PRICE	TOTAL \$ PER DAY
SALES	LIQUID PROPANE GAS	BBLS	.	.	2341.54	31.10	73821.66
SALES	UNLEADED PRE. GASOLINE	BBLS	.	.	16793.83	51.95	874437.94
SALES	UNLEADED REG. GASOLINE	BBLS	.	.	33587.67	44.10	1481212.83
SALES	JET FUEL	BBLS	.	.	8537.38	42.92	366423.30
SALES	.05% S NO.2 FUEL	BBLS	.	.	22701.79	40.74	924868.75
SALES	1.0% S NO.6 FUEL	BBLS	.	.	10449.81	33.64	351530.56
SALES	COKE	TONS	.	.	1096.68	30.00	32900.18
SALES	SULFUR	TONS	.	.	127.96	110.00	14075.48

TOTAL SALES REVENUE 4116270.70

RAW MATERIAL PURCHASED

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			MAXIMUM	MINIMUM	QUANTITY PURCHASE	PRICE	TOTAL \$ PER DAY
IPL CRUDE	BBLS	47500.00	47500.00	47500.00	38.36	1822100.00	
LLOYDMINISTER CRUDE	BBLS	47500.00	47500.00	47500.00	33.00	1567500.00	
ISOBUTANE	BBLS	.	.	3986.50	28.70	114412.64	
NORMAL BUTANE	BBLS	.	.	3066.82	24.70	75750.55	

TOTAL RAW MATERIAL COST 3579763.19

GROSS MARGIN 536507.51

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

OPERATING COST SUMMARY

PROCESS UNIT BY COMBINED MODE

	ACTIVITY TONS/DAY	OPERATING COST/DAY
CRUDE UNIT	13247.71	7471.71
VACUUM UNIT	7505.35	4225.51
DELAYED COKER	3839.90	9599.76
KEROSENE HYDROTREATER	1106.28	2780.09
DIESEL HYDROTREATER	3097.29	5187.96
GAS OIL HYDROTREATER	1554.82	1562.60
FLUID CATALYTIC CRACKER	3547.54	13551.60
ALKYLATION UNIT	941.16	6317.04
NAPHTHA HYDROTREATER	2799.00	3526.74
C5/C6 PENEX UNIT	312.73	2287.64
PLATFORMING UNIT	2484.22	4319.21
AMINE TR- SULFUR PLANT	129.25	5994.30
OFFSITE ALLOWANCE	13247.71	13009.25
<u>TOTAL PROCESS UNIT OPERATING COST</u>		<u>79833.39</u>

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY PURCHASES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
POWER CONSUMED	KWH		346007.17	0.04	13840.29
<u>TOTAL PURCHASED UTILITY COST</u>					13840.29

UNPOOLED UTILITIES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
CATALYSTCONSUMED	DOLLARS		5051.23	1.00	5051.23
LABOR MAINTNCEINSURNCEDOLLARS			74782.16	1.00	74782.16
<u>TOTAL UNPOOLED UTILITY COST</u>					79833.39
<u>TOTAL UTILITY CONVERSION COST</u>					684.96
<u>TOTAL OPERATING COST</u>					94358.63

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

MATERIAL BALANCE SUMMARY

RAW MATERIALS

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE	0.8265	47500.0	46.54	6234.62	44.89	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	46.54	7013.09	50.50	2559779.2
ISOBUTANE	0.5626	3986.5	3.91	356.18	2.57	130004.4
NORMAL BUTANE	0.5835	3066.8	3.01	284.19	2.05	103727.9
TOTAL CHARGE	.	102053.3	100.00	13888.07	100.00	5069146.2

REFINERY PRODUCTS

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
LIQUID PROPANE GAS	0.5077	2341.5	2.29	188.79	1.36	68908.8
UNLEADED PRE. GASOLINE	0.7613	16793.8	16.46	2030.39	14.62	741094.1
UNLEADED REG. GASOLINE	0.7304	33587.7	32.91	3895.96	28.05	1422023.8
JET FUEL	0.8149	8537.4	8.37	1104.85	7.96	403268.9
.05% S NO.2 FUEL	0.8497	22701.8	22.25	3063.28	22.06	1118098.9
1.0% S NO.6 FUEL	0.9809	10449.8	10.24	1627.85	11.72	594166.3
COKE	.	.	.	1096.68	7.90	400286.9
SULFUR	.	.	.	127.96	0.92	46705.0
TOTAL PRODUCTS	.	94412.0	92.51	13135.76	94.58	4794552.8

STREAMS CONVERTED TO UTILITIES

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
UNSATURATE REFINERY GAS	.	.	.	290.89	2.09	106173.4
SATURATE REFINERY GAS	.	.	.	61.82	0.45	22563.2
HYDROGEN	.	.	.	109.53	0.79	39979.5
TOTAL STREAMS CONVERTED	.	.	.	462.24	3.33	168716.0

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
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CRACKING REFINERYBASE MODEL

MATERIAL BALANCE SUMMARY CONTINUED

STREAMS NOT UTILIZED

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
FCC COKE	.	.	.	199.73	1.44	72900.2
SOX EMISSIONS	.	.	.	1.29	0.01	471.8
LLOYDMINISTER CRUDE	0.9297	0.0	0.00	0.00	0.00	0.6
PLANT LOSSES	.	.	.	87.79	0.63	32044.1
TAR FROM HF ALKY UNIT	.	.	.	1.26	0.01	460.8
TOTAL NOT UTILIZED	.	0.0	0.00	290.07	2.09	105877.4
<u>TOTAL PRODUCTS MADE</u>		94412.0	92.513	13888.07	100.000	5069146.2

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

 PLANT CAPACITY SUMMARY

PROCESS UNIT	UNITS OF CAPACITY	MAXIMUM	MINIMUM	CALENDAR DAY BASIS	STREAM DAY BASIS	INVTMNT
CRUDE UNIT	BBLS CAPACITY	100000.0	.	95000.0	100000.0	40700000.
VACUUM UNIT	BBLS CAPACITY	.	.	48570.1	51126.4	14822398.
DELAYED COKER	BBLS CAPACITY	.	.	23580.5	24821.6	55716982.
KEROSENE HYDROTREATER	BBLS CAPACITY	.	.	8511.3	8959.3	6214472.
DIESEL HYDROTREATER	BBLS CAPACITY	.	.	22439.5	23620.6	16898731.
GAS OIL HYDROTREATER	BBLS CAPACITY	.	.	10352.6	10897.5	12574929.
FLUID CATALYTIC CRACKER	BBLS CAPACITY	.	.	24701.4	26001.5	61154971.
HYDROGEN PLANT	TONS CAPACITY	.	.	71.8	79.8	.
ALKYLATION UNIT	BBLS CAPACITY	.	.	8422.9	8866.2	36039317.
NAPHTHA HYDROTREATER	BBLS CAPACITY	.	.	24100.8	25369.2	7104583.
C5/C6 PENEX UNIT	BBLS CAPACITY	.	.	3011.1	3169.5	7043340.
PLATFORMING UNIT	BBLS CAPACITY	.	.	21228.0	22345.2	33038460.
AMINE TR- SULFUR PLANT	TONS CAPACITY	.	.	129.3	136.1	19345979.
OFFSITE ALLOWANCE	BBLS CAPACITY	.	.	95000.0	95000.0	365402064.
<u>TOTAL INVESTMENT</u>						675556227.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

PRODUCT BLENDING

LIQUID PROPANE GAS

	BARRELS PER DAY	TONS PER DAY	SG 60/60
PROPANE	2341.5	188.79	0.5077
BLEND	2341.5	188.79	0.5077
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

UNLEADED PRE. GASOLINE

	BARRELS PER DAY	TONS PER DAY	SG 60/60	RESEARCH OCTANE NUMBER	MOTOR OCTANE NUMBER	VAPOR PRESSURE INDEX
NORMAL BUTANE	1355.4	125.60	0.5835	93.60	90.10	1120.00
C5/C6 PENEXATE	2473.8	253.00	0.6440	88.40	86.00	200.00
94 RON PLATFORMATE	3068.1	385.36	0.7909	94.00	83.70	24.00
98 RON PLATFORMATE	9896.5	1266.43	0.8058	98.00	87.00	25.05
BLEND	16793.8	2030.39	0.7613	95.50	86.50	139.00
MAXIMUM SPECIFICATION	.	.	.	95.50	86.50	139.00
MINIMUM SPECIFICATION	.	.	.	95.50	86.50	.

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

 PRODUCT BLENDING

UNLEADED REG. GASOLINE

	BARRELS PER DAY	TONS PER DAY	SG 60/60	RESEARCH OCTANE NUMBER	MOTOR OCTANE NUMBER	VAPOR PRESSURE INDEX
ALKYLATE FROM PROPYLENE	4107.6	458.97	0.7036	92.00	90.00	40.10
ALKYLATE FROM BUTYLENE	4315.3	482.19	0.7036	94.50	92.50	40.10
LT ST RUN LLOYD	2531.4	262.99	0.6542	70.00	68.00	160.00
FCC GASOLINE	14581.4	1752.48	0.7568	91.90	80.60	87.50
NORMAL BUTANE	2163.5	200.48	0.5835	93.60	90.10	1120.00
C5/C6 PENEXATE	461.1	47.16	0.6440	88.40	86.00	200.00
94 RON PLATFORMATE	1199.8	150.70	0.7909	94.00	83.70	24.00
98 RON PLATFORMATE	4227.5	540.99	0.8058	98.00	87.00	25.05

BLEND	33587.7	3895.96	0.7304	91.50	83.93	139.00
MAXIMUM SPECIFICATION	.	.	.	.	.	139.00
MINIMUM SPECIFICATION	.	.	.	91.50	82.50	.

JET FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	SMOKE POINT
HTD KEROSENE - HIGH SP	8537.4	1104.85	0.8149	0.01	21.00
BLEND	8537.4	1104.85	0.8149	0.01	21.00
MAXIMUM SPECIFICATION	.	.	.	0.10	.
MINIMUM SPECIFICATION	.	.	.	.	21.00

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
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CRACKING REFINERYBASE MODEL

 PRODUCT BLENDING

.05% S NO.2 FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	CETANE INDEX
HYDROTREATED SR. DIESEL	13930.7	1890.86	0.8547	0.05	51.00
HYDROTREATED COKER DIESEL	8771.1	1172.43	0.8417	0.05	46.00
BLEND	22701.8	3063.28	0.8497	0.05	49.07
MAXIMUM SPECIFICATION	.	.	.	0.05	.
MINIMUM SPECIFICATION	.	.	.	.	40.00

1.0% S NO.6 FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	VISCOSITY INDEX @ 122 F
VACUUM PITCH IPL	4227.4	647.11	0.9639	0.54	45.01
VACUUM PITCH LLOYD	881.3	143.52	1.0254	5.10	45.73
LIGHT CYCLE OIL	3444.8	520.42	0.9513	0.35	15.40
SLURRY OIL	1896.2	316.80	1.0520	1.15	32.00
BLEND	10449.8	1627.85	0.9809	1.00	33.08
MAXIMUM SPECIFICATION	.	.	0.9850	1.00	37.50
MINIMUM SPECIFICATION	.	.	.	.	.

UOP PROCESS DIVISION
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BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

PRODUCT BLENDING

COKE

	BARRELS PER DAY	TONS PER DAY	SG 60/60
HIGH SULFUR COKE	.	1096.68	.
BLEND	.	1096.68	.
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

SULFUR

	BARRELS PER DAY	TONS PER DAY	SG 60/60
SULFUR	.	127.96	.
BLEND	.	127.96	.
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY BLENDING SUMMARY

MM BTU FUEL CONSUMED

	TONS PER DAY	UTILITY UNITS/D	CONVERSN COST \$/D
UNSATURATE REFINERY GAS	290.89	12508.09	.
SATURATE REFINERY GAS	61.82	2719.94	.
HYDROGEN	109.53	14677.40	.
TOTAL...	462.24	29905.44	.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY CONVERSION SUMMARY

M LBS HIGH PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
FUEL CONSUMED	4280.98	3638.84	684.96
TOTAL...	4280.98	3638.84	684.96

M LBS MEDIUM PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
HIGH PRESSURESTEAM	2055.62	2055.62	.
TOTAL...	2055.62	2055.62	.

M LBS LOW PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
MEDIUM PRESSURESTEAM	0.10	0.10	.
TOTAL...	0.10	0.10	.

UOP PROCESS DIVISION
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CRACKING REFINERYBASE MODEL

UTILITY CONVERSION SUMMARY

M GAL COOLING WATER

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
POWER CONSUMED	22281.11	28519.82	.
TOTAL...	22281.11	28519.82	.

UOP PROCESS DIVISION
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DETAILED MATERIAL BALANCE

CRUDE UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE		0.8265	47500.0	50.00	6234.62	47.06	2275634.6
LLOYDMINISTER CRUDE		0.9297	47500.0	50.00	7013.09	52.94	2559778.6
TOTAL...		.	95000.0	100.00	13247.71	100.00	4835413.3

PRODUCTS

SATURATE REFINERY GAS		.	.	.	3.12	0.02	1137.8
PROPANE		0.5077	409.8	0.43	33.04	0.25	12060.9
ISO BUTANE		0.5626	263.4	0.28	23.54	0.18	8590.3
NORMAL BUTANE		0.5835	1130.3	1.19	104.74	0.79	38229.9
LT ST RUN	IPL	0.6542	2928.5	3.08	304.25	2.30	111051.0
LT ST RUN	LLOYD	0.6542	2531.4	2.66	262.99	1.99	95991.7
NAPHTHA	IPL	0.7389	10238.4	10.78	1201.41	9.07	438514.8
NAPHTHA	LLOYD	0.7459	6814.5	7.17	807.21	6.09	294630.5
KEROSENE	IPL	0.8109	5519.2	5.81	710.75	5.37	259422.3
KEROSENE	LLOYD	0.8324	2992.2	3.15	395.54	2.99	144371.5
DIESEL	IPL	0.8468	8929.2	9.40	1200.79	9.06	438287.2
DIESEL	LLOYD	0.8899	4917.8	5.18	695.00	5.25	253674.1
REDUCED CRUDE	IPL	0.9144	18414.3	19.38	2674.03	20.18	976019.7
REDUCED CRUDE	LLOYD	1.0086	30162.9	31.75	4831.32	36.47	1763431.5
TOTAL...		.	95251.9	100.27	13247.71	100.00	4835413.3

UOP PROCESS DIVISION
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DETAILED MATERIAL BALANCE

VACUUM UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
REDUCED CRUDE	IPL	0.9144	18414.3	37.91	2674.03	35.63	976019.7
REDUCED CRUDE	LLOYD	1.0086	30162.9	62.09	4831.32	64.37	1763431.5
TOTAL...		.	48577.3	100.00	7505.35	100.00	2739451.2

PRODUCTS

VACUUM GAS OIL	IPL	0.8996	14187.7	29.21	2026.91	27.01	739822.9
VACUUM PITCH	IPL	0.9639	4227.4	8.70	647.11	8.62	236196.8
VACUUM GAS OIL	LLOYD	0.9365	5701.1	11.74	847.90	11.30	309482.2
VACUUM PITCH	LLOYD	1.0254	24461.9	50.36	3983.42	53.07	1453949.3
TOTAL...		.	48578.2	100.00	7505.35	100.00	2739451.2

DELAYED COKER	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
VACUUM PITCH	LLOYD	1.0254	23580.5	100.00	3839.90	100.00	1401564.7
TOTAL...		.	23580.5	100.00	3839.90	100.00	1401564.7

PRODUCTS

HYDROGEN SULFIDE	.	.	.	39.55	1.03	14436.1
UNSATURATE REFINERY GAS	.	.	.	153.60	4.00	56062.6
PROPYLENE	0.5220	370.6	1.57	30.72	0.80	11212.5
PROPANE	0.5077	600.1	2.54	48.38	1.26	17659.7
BUTYLENE	0.6100	352.8	1.50	34.18	0.89	12473.9
ISO BUTANE	0.5626	77.4	0.33	6.91	0.18	2522.8
NORMAL BUTANE	0.5835	381.2	1.62	35.33	0.92	12894.4
COKER NAPHTHA	0.7440	4114.4	17.45	486.13	12.66	177438.1
COKER DISTILLATE	0.8871	8528.6	36.17	1201.51	31.29	438549.6
COKER GAS OIL	0.9646	4614.8	19.57	706.93	18.41	258028.1
HIGH SULFUR COKE	.	.	.	1096.68	28.56	400286.9
TOTAL...	.	19039.9	80.74	3839.90	100.00	1401564.7

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CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

KEROSENE HYDROTREATER CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
KEROSENE	IPL	0.8109	5519.2	64.85	710.75	64.17	259422.3
KEROSENE	LLOYD	0.8324	2992.2	35.15	395.54	35.71	144371.5
HYDROGEN		.	.	.	1.33	0.12	484.6
TOTAL...		.	8511.3	100.00	1107.61	100.00	404278.4

PRODUCTS

HYDROGEN SULFIDE	.	.	.	2.10	0.19	767.2
SATURATE REFINERY GAS	.	.	.	0.66	0.06	242.3
HTD KEROSENE - HIGH SP	0.8149	8537.4	100.31	1104.85	99.75	403268.9
TOTAL...	.	8537.4	100.31	1107.61	100.00	404278.4

DIESEL HYDROTREATER CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
DIESEL	IPL	0.8468	8929.2	39.91	1200.79	38.57	438287.2
DIESEL	LLOYD	0.8899	4917.8	21.98	695.00	22.32	253674.1
COKER DISTILLATE		0.8871	8528.6	38.12	1201.51	38.60	438549.6
HYDROGEN		.	.	.	15.81	0.51	5769.4
TOTAL...		.	22375.6	100.00	3113.10	100.00	1136280.3

PRODUCTS

HYDROGEN SULFIDE	.	.	.	42.10	1.35	15365.9
SATURATE REFINERY GAS	.	.	.	7.71	0.25	2815.5
HYDROTREATED COKER DIESE	0.8417	8771.1	39.20	1172.43	37.66	427936.7
HYDROTREATED SR. DIESEL	0.8547	13930.7	62.26	1890.86	60.74	690162.2
TOTAL...	.	22701.8	101.46	3113.10	100.00	1136280.3

UOP PROCESS DIVISION
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CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

GAS OIL HYDROTREATER	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
VACUUM GAS OIL	LLOYD	0.9365	5701.1	55.27	847.90	54.17	309482.2
COKER GAS OIL		0.9646	4614.8	44.73	706.93	45.16	258028.1
HYDROGEN		.	.	.	10.55	0.67	3849.3
TOTAL...		.	10315.9	100.00	1565.37	100.00	571359.6

PRODUCTS

HYDROGEN SULFIDE	.	.	.	34.99	2.24	12771.0
SATURATE REFINERY GAS	.	.	.	9.75	0.62	3559.9
HTD MIXED GAS OILS	0.9202	5730.2	55.55	837.38	53.49	305644.6
HTD COKER DIESEL	0.8994	4783.5	46.37	683.24	43.65	249384.1
TOTAL...	.	10513.7	101.92	1565.37	100.00	571359.6

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CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

FLUID CATALYTIC CRACKER CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HTD MIXED GAS OILS	0.9202	5730.2	23.20	837.38	23.60	305644.6
VACUUM GAS OIL IPL	0.8996	14187.7	57.44	2026.91	57.14	739822.9
HTD COKER DIESEL	0.8994	4783.5	19.37	683.24	19.26	249384.1
TOTAL...	.	24701.4	100.00	3547.54	100.00	1294851.7

PRODUCTS

HYDROGEN SULFIDE	.	.	.	3.19	0.09	1165.4
UNSATURATE REFINERY GAS	.	.	.	137.29	3.87	50110.8
PROPYLENE	0.5220	1981.4	8.02	164.25	4.63	59951.6
PROPANE	0.5077	862.4	3.49	69.53	1.96	25379.1
BUTYLENE	0.6100	2083.7	8.44	201.85	5.69	73677.1
ISO BUTANE	0.5626	1556.5	6.30	139.06	3.92	50758.2
NORMAL BUTANE	0.5835	463.2	1.88	42.93	1.21	15667.7
FCC GASOLINE	0.7568	14581.4	59.03	1752.48	49.40	639656.7
LIGHT CYCLE OIL	0.9513	3444.8	13.95	520.42	14.67	189954.7
SLURRY OIL	1.0520	1896.2	7.68	316.80	8.93	115630.3
FCC COKE	.	.	.	199.73	5.63	72900.2
TOTAL...	.	26869.6	108.78	3547.54	100.00	1294851.7

HYDROGEN PLANT

CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
NORMAL BUTANE	0.5835	1722.6	100.00	159.62	100.00	58262.4
TOTAL...	.	1722.6	100.00	159.62	100.00	58262.4

PRODUCTS

HYDROGEN	.	.	.	71.83	45.00	26218.3
PLANT LOSSES	.	.	.	87.79	55.00	32044.1
TOTAL...	.	.	.	159.62	100.00	58262.4

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

ALKYLATION UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
PROPYLENE		0.5220	2351.9	21.83	194.97	20.19	71164.2
BUTYLENE		0.6100	2436.5	22.61	236.03	24.44	86151.0
ISO BUTANE		0.5626	5985.9	55.56	534.81	55.37	195206.2
TOTAL...		.	10774.3	100.00	965.81	100.00	352521.4

PRODUCTS

PROPANE	0.5077	234.5	2.18	18.91	1.96	6902.0
NORMAL BUTANE	0.5835	48.4	0.45	4.48	0.46	1636.8
ALKYLATE FROM PROPYLENE	0.7036	4107.6	38.12	458.97	47.52	167523.9
ALKYLATE FROM BUTYLENE	0.7036	4315.3	40.05	482.19	49.93	175997.9
TAR FROM HF ALKY UNIT	.	.	.	1.26	0.13	460.8
TOTAL...	.	8705.8	80.80	965.81	100.00	352521.4

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

NAPHTHA HYDROTREATER CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
NAPHTHA	IPL	0.7389	10238.4	42.49	1201.41	42.82	438514.8
NAPHTHA	LLOYD	0.7459	6814.5	28.28	807.21	28.77	294630.5
LT ST RUN	IPL	0.6542	2928.5	12.15	304.25	10.84	111051.0
COKER NAPHTHA		0.7440	4114.4	17.08	486.13	17.33	177438.1
HYDROGEN		.	.	.	6.88	0.25	2511.6
TOTAL...		.	24095.8	100.00	2805.88	100.00	1024145.9

PRODUCTS

HYDROGEN SULFIDE	.	.	.	7.32	0.26	2671.3
SATURATE REFINERY GAS	.	.	.	0.71	0.03	257.6
PROPANE	0.5077	5.3	0.02	0.43	0.02	155.4
NORMAL BUTANE	0.5835	5.1	0.02	0.47	0.02	173.1
HYDROTREATED LT NAPHTHA	0.6540	3011.1	12.50	312.73	11.15	114147.2
HYDROTREATED NAPHTHA	0.7270	21517.1	89.30	2484.22	88.54	906741.4
TOTAL...	.	24538.5	101.84	2805.88	100.00	1024145.9

C5/C6 PENEX UNIT CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HYDROTREATED LT NAPHTHA	0.6540	3011.1	100.00	312.73	99.13	114147.2
HYDROGEN	.	.	.	2.75	0.87	1004.5
TOTAL...	.	3011.1	100.00	315.48	100.00	115151.7

PRODUCTS

SATURATE REFINERY GAS	.	.	.	15.32	4.86	5593.2
C5/C6 PENEXATE	0.6440	2934.9	97.47	300.16	95.14	109558.5
TOTAL...	.	2934.9	97.47	315.48	100.00	115151.7

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

PLATFORMING UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HYDROTREATED NAPHTHA		0.7270	21517.1	100.00	2484.22	100.00	906741.4
TOTAL...		.	21517.1	100.00	2484.22	100.00	906741.4
PRODUCTS							
HYDROGEN		.	.	.	75.02	3.02	27380.5
SATURATE REFINERY GAS		.	.	.	24.54	0.99	8956.9
PROPANE		0.5077	229.4	1.07	18.50	0.74	6751.8
ISO BUTANE		0.5626	102.1	0.47	9.12	0.37	3330.5
NORMAL BUTANE		0.5835	146.4	0.68	13.56	0.55	4950.5
98 RON PLATFORMATE		0.8058	14124.0	65.64	1807.42	72.76	659708.4
94 RON PLATFORMATE		0.7909	4268.0	19.84	536.06	21.58	195662.8
TOTAL...		.	18869.9	87.70	2484.22	100.00	906741.4

AMINE TR- SULFUR PLANT CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HYDROGEN SULFIDE	.	.	.	129.25	100.00	47176.8
TOTAL...	.	.	.	129.25	100.00	47176.8
PRODUCTS						
SULFUR	.	.	.	127.96	99.00	46705.0
SOX EMISSIONS	.	.	.	1.29	1.00	471.8
TOTAL...	.	.	.	129.25	100.00	47176.8

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

OFFSITE ALLOWANCE	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
	I						
	IPL CRUDE	0.8265	47500.0	50.00	6234.62	47.06	2275634.6
	LLOYDMINISTER CRUDE	0.9297	47500.0	50.00	7013.09	52.94	2559779.2
	TOTAL...	.	95000.0	100.00	13247.71	100.00	4835413.9

PRODUCTS

IPL CRUDE	0.8265	47500.0	50.00	6234.62	47.06	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	50.00	7013.09	52.94	2559779.2
TOTAL...	.	95000.0	100.00	13247.71	100.00	4835413.9

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY CONSUMPTION SUMMARY

PROCESS UNITS	COMBINED MODE	LABOR MAINTNCE INSURANCE DOLLARS	CATALYST CONSUMED DOLLARS	POWER CONSUMED KWH	FUEL CONSUMED MM BTU	HIGH PRESSURE STEAM M LBS	MEDIUM PRESSURE STEAM M LBS	LOW PRESSURE STEAM M LBS	COOLING WATER M GAL
CRUDE UNIT		7471.71	.	58952.30	6557.62	.	543.16	.	168.25
VACUUM UNIT		4225.51	.	13029.28	4000.35	.	825.59	112.58	5208.71
DELAYED COKER		9599.76	.	55006.62	4607.88	.	-172.80	66.05	1787.48
KEROSENE HYDROTREATER		2780.09	.	7821.43	297.59	66.38	.	.	27.66
DIESEL HYDROTREATER		5187.96	.	29098.28	833.17	185.84	.	.	77.43
GAS OIL HYDROTREATER		1562.60	.	19012.87	418.25	93.29	.	.	38.87
FLUID CATALYTIC CRACKER		10319.79	3231.81	29263.65	.	1901.48	-1003.95	-1901.48	2440.71
ALKYLATION UNIT		5782.46	534.58	16638.69	1884.19	.	.	752.92	15594.95
NAPHTHA HYDROTREATER		3526.74	.	19099.38	1878.97	.	.	.	1465.56
C5/C6 PENEX UNIT		1973.34	314.30	4527.11	.	.	.	194.52	211.41
PLATFORMING UNIT		3477.91	841.29	36205.69	4766.57	-663.77	.	.	1270.94
AMINE TR- SULFUR PLANT		5865.05	129.25	13357.76	379.87	.	-189.87	775.51	227.87
OFFSITE ALLOWANCE		13009.25	.	21713.00	.	.	2053.39	.	.
TOTAL...		74782.16	5051.23	323726.05	25624.46	1583.22	2055.52	0.10	28519.82

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

UTILITY COST

COST

LABOR	MAINTNCE	INSURANCE	DOLLARS	1.00
CATALYST	CONSUMED		DOLLARS	1.00
POWER	CONSUMED		KWH	0.06
FUEL	CONSUMED		MM BTU	3.75
HIGH	PRESSURE	STEAM	M LBS	7.20
MEDIUM	PRESSURE	STEAM	M LBS	5.28
LOW	PRESSURE	STEAM	M LBS	4.80
COOLING	WATER		M GAL	0.10

UTILITY CONSUMPTION COST SUMMARY

	LABOR MAINTNCE INSURANCE DOLLARS	CATALYST CONSUMED DOLLARS	POWER CONSUMED DOLLARS	FUEL CONSUMED DOLLARS	HIGH PRESSURE STEAM DOLLARS	MEDIUM PRESSURE STEAM DOLLARS	LOW PRESSURE STEAM DOLLARS	COOLING WATER DOLLARS
CRUDE UNIT	7471.71	.	3537.14	24591.06	.	2867.86	.	16.82
VACUUM UNIT	4225.51	.	781.76	15001.31	.	4359.10	540.38	520.87
DELAYED COKER	9599.76	.	3300.40	17279.57	.	-912.36	317.02	178.75
KEROSENE HYDROTREATER	2780.09	.	469.29	1115.96	477.91	.	.	2.77
DIESEL HYDROTREATER	5187.96	.	1745.90	3124.39	1338.03	.	.	7.74
GAS OIL HYDROTREATER	1562.60	.	1140.77	1568.43	671.68	.	.	3.89
FLUID CATALYTIC CRACKER	10319.79	3231.81	1755.82	.	13690.66	-5300.87	-9127.11	244.07
ALKYLATION UNIT	5782.46	534.58	998.32	7065.73	.	.	3614.04	1559.49
NAPHTHA HYDROTREATER	3526.74	.	1145.96	7046.13	.	.	.	146.56
C5/C6 PENEX UNIT	1973.34	314.30	271.63	.	.	.	933.69	21.14
PLATFORMING UNIT	3477.91	841.29	2172.34	17874.63	-4779.12	.	.	127.09
AMINE TR- SULFUR PLANT	5865.05	129.25	801.47	1424.51	.	-1002.52	3722.44	22.79
OFFSITE ALLOWANCE	13009.25	.	1302.78	.	.	10841.93	.	.
TOTAL COST (BY UTILITY)	74782.16	5051.23	19423.56	96091.71	11399.17	10853.14	0.47	2851.98

UTILITY CONSUMPTION COST SUMMARY

TOTAL
COST
(BY
UNIT)

CRUDE UNIT	38484.59
VACUUM UNIT	25428.94
DELAYED COKER	29763.13
KEROSENE HYDROTREATER	4846.02
DIESEL HYDROTREATER	11404.02
GAS OIL HYDROTREATER	4947.37
FLUID CATALYTIC CRACKER	14814.17
ALKYLATION UNIT	19554.62
NAPHTHA HYDROTREATER	11865.38
C5/C6 PENEX UNIT	3514.10
PLATFORMING UNIT	19714.15
AMINE TR- SULFUR PLANT	10962.99
OFFSITE ALLOWANCE	25153.96
TOTAL COST (BY UTILITY)	220453.43

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

PLANT CAPACITY SUMMARY II
--- BY CAPACITY TYPES ---

PROCESS UNIT	UNITS OF CAPACITY	MAXIMUM	MINIMUM	CALENDAR DAY BASIS	STREAM DAY BASIS	INVTMNT	PER UNIT OPERATIN COST
CRUDE UNIT	BBLS CAPACITY	100000.0	.	95000.0	100000.0	40700000.	0.41
VACUUM UNIT	BBLS CAPACITY	.	.	48570.1	51126.4	14322398.	0.52
DELAYED COKER	BBLS CAPACITY	.	.	23580.5	24821.6	55716982.	1.26
KEROSENE HYDROTREATER	BBLS CAPACITY	.	.	8511.3	8959.3	6214472.	0.57
DIESEL HYDROTREATER	BBLS CAPACITY	.	.	22439.5	23620.6	16898731.	0.51
GAS OIL HYDROTREATER	BBLS CAPACITY	.	.	10352.6	10897.5	12574929.	0.48
FLUID CATALYTIC CRACKER	BBLS CAPACITY	.	.	24701.4	26001.5	61154971.	0.60
ALKYLATION UNIT	BBLS CAPACITY	.	.	8422.9	8866.2	36039317.	2.32
NAPHTHA HYDROTREATER	BBLS CAPACITY	.	.	24100.8	25369.2	7104583.	0.49
C5/C6 PENEX UNIT	BBLS CAPACITY	.	.	3011.1	3169.5	7043340.	1.17
PLATFORMING UNIT	BBLS CAPACITY	.	.	21228.0	22345.2	33038460.	0.93
OFFSITE ALLOWANCE	BBLS CAPACITY	.	.	95000.0	95000.0	365402064.	0.26
HYDROGEN PLANT	TONS CAPACITY	.	.	71.8	79.8	.	.
AMINE TR- SULFUR PLANT	TONS CAPACITY	.	.	129.3	136.1	19345979.	84.82

TOTAL INVESTMENT

675556227.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY-1994
BASE CASE - CONVENTIONAL REFINERY
CRACKING REFINERYBASE MODEL

RAW MATERIAL PURCHASE SUMMARY

TIME 1 BASE MODEL

PURCHASED MATERIALS

		MAXIMUM	MINIMUM	QUANTITY BOUGHT	PRICE	TOTAL \$ PER DAY
IPL CRUDE	BBLS	47500.00	47500.00	47500.00	38.36	1822100.00
LLOYDMINISTER CRUDE	BBLS	47500.00	47500.00	47500.00	33.00	1567500.00
ISOBUTANE	BBLS	.	.	3986.50	28.70	114412.64
NORMAL BUTANE	BBLS	.	.	3066.82	24.70	75750.55

TOTAL RAW MATERIAL PURCHASE

3579763.19

RAW MATERIAL PURCHASE RESTRICTIONS:

Appendix D.3
Co-Processing Refinery Base Case

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASEOVERALL ECONOMIC SUMMARY

TIME 1 BASE MODEL

	DOLLARS PER DAY	\$MM PER YEAR
<u>GROSS REVENUES AT MARKET</u>		
CRACKING REFINERY	2222805.67	811.32
TOTAL...	2222805.67	811.32
<u>RAW MATERIAL PURCHASES</u>		
CRACKING REFINERY	1246994.47	455.15
TOTAL...	1246994.47	455.15
<u>PROCESS UNIT OPERATING COSTS</u>		
CRACKING REFINERY	194381.47	70.95
TOTAL...	194381.47	70.95
<u>UTILITY POOL COSTS</u>		
CRACKING REFINERY	36813.52	13.44
TOTAL...	36813.52	13.44
<u>TRANSPORTATION COSTS</u>		
CRACKING REFINERY	10.18	0.00
TOTAL...	10.18	0.00
<u>CAPITAL CHARGES</u>		
CRACKING REFINERY	698444.18	254.93
TOTAL...	698444.18	254.93
<u>OPTIMIZED PROFIT</u>	46161.84	16.85

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE

MARKET CENTER SUMMARY

PRODUCT SALESBASE MODEL

PRODUCT SALES

		MAXIMUM	MINIMUM	QUANTITY SOLD	PRICE \$	TOTAL \$ PER DAY
UNLEADED PRE. GASOLINE	BBLS	.	.	17870.22	24.76	442466.54
UNLEADED REG. GASOLINE	BBLS	.	.	35740.43	23.33	833824.33
JET FUEL	BBLS	.	.	8537.38	22.26	190041.98
.05% S NO.2 FUEL	BBLS	.	.	25512.28	21.22	541370.65
1.0% S NO.6 FUEL	BBLS	.	.	13614.03	13.49	183653.22
COKE	TONS	.	.	303.15	30.00	9094.62
SULFUR	TONS	.	.	223.54	100.00	22354.34
<u>TOTAL SALES REVENUE</u>						2222805.67

MARKET RESTRICTIONS:

SALES	PRODUCT SALES	FIXED	UNLEADED REG TO UNLEADED PRM
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

REFINERY ECONOMIC SUMMARY

	DOLLARS PER DAY	\$MM PER YEAR
SALES REVENUE	2222795.49	811.3204
COST OF RAW MATERIALS	1246994.47	455.1530
PROCESS UNIT OPERATING COST	194381.47	70.9492
UTILITY POOL COSTS	36813.52	13.4369
CAPITAL CHARGES	698444.18	254.9321
OPTIMIZED PROFIT	46161.84	16.8491

PROFITABILITY ANALYSIS

GROSS PROFIT	744606.03	271.7812
DEPRECIATION	226252.00	82.5820

PROFIT BEFORE TAXES		189.1992
ALLOWANCE FOR FEDRL INCOME TAX		64.3277
NET PROFIT AFTER TAXES		124.8715

CASH FLOW 207.4535

INVESTMENT

PLANT INVESTMENT	\$MM	894.4987
OFFSITE INVESTMENT		.
TOTAL INVESTMENT		894.4987

SIMPLE PAYOUT TIME (IN YEARS) 4.3

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

GROSS MARGIN SUMMARY

MARKET	PRODUCTS		MAXIMUM	MINIMUM	QUANTITY SOLD	PRICE	TOTAL \$ PER DAY
SALES	UNLEADED PRE. GASOLINE	BBLS	.	.	17870.22	24.76	442464.75
SALES	UNLEADED REG. GASOLINE	BBLS	.	.	35740.43	23.33	833820.75
SALES	JET FUEL	BBLS	.	.	8537.38	22.26	190041.12
SALES	.05% S NO.2 FUEL	BBLS	.	.	25512.28	21.22	541368.10
SALES	1.0% S NO.6 FUEL	BBLS	.	.	13614.03	13.49	183651.86
SALES	COKE	TONS	.	.	303.15	30.00	9094.59
SALES	SULFUR	TONS	.	.	223.54	100.00	22354.32
<u>TOTAL SALES REVENUE</u>							2222795.49
<u>RAW MATERIAL PURCHASED</u>							
			MAXIMUM	MINIMUM	QUANTITY PURCHASE	PRICE	TOTAL \$ PER DAY
	IPL CRUDE	BBLS	47500.00	47500.00	47500.00	13.47	639825.00
	LLOYDMINISTER CRUDE	BBLS	47500.00	47500.00	47500.00	10.48	497800.00
	ILL. NO. 6 COAL	TONS	.	.	2104.91	34.00	71566.97
	ISOBUTANE	BBLS	.	.	869.54	16.81	14616.99
	NORMAL BUTANE	BBLS	.	.	1504.58	15.41	23185.50
<u>TOTAL RAW MATERIAL COST</u>							1246994.47
<u>GROSS MARGIN</u>							975801.03

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

OPERATING COST SUMMARY

PROCESS UNIT BY COMBINED MODE

	ACTIVITY TONS/DAY	OPERATING COST/DAY
CRUDE UNIT	13247.71	7471.71
VACUUM UNIT	7505.35	4225.51
CO-PROCESSING UNIT	3789.90	100167.08
DELAYED COKER	535.51	1338.78
KEROSENE HYDROTREATER	1106.28	2780.09
DIESEL HYDROTREATER	3468.24	5809.29
GAS OIL HYDROTREATER	2113.25	2123.82
FLUID CATALYTIC CRACKER	4112.24	15708.75
CATALYTIC CONDENSATION	187.61	1988.68
HYDROGEN PLANT	111.61	12756.47
ALKYLATION UNIT	479.18	3216.24
NAPHTHA HYDROTREATER	3149.13	3967.91
C5/C6 PENEX UNIT	317.86	2325.13
PLATFORMING UNIT	2830.70	4953.73
AMINE TR- SULFUR PLANT	225.80	10471.99
OFFSITE ALLOWANCE	15352.62	15076.27
<u>TOTAL PROCESS UNIT OPERATING COST</u>		<u>194381.47</u>

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERY

.....BASE MODEL

UTILITY PURCHASES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
POWER CONSUMED	KWH		900653.06	0.04	36026.12
<u>TOTAL PURCHASED UTILITY COST</u>					36026.12

UTILITY SALES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
LOW PRESSURESTEAM	M LBS		4084.22	0.00	0.41
<u>TOTAL UTILITY SALES CREDIT</u>					0.41

UNPOOLED UTILITIES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
CATALYSTCONSUMED	DOLLARS		78846.21	1.00	78846.21
LABOR MAINTNCEINSURNCED	DOLLARS		115535.27	1.00	115535.27

TOTAL UNPOOLED UTILITY COST 194381.47

TOTAL UTILITY CONVERSION COST 787.81

TOTAL OPERATING COST 231195.00

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERY

.....BASE MODEL

MATERIAL BALANCE SUMMARY

RAW MATERIALS

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE	0.8265	47500.0	48.78	6234.62	40.04	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	48.78	7013.09	45.04	2559779.2
ILL. NO. 6 COAL	.	.	.	2104.91	13.52	768292.5
ISOBUTANE	0.5626	869.5	0.89	77.69	0.50	28356.7
NORMAL BUTANE	0.5835	1504.6	1.55	139.42	0.90	50888.6
TOTAL CHARGE	.	97374.1	100.00	15569.73	100.00	5682951.7

REFINERY PRODUCTS

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
UNLEADED PRE. GASOLINE	0.7579	17870.2	18.35	2150.86	13.81	785062.3
UNLEADED REG. GASOLINE	0.7383	35740.4	36.70	4190.57	26.92	1529556.5
JET FUEL	0.8149	8537.4	8.77	1104.85	7.10	403268.9
.05% S NO.2 FUEL	0.8559	25512.3	26.20	3467.74	22.27	1265726.8
1.0% S NO.6 FUEL	0.9850	13614.0	13.98	2129.59	13.68	777300.4
COKE	.	.	.	303.15	1.95	110651.2
SULFUR	.	.	.	223.54	1.44	81593.4
TOTAL PRODUCTS	.	101274.3	104.01	13570.30	87.16	4953159.5

STREAMS CONVERTED TO UTILITIES

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
UNSATURATE REFINERY GAS	.	.	.	208.78	1.34	76204.6
SATURATE REFINERY GAS	.	.	.	517.24	3.32	188792.6
PROPYLENE	0.5077	86.1	0.09	6.94	0.04	2533.7
PROPANE	0.5077	1717.8	1.76	138.50	0.89	50551.5
REFINERY FUEL OIL	.	.	.	13.32	0.09	4862.3
TOTAL STREAMS CONVERTED	.	1803.8	1.85	884.78	5.68	322944.7

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

MATERIAL BALANCE SUMMARY CONTINUED

STREAMS NOT UTILIZED

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
FCC COKE	.	.	.	219.59	1.41	80151.7
SOX EMISSIONS	.	.	.	2.26	0.02	826.6
LLOYDMINISTER CRUDE	0.9297	0.0	0.00	0.00	0.00	0.6
PLANT LOSSES	.	.	.	892.31	5.73	325693.8
TAR FROM HF ALKY UNIT	.	.	.	0.48	0.00	174.9
TOTAL NOT UTILIZED	.	0.0	0.00	1114.65	7.16	406847.6
<u>TOTAL PRODUCTS MADE</u>		103078.2	105.858	15569.73	100.000	5682951.7

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

PLANT CAPACITY SUMMARY

PROCESS UNIT	UNITS OF CAPACITY	MAXIMUM	MINIMUM	CALENDAR DAY BASIS	STREAM DAY BASIS	INVSTMNT
CRUDE UNIT	BBLS CAPACITY	100000.0	.	95000.0	100000.0	40680000.
VACUUM UNIT	BBLS CAPACITY	.	.	48570.1	51126.4	14322398.
CO-PROCESSING UNIT	BBLS CAPACITY	.	25683.0	23273.5	25745.0	182348089.
DELAYED COKER	BBLS CAPACITY	.	.	3026.5	3185.7	22463670.
KEROSENE HYDROTREATER	BBLS CAPACITY	.	.	8511.3	8959.3	6214466.
DIESEL HYDROTREATER	BBLS CAPACITY	.	.	25034.0	26351.6	18243841.
GAS OIL HYDROTREATER	BBLS CAPACITY	.	.	13811.5	14538.4	15386492.
FLUID CATALYTIC CRACKER	BBLS CAPACITY	.	10000.0	28192.7	29676.5	67262118.
CATALYTIC CONDENSATION	BBLS CAPACITY	.	.	2326.4	2448.8	6665561.
HYDROGEN PLANT	TONS CAPACITY	.	.	111.6	124.0	29679286.
ALKYLATION UNIT	BBLS CAPACITY	.	.	4288.4	4514.1	21358654.
NAPHTHA HYDROTREATER	BBLS CAPACITY	.	.	26809.6	28220.6	7654555.
C5/C6 PENEX UNIT	BBLS CAPACITY	.	.	3060.4	3221.5	7123952.
PLATFORMING UNIT	BBLS CAPACITY	.	.	24025.7	25290.2	36392833.
AMINE TR- SULFUR PLANT	TONS CAPACITY	.	.	225.8	237.7	28588701.
OFFSITE ALLOWANCE	BBLS CAPACITY	.	.	102824.0	102824.0	390114076.
<u>TOTAL INVESTMENT</u>						894498692.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
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CRACKING REFINERYBASE MODEL

PRODUCT BLENDING

UNLEADED PRE. GASOLINE

	BARRELS PER DAY	TONS PER DAY	SG 60/60	RESEARCH OCTANE NUMBER	MOTOR OCTANE NUMBER	VAPOR PRESSURE INDEX
ALKYLATE FROM BUTYLENE	2536.0	283.37	0.7036	94.50	94.50	40.10
FCC GASOLINE	5722.2	690.55	0.7599	92.90	81.30	88.00
NORMAL BUTANE	1503.7	139.34	0.5835	93.60	90.10	1120.00
98 RON PLATFORMATE	8108.3	1037.60	0.8058	98.00	87.00	24.00
BLEND	17870.2	2150.86	0.7579	95.50	86.50	139.00
MAXIMUM SPECIFICATION	.	.	.	.	.	139.00
MINIMUM SPECIFICATION	.	.	.	95.50	86.50	.

UNLEADED REG. GASOLINE

	BARRELS PER DAY	TONS PER DAY	SG 60/60	RESEARCH OCTANE NUMBER	MOTOR OCTANE NUMBER	VAPOR PRESSURE INDEX
ALKYLATE FROM BUTYLENE	1752.4	195.81	0.7036	94.50	94.50	40.10
LT ST RUN IPL	10.5	1.09	0.6542	60.30	58.80	159.00
LT ST RUN LLOYD	2531.4	262.99	0.6542	70.00	68.00	160.00
FCC GASOLINE	9533.6	1150.50	0.7599	92.90	81.30	88.00
NORMAL BUTANE	1946.8	180.40	0.5835	93.60	90.10	1120.00
C5/C6 PENEXATE	2983.0	305.08	0.6440	88.40	86.00	200.00
MOTOR FUEL POLYMER	1537.4	180.67	0.7400	94.00	83.00	25.50
98 RON PLATFORMATE	12709.3	1626.37	0.8058	98.00	87.00	24.00
CO-PROC LT GASO.	2736.1	287.65	0.6620	75.00	74.00	194.00
BLEND	35740.4	4190.57	0.7383	91.50	83.41	139.00
MAXIMUM SPECIFICATION	.	.	.	.	.	139.00
MINIMUM SPECIFICATION	.	.	.	91.50	82.50	.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
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CRACKING REFINERYBASE MODEL

PRODUCT BLENDING

JET FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	SMOKE POINT
HTD KEROSENE - HIGH SP	8537.4	1104.85	0.8149	0.01	21.00
BLEND	8537.4	1104.85	0.8149	0.01	21.00
MAXIMUM SPECIFICATION	.	.	.	0.10	.
MINIMUM SPECIFICATION	.	.	.	.	21.00

.05% S NO.2 FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	CETANE INDEX
HTD.COPROCESSING DIESEL	11664.4	1588.80	0.8577	0.05	32.00
HTD.SR DIESEL	13517.1	1834.73	0.8547	0.05	51.00
HTD. COKER DIESEL	330.8	44.21	0.8417	0.05	46.00
BLEND	25512.3	3467.74	0.8559	0.05	42.25
MAXIMUM SPECIFICATION	.	.	.	0.05	.
MINIMUM SPECIFICATION	.	.	.	.	40.00

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERY

.....BASE MODEL

PRODUCT BLENDING

1.0% S NO.6 FUEL

		BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	VISCOSTY INDEX @ 122 F
DIESEL	IPL	418.5	56.27	0.8468	0.17	14.85
VACUUM PITCH	IPL	4227.4	647.11	0.9639	0.54	45.01
VACUUM PITCH	LLOYD	1179.1	192.01	1.0254	5.10	45.73
LIGHT CYCLE OIL		4634.6	703.19	0.9554	0.34	15.40
SLURRY OIL		3154.4	531.00	1.0600	1.04	33.20
BLEND		13614.0	2129.59	0.9850	1.00	31.56
MAXIMUM SPECIFICATION		.	.	0.9850	1.00	37.50
MINIMUM SPECIFICATION		.	.	.	.	.

COKE

	BARRELS PER DAY	TONS PER DAY	SG 60/60
HIGH SULFUR COKE	.	303.15	.
BLEND	.	303.15	.
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERY

.....BASE MODEL

PRODUCT BLENDING

SULFUR

	BARRELS PER DAY	TONS PER DAY	SG 60/60
SULFUR	.	223.54	.
BLEND	.	223.54	.
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

1.5% S REFINERY FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	VISCOSTY INDEX @ 122 F
VACUUM PITCH LLOYD	9.3	1.51	1.0254	5.10	45.73
SLURRY OIL	70.2	11.82	1.0600	1.04	33.20
BLEND	79.5	13.33	1.0560	1.50	34.62
MAXIMUM SPECIFICATION	.	.	.	1.50	45.00
MINIMUM SPECIFICATION	.	.	.	.	.

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

UTILITY BLENDING SUMMARY

MM BTU FUEL CONSUMED

	TONS PER DAY	UTILITY UNITS/D	CONVERSN COST \$/D
UNSATURATE REFINERY GAS	208.78	8977.53	.
SATURATE REFINERY GAS	517.24	22758.56	.
PROPYLENE	6.94	291.55	.
PROPANE	138.50	6093.87	.
REFINERY FUEL OIL	13.32	506.22	.
TOTAL...	884.78	38627.72	.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

UTILITY CONVERSION SUMMARY

M LBS HIGH PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
FUEL CONSUMED	4923.81	4185.24	787.81
TOTAL...	4923.81	4185.24	787.81

M LBS MEDIUM PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
HIGH PRESSURESTEAM	2228.84	2228.84	.
TOTAL...	2228.84	2228.84	.

M GAL COOLING WATER

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
POWER CONSUMED	21014.20	26898.17	.
TOTAL...	21014.20	26898.17	.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

CRUDE UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE		0.8265	47500.0	50.00	6234.62	47.06	2275634.6
LLOYDMINISTER CRUDE		0.9297	47500.0	50.00	7013.09	52.94	2559778.6
TOTAL...		.	95000.0	100.00	13247.71	100.00	4835413.3

PRODUCTS

SATURATE REFINERY GAS		.	.	.	3.12	0.02	1137.8
PROPANE		0.5077	409.8	0.43	33.04	0.25	12060.9
ISO BUTANE		0.5626	263.4	0.28	23.54	0.18	8590.3
NORMAL BUTANE		0.5835	1130.3	1.19	104.74	0.79	38229.9
LT ST RUN	IPL	0.6542	2928.5	3.08	304.25	2.30	111051.0
LT ST RUN	LLOYD	0.6542	2531.4	2.66	262.99	1.99	95991.7
NAPHTHA	IPL	0.7389	10238.4	10.78	1201.41	9.07	438514.8
NAPHTHA	LLOYD	0.7459	6814.5	7.17	807.21	6.09	294630.5
KEROSENE	IPL	0.8109	5519.2	5.81	710.75	5.37	259422.3
KEROSENE	LLOYD	0.8324	2992.2	3.15	395.54	2.99	144371.5
DIESEL	IPL	0.8468	8929.2	9.40	1200.79	9.06	438287.2
DIESEL	LLOYD	0.8899	4917.8	5.18	695.00	5.25	253674.1
REDUCED CRUDE	IPL	0.9144	18414.3	19.38	2674.03	20.18	976019.7
REDUCED CRUDE	LLOYD	1.0086	30162.9	31.75	4831.32	36.47	1763431.5
TOTAL...		.	95251.9	100.27	13247.71	100.00	4835413.3

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERY

.....BASE MODEL

DETAILED MATERIAL BALANCE

VACUUM UNIT	CHARGE		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
REDUCED CRUDE	IPL		0.9144	18414.3	37.91	2674.03	35.63	976019.7
REDUCED CRUDE	LLOYD		1.0086	30162.9	62.09	4831.32	64.37	1763431.5
TOTAL...			.	48577.3	100.00	7505.35	100.00	2739451.2

PRODUCTS

VACUUM GAS OIL	IPL		0.8996	14187.7	29.21	2026.91	27.01	739822.9
VACUUM PITCH	IPL		0.9639	4227.4	8.70	647.11	8.62	236196.8
VACUUM GAS OIL	LLOYD		0.9365	5701.1	11.74	847.90	11.30	309482.2
VACUUM PITCH	LLOYD		1.0254	24461.9	50.36	3983.42	53.07	1453949.3
TOTAL...			.	48578.2	100.00	7505.35	100.00	2739451.2

CO-PROCESSING UNIT	CHARGE		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
ILLINOIS NO. 6 COAL			.	.	.	2104.91	34.73	768292.5
VACUUM PITCH	LLOYD		1.0254	23273.5	100.00	3789.90	62.54	1383313.9
HYDROGEN			.	.	.	165.62	2.73	60450.8
TOTAL...			.	23273.5	100.00	6060.43	100.00	2212057.2

PRODUCTS

HYDROGEN SULFIDE			.	.	.	176.99	2.92	64600.8
SATURATE REFINERY GAS			.	.	.	682.56	11.26	249134.8
CO-PROC LT GASO.			0.6620	2736.1	11.76	287.65	4.75	104993.5
CO-PROC NAPHTHA			0.7800	6131.4	26.34	759.50	12.53	277216.1
CO-PROC DIESEL			0.8838	11281.6	48.47	1583.42	26.13	577948.5
CO-PROC GAS OIL			0.9825	8086.0	34.74	1261.66	20.82	460505.2
CO-PROC PITCH			1.1142	3026.5	13.00	535.51	8.84	195462.3
PLANT LOSSES			.	.	.	773.14	12.76	282196.0
TOTAL...			.	31261.6	134.32	6060.43	100.00	2212057.2

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERY

.....BASE MODEL

DETAILED MATERIAL BALANCE

DELAYED COKER

CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
CO-PROC PITCH	1.1142	3026.5	100.00	535.51	100.00	195462.3
TOTAL...	.	3026.5	100.00	535.51	100.00	195462.3

PRODUCTS

HYDROGEN SULFIDE	.	.	.	2.57	0.48	938.2
UNSATURATE REFINERY GAS	.	.	.	53.34	9.96	19468.0
PROPYLENE	0.5220	105.3	3.48	8.73	1.63	3186.0
PROPANE	0.5077	182.0	6.01	14.67	2.74	5355.7
BUTYLENE	0.6100	120.5	3.98	11.67	2.18	4261.1
ISO BUTANE	0.5626	27.0	0.89	2.41	0.45	879.6
NORMAL BUTANE	0.5835	130.6	4.32	12.10	2.26	4417.4
COKER NAPHTHA	0.7424	660.4	21.82	77.86	14.54	28420.2
COKER DISTILLATE	0.8961	318.4	10.52	45.30	8.46	16536.1
COKER GAS OIL	0.9639	24.1	0.80	3.70	0.69	1348.7
HIGH SULFUR COKE	.	.	.	303.15	56.61	110651.2
TOTAL...	.	1568.3	51.82	535.51	100.00	195462.3

KEROSENE HYDROTREATER

CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
KEROSENE IPL	0.8109	5519.2	64.85	710.75	64.17	259422.3
KEROSENE LLOYD	0.8324	2992.2	35.15	395.54	35.71	144371.5
HYDROGEN	.	.	.	1.33	0.12	484.6
TOTAL...	.	8511.3	100.00	1107.61	100.00	404278.4

PRODUCTS

HYDROGEN SULFIDE	.	.	.	2.10	0.19	767.2
SATURATE REFINERY GAS	.	.	.	0.66	0.06	242.3
HTD KEROSENE - HIGH SP	0.8149	8537.4	100.31	1104.85	99.75	403268.9
TOTAL...	.	8537.4	100.31	1107.61	100.00	404278.4

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

DIESEL HYDROTREATER	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
DIESEL	IPL	0.8468	8510.7	34.00	1144.51	32.88	417747.3
DIESEL	LLOYD	0.8899	4917.8	19.65	695.00	19.97	253674.1
CO-PROC DIESEL		0.8838	11281.6	45.07	1583.42	45.49	577948.5
COKER DISTILLATE		0.8961	318.4	1.27	45.30	1.30	16536.1
HYDROGEN		.	.	.	12.52	0.36	4571.3
TOTAL...		.	25028.4	100.00	3480.76	100.00	1270477.3

PRODUCTS

HYDROGEN SULFIDE	.	.	.	9.16	0.26	3343.6
SATURATE REFINERY GAS	.	.	.	3.85	0.11	1407.0
HTD. COKER DIESEL	0.8417	330.8	1.32	44.21	1.27	16137.6
HTD.SR DIESEL	0.8547	13517.1	54.01	1834.73	52.71	669675.7
HTD.COPROCESSING DIESEL	0.8577	11664.4	46.60	1588.80	45.65	579913.6
TOTAL...	.	25512.3	101.93	3480.76	100.00	1270477.3

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

GAS OIL HYDROTREATER CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
VACUUM GAS OIL LLOYD	0.9365	5701.1	41.28	847.90	39.91	309482.2
COKER GAS OIL	0.9639	24.1	0.17	3.70	0.17	1348.7
CO-PROC GAS OIL	0.9825	8086.0	58.55	1261.66	59.38	460505.2
HYDROGEN	.	.	.	11.50	0.54	4198.0
TOTAL...	.	13811.3	100.00	2124.75	100.00	775534.1

PRODUCTS

HYDROGEN SULFIDE	.	.	.	27.50	1.29	10037.2
SATURATE REFINERY GAS	.	.	.	11.92	0.56	4352.5
HTD MIXED GAS OILS	0.9202	5730.2	41.49	837.38	39.41	305644.6
HTD COKER DIESEL	0.8994	25.0	0.18	3.57	0.17	1303.5
HTD CO-PROCESSING GAS OI	0.9498	8249.8	59.73	1244.37	58.57	454196.3
TOTAL...	.	14005.0	101.40	2124.75	100.00	775534.1

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

FLUID CATALYTIC CRACKER CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HTD MIXED GAS OILS	0.9202	5730.2	20.33	837.38	20.36	305644.6
VACUUM GAS OIL IPL	0.8996	14187.7	50.32	2026.91	49.29	739822.9
HTD CO-PROCESSING GAS OI	0.9498	8249.8	29.26	1244.37	30.26	454196.3
HTD COKER DIESEL	0.8994	25.0	0.09	3.57	0.09	1303.5
TOTAL...	.	28192.7	100.00	4112.24	100.00	1500967.4

PRODUCTS

HYDROGEN SULFIDE	.	.	.	3.70	0.09	1350.9
UNSATURATE REFINERY GAS	.	.	.	155.44	3.78	56736.6
PROPYLENE	0.5220	2157.9	7.65	178.88	4.35	65292.1
PROPANE	0.5077	841.6	2.99	67.85	1.65	24766.0
BUTYLENE	0.6100	2300.8	8.16	222.88	5.42	81352.4
ISO BUTANE	0.5626	1509.7	5.35	134.88	3.28	49231.7
NORMAL BUTANE	0.5835	452.7	1.61	41.94	1.02	15309.9
FCC GASOLINE	0.7599	15255.8	54.11	1841.05	44.77	671983.1
LIGHT CYCLE OIL	0.9554	4634.6	16.44	703.19	17.10	256665.4
SLURRY OIL	1.0600	3224.6	11.44	542.82	13.20	198127.7
FCC COKE	.	.	.	219.59	5.34	80151.7
TOTAL...	.	30377.6	107.75	4112.24	100.00	1500967.4

CATALYTIC CONDENSATION CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
PROPYLENE	0.5220	2263.2	100.00	187.61	100.00	68478.1
TOTAL...	.	2263.2	100.00	187.61	100.00	68478.1

PRODUCTS

PROPYLENE	0.5077	86.1	3.80	6.94	3.70	2533.7
MOTOR FUEL POLYMER	0.7400	1537.4	67.93	180.67	96.30	65944.4
TOTAL...	.	1623.5	71.73	187.61	100.00	68478.1

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

HYDROGEN PLANT

CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
SATURATE REFINERY GAS	.	.	.	230.78	100.00	84233.7
TOTAL...	.	.	.	230.78	100.00	84233.7

PRODUCTS

HYDROGEN	.	.	.	111.61	48.36	40735.9
PLANT LOSSES	.	.	.	119.17	51.64	43497.8
TOTAL...	.	.	.	230.78	100.00	84233.7

REFINERY FUEL BLENDING CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
1.5% S REFINERY FUEL	1.0560	79.5	100.00	13.33	100.00	4864.8
TOTAL...	.	79.5	100.00	13.33	100.00	4864.8

PRODUCTS

REFINERY FUEL OIL	.	.	.	13.32	99.95	4862.3
SOX EMISSIONS	.	.	.	0.01	0.05	2.4
TOTAL...	.	.	.	13.33	100.00	4864.8

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
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DETAILED MATERIAL BALANCE

ALKYLATION UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
	BUTYLENE	0.6100	2421.3	46.43	234.56	48.45	85613.5
	ISO BUTANE	0.5626	2793.2	53.57	249.56	51.55	91087.9
	TOTAL...	.	5214.4	100.00	484.11	100.00	176701.4

PRODUCTS

	NORMAL BUTANE	0.5835	48.1	0.92	4.46	0.92	1626.6
	ALKYLATE FROM BUTYLENE	0.7036	4288.4	82.24	479.18	98.98	174899.9
	TAR FROM HF ALKY UNIT	.	.	.	0.48	0.10	174.9
	TOTAL...	.	4336.5	83.16	484.11	100.00	176701.4

NAPHTHA HYDROTREATER CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
	NAPHTHA IPL	0.7389	10238.4	38.26	1201.41	38.09	438514.8
	NAPHTHA LLOYD	0.7459	6814.5	25.46	807.21	25.59	294630.5
	LT ST RUN IPL	0.6542	2918.0	10.90	303.16	9.61	110652.6
	COKER NAPHTHA	0.7424	660.4	2.47	77.86	2.47	28420.2
	CO-PROC NAPHTHA	0.7800	6131.4	22.91	759.50	24.08	277216.1
	HYDROGEN	.	.	.	5.30	0.17	1936.3
	TOTAL...	.	26762.7	100.00	3154.44	100.00	1151370.5

PRODUCTS

	HYDROGEN SULFIDE	.	.	.	3.78	0.12	1379.7
	SATURATE REFINERY GAS	.	.	.	0.88	0.03	321.6
	PROPANE	0.5077	7.0	0.03	0.57	0.02	206.6
	NORMAL BUTANE	0.5835	7.0	0.03	0.65	0.02	237.2
	LNA	0.6540	3060.4	11.44	317.86	10.08	116018.1
	HHN	0.7247	24595.9	91.90	2830.70	89.74	1033207.3
	TOTAL...	.	27670.4	103.39	3154.44	100.00	1151370.5

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

C5/C6 PENEX UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
LNA		0.6540	3060.4	100.00	317.86	99.13	116018.1
HYDROGEN		.	.	.	2.80	0.87	1021.0
TOTAL...		.	3060.4	100.00	320.65	100.00	117039.0

PRODUCTS

SATURATE REFINERY GAS	.	.	.	15.58	4.86	5684.9
C5/C6 PENEXATE	0.6440	2983.0	97.47	305.08	95.14	111354.2
TOTAL...	.	2983.0	97.47	320.65	100.00	117039.0

PLATFORMING UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HHN		0.7247	24595.9	100.00	2830.70	100.00	1033207.3
TOTAL...		.	24595.9	100.00	2830.70	100.00	1033207.3

PRODUCTS

HYDROGEN	.	.	.	87.47	3.09	31926.1
SATURATE REFINERY GAS	.	.	.	29.44	1.04	10745.4
PROPANE	0.5077	277.4	1.13	22.36	0.79	8162.3
ISO BUTANE	0.5626	123.6	0.50	11.04	0.39	4029.5
NORMAL BUTANE	0.5835	177.2	0.72	16.42	0.58	5992.6
98 RON PLATFORMATE	0.8058	20817.6	84.64	2663.98	94.11	972351.4
TOTAL...	.	21395.7	86.99	2830.70	100.00	1033207.3

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

DETAILED MATERIAL BALANCE

AMINE TR- SULFUR PLANT CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HYDROGEN SULFIDE	.	.	.	225.80	100.00	82417.5
TOTAL...	.	.	.	225.80	100.00	82417.5

PRODUCTS

SULFUR	.	.	.	223.54	99.00	81593.4
SOX EMISSIONS	.	.	.	2.26	1.00	824.2
TOTAL...	.	.	.	225.80	100.00	82417.5

OFFSITE ALLOWANCE CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE	0.8265	47500.0	50.00	6234.62	40.61	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	50.00	7013.09	45.68	2559779.2
ILLINOIS NO. 6 COAL	.	.	.	2104.91	13.71	768292.5
TOTAL...	.	95000.0	100.00	15352.62	100.00	5603706.4

PRODUCTS

IPL CRUDE	0.8265	47500.0	50.00	6234.62	40.61	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	50.00	7013.09	45.68	2559779.2
ILLINOIS NO. 6 COAL	.	.	.	2104.91	13.71	768292.5
TOTAL...	.	95000.0	100.00	15352.62	100.00	5603706.4

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

UTILITY CONSUMPTION SUMMARY

PROCESS UNITS	COMBINED MODE	LABOR MAINTNCE INSURANCE DOLLARS	POWER CONSUMED KWH	FUEL CONSUMED MM BTU	HIGH PRESSURE STEAM M LBS	MEDIUM PRESSURE STEAM M LBS	LOW PRESSURE STEAM M LBS	COOLING WATER M GAL	CATALYST CONSUMED DOLLARS
CRUDE UNIT		7471.71	58952.30	6557.62	.	543.16	.	168.25	.
VACUUM UNIT		4225.51	13029.28	4000.35	.	825.59	112.58	5208.71	.
CO-PROCESSING UNIT		27363.09	557115.45	4495.58	594.26	.	-3937.71	6158.21	72804.00
DELAYED COKER		1338.78	7671.22	642.62	.	-24.10	9.21	249.28	.
KEROSENE HYDROTREATER		2780.09	7821.43	297.59	66.38	.	.	27.66	.
DIESEL HYDROTREATER		5809.29	30289.39	932.96	208.09	.	.	86.71	.
GAS OIL HYDROTREATER		2123.82	23249.36	568.46	126.79	.	.	52.83	.
FLUID CATALYTIC CRACKER		11962.50	33921.86	.	2204.16	-1163.76	-2204.16	2829.22	3746.25
CATALYTIC CONDENSATION		1500.89	7035.42	600.36	82.55	.	.	422.13	487.79
HYDROGEN PLANT		12756.47	26450.42	6266.63	-550.21	.	.	97.82	.
ALKYLATION UNIT		2944.07	8471.39	959.31	.	.	383.34	7939.98	272.17
NAPHTHA HYDROTREATER		3967.91	17345.52	2114.01	.	.	.	1648.89	.
C5/C6 PENEX UNIT		2005.68	4601.31	.	.	.	197.71	214.87	319.45
PLATFORMING UNIT		3962.99	35185.66	5604.80	-775.61	.	.	1395.54	990.75
AMINE TR- SULFUR PLANT		10246.19	23335.90	663.63	.	-331.70	1354.81	398.09	225.80
OFFSITE ALLOWANCE		15076.27	25162.94	.	.	2379.66	.	.	.
TOTAL...		115535.27	879638.86	33703.91	1956.41	2228.84	-4084.22	26898.17	78846.21

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

UTILITY CONSUMPTION SUMMARY

PROCESS UNITS

COMBINED MODE

M

CRUDE UNIT	.
VACUUM UNIT	.
CO-PROCESSING UNIT	.
DELAYED COKER	.
KEROSENE HYDROTREATER	.
DIESEL HYDROTREATER	.
GAS OIL HYDROTREATER	.
FLUID CATALYTIC CRACKER	.
CATALYTIC CONDENSATION	.
HYDROGEN PLANT	.
ALKYLATION UNIT	.
NAPHTHA HYDROTREATER	.
C5/C6 PENEX UNIT	.
PLATFORMING UNIT	.
AMINE TR- SULFUR PLANT	.
OFFSITE ALLOWANCE	.
TOTAL...	.

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UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

UTILITY COST

COST

LABOR	MAINTNCE	INSURANCE	DOLLARS	1.00
CATALYST	CONSUMED		DOLLARS	1.00
POWER	CONSUMED		KWH	0.05
FUEL	CONSUMED		MM BTU	2.25
HIGH	PRESSURE	STEAM	M LBS	3.65
MEDIUM	PRESSURE	STEAM	M LBS	3.20
LOW	PRESSURE	STEAM	M LBS	3.15
COOLING	WATER		M GAL	0.10

UTILITY CONSUMPTION COST SUMMARY

	LABOR MAINTNCE INSURANCE DOLLARS	CATALYST CONSUMED DOLLARS	POWER CONSUMED DOLLARS	FUEL CONSUMED DOLLARS	HIGH PRESSURE STEAM DOLLARS	MEDIUM PRESSURE STEAM DOLLARS	LOW PRESSURE STEAM DOLLARS	COOLING WATER DOLLARS
CRUDE UNIT	7471.71	.	2947.61	14754.63	.	1738.10	.	16.82
VACUUM UNIT	4225.51	.	651.46	9000.79	.	2641.88	354.63	520.87
CO-PROCESSING UNIT	27363.09	72804.00	27855.77	10115.06	2169.04	.	-12403.78	615.82
DELAYED COKER	1338.78	.	383.56	1445.89	.	-77.11	29.01	24.93
KEROSENE HYDROTREATER	2780.09	.	391.07	669.58	242.28	.	.	2.77
DIESEL HYDROTREATER	5809.29	.	1514.47	2099.15	759.54	.	.	8.67
GAS OIL HYDROTREATER	2123.82	.	1162.47	1279.04	462.80	.	.	5.28
FLUID CATALYTIC CRACKER	11962.50	3746.25	1696.09	.	8045.19	-3724.04	-6943.10	282.92
CATALYTIC CONDENSATION	1500.89	487.79	351.77	1350.80	301.30	.	.	42.21
HYDROGEN PLANT	12756.47	.	1322.52	14099.92	-2008.28	.	.	9.78
ALKYLATION UNIT	2944.07	272.17	423.57	2158.46	.	.	1207.53	794.00
NAPHTHA HYDROTREATER	3967.91	.	867.28	4756.53	.	.	.	164.89
C5/C6 PENEX UNIT	2005.68	319.45	230.07	.	.	.	622.78	21.49
PLATFORMING UNIT	3962.99	990.75	1759.28	12610.79	-2830.99	.	.	139.55
AMINE TR- SULFUR PLANT	10246.19	225.80	1166.80	1493.17	.	-1061.45	4267.65	39.81
OFFSITE ALLOWANCE	15076.27	.	1258.15	.	.	7614.90	.	.
TOTAL COST (BY UTILITY)	115535.27	78846.21	43981.94	75833.80	7140.88	7132.28	-12865.29	2689.82

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UTILITY CONSUMPTION COST SUMMARY

TOTAL
COST
(BY
UNIT)

CRUDE UNIT	26928.88
VACUUM UNIT	17395.14
CO-PROCESSING UNIT	128518.99
DELAYED COKER	3145.06
KEROSENE HYDROTREATER	4085.79
DIESEL HYDROTREATER	10191.13
GAS OIL HYDROTREATER	5033.41
FLUID CATALYTIC CRACKER	15065.81
CATALYTIC CONDENSATION	4034.77
HYDROGEN PLANT	26180.41
ALKYLATION UNIT	7799.79
NAPHTHA HYDROTREATER	9756.61
C5/C6 PENEX UNIT	3199.46
PLATFORMING UNIT	16632.37
AMINE TR- SULFUR PLANT	16377.97
OFFSITE ALLOWANCE	23949.32
TOTAL COST (BY UTILITY)	318294.90

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

PLANT CAPACITY SUMMARY II

--- BY CAPACITY TYPES ---

PROCESS UNIT	UNITS OF CAPACITY	MAXIMUM	MINIMUM	CALENDAR DAY BASIS	STREAM DAY BASIS	INVTMNT	PER UNIT OPERATIN COST
CRUDE UNIT	BBLS CAPACITY	100000.0	.	95000.0	100000.0	40680000.	0.28
VACUUM UNIT	BBLS CAPACITY	.	.	48570.1	51126.4	14322398.	0.36
CO-PROCESSING UNIT	BBLS CAPACITY	.	25683.0	23273.5	25745.0	182348089.	5.52
DELAYED COKER	BBLS CAPACITY	.	.	3026.5	3185.7	22463670.	1.04
KEROSENE HYDROTREATER	BBLS CAPACITY	.	.	8511.3	8959.3	6214466.	0.48
DIESEL HYDROTREATER	BBLS CAPACITY	.	.	25034.0	26351.6	18243841.	0.41
GAS OIL HYDROTREATER	BBLS CAPACITY	.	.	13811.5	14538.4	15386492.	0.36
FLUID CATALYTIC CRACKER	BBLS CAPACITY	.	10000.0	28192.7	29676.5	67262118.	0.53
CATALYTIC CONDENSATION	BBLS CAPACITY	.	.	2326.4	2448.8	6665561.	1.73
ALKYLATION UNIT	BBLS CAPACITY	.	.	4288.4	4514.1	21358654.	1.82
NAPHTHA HYDROTREATER	BBLS CAPACITY	.	.	26809.6	28220.6	7654555.	0.36
C5/C6 PENEX UNIT	BBLS CAPACITY	.	.	3060.4	3221.5	7123952.	1.05
PLATFORMING UNIT	BBLS CAPACITY	.	.	24025.7	25290.2	36392833.	0.69
OFFSITE ALLOWANCE	BBLS CAPACITY	.	.	102824.0	102824.0	390114076.	0.23
HYDROGEN PLANT	TONS CAPACITY	.	.	111.6	124.0	29679286.	234.58
AMINE TR- SULFUR PLANT	TONS CAPACITY	.	.	225.8	237.7	28588701.	72.53

TOTAL INVESTMENT

894498692.

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERY BASE MODEL

RAW MATERIAL PURCHASE SUMMARY

TIME 1 BASE MODEL

PURCHASED MATERIALS

		MAXIMUM	MINIMUM	QUANTITY BOUGHT	PRICE	TOTAL \$ PER DAY
IPL CRUDE	BBLs	47500.00	47500.00	47500.00	13.47	639825.00
LLOYDMINISTER CRUDE	BBLs	47500.00	47500.00	47500.00	10.48	497800.00
ILL. NO. 6 COAL	TONS	.	.	2104.91	34.00	71566.97
ISOBUTANE	BBLs	.	.	869.54	16.81	14616.99
NORMAL BUTANE	BBLs	.	.	1504.58	15.41	23185.50

TOTAL RAW MATERIAL PURCHASE

1246994.47

RAW MATERIAL PURCHASE RESTRICTIONS:

Appendix D.4
Co-Processing Refinery High Crude Price Case

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASEOVERALL ECONOMIC SUMMARY

TIME 1 BASE MODEL

	DOLLARS PER DAY	\$MM PER YEAR
<u>GROSS REVENUES AT MARKET</u>		
CRACKING REFINERY	4561120.46	1664.81
TOTAL...	4561120.46	1664.81
<u>RAW MATERIAL PURCHASES</u>		
CRACKING REFINERY	3625034.05	1323.14
TOTAL...	3625034.05	1323.14
<u>PROCESS UNIT OPERATING COSTS</u>		
CRACKING REFINERY	203181.50	74.16
TOTAL...	203181.50	74.16
<u>UTILITY POOL COSTS</u>		
CRACKING REFINERY	38534.98	14.07
TOTAL...	38534.98	14.07
<u>TRANSPORTATION COSTS</u>		
CRACKING REFINERY	10.50	0.00
TOTAL...	10.50	0.00
<u>CAPITAL CHARGES</u>		
CRACKING REFINERY	710983.63	259.51
TOTAL...	710983.63	259.51
<u>OPTIMIZED PROFIT</u>	-16624.20	-6.07

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE

MARKET CENTER SUMMARY

PRODUCT SALESBASE MODEL

PRODUCT SALES

		MAXIMUM	MINIMUM	QUANTITY SOLD	PRICE \$	TOTAL \$ PER DAY
UNLEADED PRE. GASOLINE	BBLS	.	.	19039.22	51.95	989087.26
UNLEADED REG. GASOLINE	BBLS	.	.	38078.43	44.10	1679258.96
JET FUEL	BBLS	.	.	8537.38	42.92	366424.15
.05% S NO.2 FUEL	BBLS	.	.	26416.16	40.74	1076194.33
1.0% S NO.6 FUEL	BBLS	.	.	12332.03	33.64	414849.51
COKE	TONS	.	.	318.63	30.00	9559.01
SULFUR	TONS	.	.	234.07	110.00	25747.23
<u>TOTAL SALES REVENUE</u>						4561120.46

MARKET RESTRICTIONS:

SALES	PRODUCT SALES	FIXED	UNLEADED REG TO UNLEADED PRM
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

REFINERY ECONOMIC SUMMARY

	DOLLARS PER DAY	\$MM PER YEAR
SALES REVENUE	4561109.96	1664.8051
COST OF RAW MATERIALS	3625034.05	1323.1374
PROCESS UNIT OPERATING COST	203181.50	74.1612
UTILITY POOL COSTS	38534.98	14.0653
CAPITAL CHARGES	710983.63	259.5090
OPTIMIZED PROFIT	-16624.20	-6.0678

PROFITABILITY ANALYSIS

GROSS PROFIT	694359.43	253.4412
DEPRECIATION	226252.00	82.5820

PROFIT BEFORE TAXES		170.8592
ALLOWANCE FOR FEDRL INCOME TAX		58.0921
NET PROFIT AFTER TAXES		112.7671

<u>CASH FLOW</u>		195.3491
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INVESTMENT

	\$MM
PLANT INVESTMENT	910.5580
OFFSITE INVESTMENT	.
TOTAL INVESTMENT	910.5580

<u>SIMPLE PAYOUT TIME (IN YEARS)</u>	4.7
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UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

GROSS MARGIN SUMMARY

MARKET	PRODUCTS		MAXIMUM	MINIMUM	QUANTITY SOLD	PRICE	TOTAL \$ PER DAY
SALES	UNLEADED PRE. GASOLINE	BBLS	.	.	19039.22	51.95	989085.35
SALES	UNLEADED REG. GASOLINE	BBLS	.	.	38078.43	44.10	1679255.16
SALES	JET FUEL	BBLS	.	.	8537.38	42.92	366423.30
SALES	.05% S NO.2 FUEL	BBLS	.	.	26416.16	40.74	1076191.69
SALES	1.0% S NO.6 FUEL	BBLS	.	.	12332.03	33.64	414848.28
SALES	COKE	TONS	.	.	318.63	30.00	9558.98
SALES	SULFUR	TONS	.	.	234.07	110.00	25747.21

TOTAL SALES REVENUE

4561109.96

RAW MATERIAL PURCHASED

		MAXIMUM	MINIMUM	QUANTITY PURCHASE	PRICE	TOTAL \$ PER DAY
IPL CRUDE	BBLS	47500.00	47500.00	47500.00	38.36	1822100.00
LLOYDMINISTER CRUDE	BBLS	47500.00	47500.00	47500.00	33.00	1567500.00
ILL. NO. 6 COAL	TONS	.	.	2212.39	35.00	77433.75
ISOBUTANE	BBLS	.	.	3999.09	28.70	114773.89
NORMAL BUTANE	BBLS	.	.	1750.06	24.70	43226.41

TOTAL RAW MATERIAL COST

3625034.05

GROSS MARGIN

936075.91

UOP PROCESS DIVISION
DOE COPROCESSING STUDY
COPROCESSING CASE
CRACKING REFINERYBASE MODEL

OPERATING COST SUMMARY

PROCESS UNIT BY COMBINED MODE

ACTIVITY	OPERATING
TONS/DAY	COST/DAY
CRUDE UNIT	13247.71 7471.71
VACUUM UNIT	7505.35 4225.51
CO-PROCESSING UNIT	3983.42 105281.86
DELAYED COKER	562.86 1407.14
KEROSENE HYDROTREATER	1106.28 2780.09
DIESEL HYDROTREATER	3591.05 6015.02
GAS OIL HYDROTREATER	2177.86 2188.75
FLUID CATALYTIC CRACKER	4175.96 15952.18
HYDROGEN PLANT	120.33 13753.27
ALKYLATION UNIT	936.67 6286.94
NAPHTHA HYDROTREATER	3226.12 4064.91
C5/C6 PENEX UNIT	352.83 2580.93
PLATFORMING UNIT	2872.72 5026.46
AMINE TR- SULFUR PLANT	236.43 10964.92
OFFSITE ALLOWANCE	15460.10 15181.82

TOTAL PROCESS UNIT OPERATING COST

203181.50

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UTILITY PURCHASES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
POWER CONSUMED	KWH		944060.84	0.04	37762.43
<u>TOTAL PURCHASED UTILITY COST</u>					37762.43

UTILITY SALES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
LOW PRESSURESTEAM	M LBS		3867.45	0.00	0.39
<u>TOTAL UTILITY SALES CREDIT</u>					0.39

UNPOOLED UTILITIES

UTILITY	UNIT		UNITS PER DAY	DOLLARS PER UNIT	DOLLARS PER DAY
CATALYSTCONSUMED	DOLLARS		82453.55	1.00	82453.55
LABOR MAINTNCEINSURNCEDOLLARS			120727.95	1.00	120727.95
<u>TOTAL UNPOOLED UTILITY COST</u>					203181.50
<u>TOTAL UTILITY CONVERSION COST</u>					772.93
<u>TOTAL OPERATING COST</u>					241716.48

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MATERIAL BALANCE SUMMARY

RAW MATERIALS

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE	0.8265	47500.0	47.15	6234.62	39.02	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	47.15	7013.09	43.89	2559779.2
ILL. NO. 6 COAL	.	.	.	2212.39	13.85	807523.4
ISOBUTANE	0.5626	3999.1	3.97	357.30	2.24	130414.8
NORMAL BUTANE	0.5835	1750.1	1.74	162.17	1.02	59191.5
TOTAL CHARGE	.	100749.1	100.00	15979.57	100.00	5832543.6

REFINERY PRODUCTS

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
UNLEADED PRE. GASOLINE	0.7514	19039.2	18.90	2271.86	14.22	829229.8
UNLEADED REG. GASOLINE	0.7364	38078.4	37.80	4452.98	27.87	1625337.1
JET FUEL	0.8149	8537.4	8.47	1104.85	6.91	403268.9
.05% S NO.2 FUEL	0.8559	26416.2	26.22	3590.68	22.47	1310598.1
1.0% S NO.6 FUEL	0.9850	12332.0	12.24	1929.05	12.07	704104.1
COKE	.	.	.	318.63	1.99	116301.3
SULFUR	.	.	.	234.07	1.47	85434.0
TOTAL PRODUCTS	.	104403.2	103.63	13902.12	87.00	5074273.4

STREAMS CONVERTED TO UTILITIES

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
UNSATURATE REFINERY GAS	.	.	.	213.91	1.34	78077.9
SATURATE REFINERY GAS	.	.	.	536.70	3.36	195895.0
PROPANE	0.5077	1973.6	1.96	159.13	1.00	58081.0
TOTAL STREAMS CONVERTED	.	1973.6	1.96	909.74	5.69	332053.9

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MATERIAL BALANCE SUMMARY CONTINUED

STREAMS NOT UTILIZED

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
FCC COKE	.	.	.	223.00	1.40	81393.7
SOX EMISSIONS	.	.	.	2.36	0.02	863.0
LLOYDMINISTER CRUDE	0.9297	0.0	0.00	0.00	0.00	0.6
PLANT LOSSES	.	.	.	941.10	5.89	343502.4
TAR FROM HF ALKY UNIT	.	.	.	1.25	0.01	456.7
TOTAL NOT UTILIZED	.	0.0	0.00	1167.72	7.31	426216.3

<u>TOTAL PRODUCTS MADE</u>		106376.8	105.586	15979.57	100.000	5832543.6
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DOE COPROCESSING STUDY
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 PLANT CAPACITY SUMMARY

PROCESS UNIT	UNITS OF CAPACITY	MAXIMUM	MINIMUM	CALENDAR DAY BASIS	STREAM DAY BASIS	INVTMNT
CRUDE UNIT	BBLS CAPACITY	100000.0	.	95000.0	100000.0	40680000.
VACUUM UNIT	BBLS CAPACITY	.	.	48570.1	51126.4	14322398.
CO-PROCESSING UNIT	BBLS CAPACITY	.	25683.0	24461.9	27059.6	183564703.
DELAYED COKER	BBLS CAPACITY	.	.	3181.0	3348.4	23202738.
KEROSENE HYDROTREATER	BBLS CAPACITY	.	.	8511.3	8959.3	6214466.
DIESEL HYDROTREATER	BBLS CAPACITY	.	.	25921.5	27285.8	18694196.
GAS OIL HYDROTREATER	BBLS CAPACITY	.	.	14225.6	14974.3	15708011.
FLUID CATALYTIC CRACKER	BBLS CAPACITY	.	10000.0	28615.3	30121.3	67986422.
HYDROGEN PLANT	TONS CAPACITY	.	.	120.3	133.7	31166819.
ALKYLATION UNIT	BBLS CAPACITY	.	.	8382.8	8824.0	35906187.
NAPHTHA HYDROTREATER	BBLS CAPACITY	.	.	27488.2	28934.9	7789679.
C5/C6 PENEX UNIT	BBLS CAPACITY	.	.	3397.1	3575.9	7663919.
PLATFORMING UNIT	BBLS CAPACITY	.	.	24382.3	25665.6	36808091.
AMINE TR- SULFUR PLANT	TONS CAPACITY	.	.	236.4	248.9	29524162.
OFFSITE ALLOWANCE	BBLS CAPACITY	.	.	103223.5	103223.5	391326199.
<u>TOTAL INVESTMENT</u>						910557988.

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PRODUCT BLENDING

UNLEADED PRE. GASOLINE

	BARRELS PER DAY	TONS PER DAY	SG 60/60	RESEARCH OCTANE NUMBER	MOTOR OCTANE NUMBER	VAPOR PRESSURE INDEX
ALKYLATE FROM BUTYLENE	4362.5	487.45	0.7036	94.50	94.50	40.10
FCC GASOLINE	4817.7	581.39	0.7599	92.90	81.30	88.00
NORMAL BUTANE	1652.3	153.11	0.5835	93.60	90.10	1120.00
94 RON PLATFORMATE	122.2	15.35	0.7909	94.00	83.70	24.00
98 RON PLATFORMATE	8084.5	1034.56	0.8058	98.00	87.00	24.00
BLEND	19039.2	2271.86	0.7514	95.50	87.52	139.00
MAXIMUM SPECIFICATION	.	.	.	.	.	139.00
MINIMUM SPECIFICATION	.	.	.	95.50	86.50	.

UNLEADED REG. GASOLINE

	BARRELS PER DAY	TONS PER DAY	SG 60/60	RESEARCH OCTANE NUMBER	MOTOR OCTANE NUMBER	VAPOR PRESSURE INDEX
ALKYLATE FROM PROPYLENE	4020.3	449.22	0.7036	92.00	92.00	40.10
LT ST RUN LLOYD	2212.5	229.86	0.6542	70.00	68.00	160.00
FCC GASOLINE	10674.5	1288.18	0.7599	92.90	81.30	88.00
NORMAL BUTANE	2060.7	190.96	0.5835	93.60	90.10	1120.00
C5/C6 PENEXATE	3311.2	338.64	0.6440	88.40	86.00	200.00
98 RON PLATFORMATE	12923.4	1653.77	0.8058	98.00	87.00	24.00
CO-PROC LT GASO.	2875.9	302.34	0.6620	75.00	74.00	194.00
BLEND	38078.4	4452.98	0.7364	91.50	83.93	139.00
MAXIMUM SPECIFICATION	.	.	.	.	.	139.00
MINIMUM SPECIFICATION	.	.	.	91.50	82.50	.

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PRODUCT BLENDING

JET FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	SMOKE POINT
HTD KEROSENE - HIGH SP	8537.4	1104.85	0.8149	0.01	21.00
BLEND	8537.4	1104.85	0.8149	0.01	21.00
MAXIMUM SPECIFICATION	.	.	.	0.10	.
MINIMUM SPECIFICATION	.	.	.	.	21.00

.05% S NO.2 FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	CETANE INDEX
HTD.COPROCESSING DIESEL	12260.0	1669.93	0.8577	0.05	32.00
HTD.SR DIESEL	13808.5	1874.28	0.8547	0.05	51.00
HTD. COKER DIESEL	347.7	46.47	0.8417	0.05	46.00
BLEND	26416.2	3590.68	0.8559	0.05	42.12
MAXIMUM SPECIFICATION	.	.	.	0.05	.
MINIMUM SPECIFICATION	.	.	.	.	40.00

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PRODUCT BLENDING

1.0% S NO.6 FUEL

	BARRELS PER DAY	TONS PER DAY	SG 60/60	SULFUR WT PCT	VISCOSTY INDEX @ 122 F
DIESEL IPL	123.6	16.62	0.8468	0.17	14.85
VACUUM PITCH IPL	4227.4	647.11	0.9639	0.54	45.01
LIGHT CYCLE OIL	4706.5	714.09	0.9554	0.34	15.40
SLURRY OIL	3274.5	551.23	1.0600	1.04	33.20
BLEND	12332.0	1929.05	0.9850	0.61	30.41
MAXIMUM SPECIFICATION	.	.	0.9850	1.00	37.50
MINIMUM SPECIFICATION	.	.	.	.	.

COKE

	BARRELS PER DAY	TONS PER DAY	SG 60/60
HIGH SULFUR COKE	.	318.63	.
BLEND	.	318.63	.
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

SULFUR

	BARRELS PER DAY	TONS PER DAY	SG 60/60
SULFUR	.	234.07	.
BLEND	.	234.07	.
MAXIMUM SPECIFICATION	.	.	.
MINIMUM SPECIFICATION	.	.	.

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UTILITY BLENDING SUMMARY

MM BTU FUEL CONSUMED

	TONS PER DAY	UTILITY UNITS/D	CONVERSN COST \$/D
UNSATURATE REFINERY GAS	213.91	9198.22	.
SATURATE REFINERY GAS	536.70	23614.74	.
PROPANE	159.13	7001.54	.
TOTAL...	909.74	39814.50	.

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UTILITY CONVERSION SUMMARY

M LBS HIGH PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
FUEL CONSUMED	4830.82	4106.20	772.93
TOTAL...	4830.82	4106.20	772.93

M LBS MEDIUM PRESSURESTEAM

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
HIGH PRESSURESTEAM	2210.62	2210.62	.
TOTAL...	2210.62	2210.62	.

M GAL COOLING WATER

	FACTOR UNITS/D	UTILITY UNITS/D	CONVERSN COST \$/D
POWER CONSUMED	26988.13	34544.80	.
TOTAL...	26988.13	34544.80	.

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UOP PROCESS DIVISION
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DETAILED MATERIAL BALANCE

CRUDE UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
IPL CRUDE		0.8265	47500.0	50.00	6234.62	47.06	2275634.6
LLOYDMINISTER CRUDE		0.9297	47500.0	50.00	7013.09	52.94	2559778.6
TOTAL...		.	95000.0	100.00	13247.71	100.00	4835413.3

PRODUCTS

SATURATE REFINERY GAS		.	.	.	3.12	0.02	1137.8
PROPANE		0.5077	409.8	0.43	33.04	0.25	12060.9
ISO BUTANE		0.5626	263.4	0.28	23.54	0.18	8590.3
NORMAL BUTANE		0.5835	1130.3	1.19	104.74	0.79	38229.9
LT ST RUN	IPL	0.6542	2928.5	3.08	304.25	2.30	111051.0
LT ST RUN	LLOYD	0.6542	2531.4	2.66	262.99	1.99	95991.7
NAPHTHA	IPL	0.7389	10238.4	10.78	1201.41	9.07	438514.8
NAPHTHA	LLOYD	0.7459	6814.5	7.17	807.21	6.09	294630.5
KEROSENE	IPL	0.8109	5519.2	5.81	710.75	5.37	259422.3
KEROSENE	LLOYD	0.8324	2992.2	3.15	395.54	2.99	144371.5
DIESEL	IPL	0.8468	8929.2	9.40	1200.79	9.06	438287.2
DIESEL	LLOYD	0.8899	4917.8	5.18	695.00	5.25	253674.1
REDUCED CRUDE	IPL	0.9144	18414.3	19.38	2674.03	20.18	976019.7
REDUCED CRUDE	LLOYD	1.0086	30162.9	31.75	4831.32	36.47	1763431.5
TOTAL...		.	95251.9	100.27	13247.71	100.00	4835413.3

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DETAILED MATERIAL BALANCE

VACUUM UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
REDUCED CRUDE	IPL	0.9144	18414.3	37.91	2674.03	35.63	976019.7
REDUCED CRUDE	LLOYD	1.0086	30162.9	62.09	4831.32	64.37	1763431.5
TOTAL...		.	48577.3	100.00	7505.35	100.00	2739451.2

PRODUCTS

VACUUM GAS OIL	IPL	0.8996	14187.7	29.21	2026.91	27.01	739822.9
VACUUM PITCH	IPL	0.9639	4227.4	8.70	647.11	8.62	236196.8
VACUUM GAS OIL	LLOYD	0.9365	5701.1	11.74	847.90	11.30	309482.2
VACUUM PITCH	LLOYD	1.0254	24461.9	50.36	3983.42	53.07	1453949.3
TOTAL...		.	48578.2	100.00	7505.35	100.00	2739451.2

CO-PROCESSING UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
ILLINOIS NO. 6 COAL		.	.	.	2212.39	34.73	807523.4
VACUUM PITCH	LLOYD	1.0254	24461.9	100.00	3983.42	62.54	1453949.3
HYDROGEN		.	.	.	174.08	2.73	63537.6
TOTAL...		.	24461.9	100.00	6369.89	100.00	2325010.3

PRODUCTS

HYDROGEN SULFIDE		.	.	.	186.03	2.92	67899.4
SATURATE REFINERY GAS		.	.	.	717.41	11.26	261856.3
CO-PROC LT GASO.		0.6620	2875.9	11.76	302.34	4.75	110354.7
CO-PROC NAPHTHA		0.7800	6444.5	26.34	798.28	12.53	291371.4
CO-PROC DIESEL		0.8838	11857.6	48.47	1664.27	26.13	607460.0
CO-PROC GAS OIL		0.9825	8498.9	34.74	1326.08	20.82	484019.7
CO-PROC PITCH		1.1142	3181.0	13.00	562.86	8.84	205443.0
PLANT LOSSES		.	.	.	812.62	12.76	296605.7
TOTAL...		.	32857.9	134.32	6369.89	100.00	2325010.3

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DETAILED MATERIAL BALANCE

DELAYED COKER

CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
CO-PROC PITCH	1.1142	3181.0	100.00	562.86	100.00	205443.0
TOTAL...	.	3181.0	100.00	562.86	100.00	205443.0

PRODUCTS

HYDROGEN SULFIDE	.	.	.	2.70	0.48	986.1
UNSATURATE REFINERY GAS	.	.	.	56.06	9.96	20462.1
PROPYLENE	0.5220	110.7	3.48	9.17	1.63	3348.7
PROPANE	0.5077	191.3	6.01	15.42	2.74	5629.1
BUTYLENE	0.6100	126.7	3.98	12.27	2.18	4478.7
ISO BUTANE	0.5626	28.3	0.89	2.53	0.45	924.5
NORMAL BUTANE	0.5835	137.3	4.32	12.72	2.26	4643.0
COKER NAPHTHA	0.7424	694.1	21.82	81.84	14.54	29871.4
COKER DISTILLATE	0.8961	334.6	10.52	47.62	8.46	17380.5
COKER GAS OIL	0.9639	25.4	0.80	3.88	0.69	1417.6
HIGH SULFUR COKE	.	.	.	318.63	56.61	116301.3
TOTAL...	.	1648.4	51.82	562.86	100.00	205443.0

KEROSENE HYDROTREATER CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
KEROSENE IPL	0.8109	5519.2	64.85	710.75	64.17	259422.3
KEROSENE LLOYD	0.8324	2992.2	35.15	395.54	35.71	144371.5
HYDROGEN	.	.	.	1.33	0.12	484.6
TOTAL...	.	8511.3	100.00	1107.61	100.00	404278.4

PRODUCTS

HYDROGEN SULFIDE	.	.	.	2.10	0.19	767.2
SATURATE REFINERY GAS	.	.	.	0.66	0.06	242.3
HTD KEROSENE - HIGH SP	0.8149	8537.4	100.31	1104.85	99.75	403268.9
TOTAL...	.	8537.4	100.31	1107.61	100.00	404278.4

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DETAILED MATERIAL BALANCE

DIESEL HYDROTREATER	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
DIESEL	IPL	0.8468	8805.6	33.98	1184.17	32.86	432220.4
DIESEL	LLOYD	0.8899	4917.8	18.98	695.00	19.28	253674.1
CO-PROC DIESEL		0.8838	11857.6	45.75	1664.27	46.18	607460.0
COKER DISTILLATE		0.8961	334.6	1.29	47.62	1.32	17380.5
HYDROGEN		.	.	.	13.06	0.36	4765.1
TOTAL...		.	25915.6	100.00	3604.11	100.00	1315500.1

PRODUCTS

HYDROGEN SULFIDE	.	.	.	9.43	0.26	3441.0
SATURATE REFINERY GAS	.	.	.	4.00	0.11	1461.0
HTD. COKER DIESEL	0.8417	347.7	1.34	46.47	1.29	16961.6
HTD.SR DIESEL	0.8547	13808.5	53.28	1874.28	52.00	684111.2
HTD.COPROCESSING DIESEL	0.8577	12260.0	47.31	1669.93	46.33	609525.4
TOTAL...	.	26416.2	101.93	3604.11	100.00	1315500.1

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DETAILED MATERIAL BALANCE

GAS OIL HYDROTREATER	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
VACUUM GAS OIL	LLOYD	0.9365	5701.1	40.08	847.90	38.72	309482.2
COKER GAS OIL		0.9639	25.4	0.18	3.88	0.18	1417.6
CO-PROC GAS OIL		0.9825	8498.9	59.74	1326.08	60.56	484019.7
HYDROGEN		.	.	.	11.95	0.55	4363.4
TOTAL...		.	14225.4	100.00	2189.82	100.00	799282.9

PRODUCTS

HYDROGEN SULFIDE	.	.	.	28.49	1.30	10399.6
SATURATE REFINERY GAS	.	.	.	12.27	0.56	4480.0
HTD MIXED GAS OILS	0.9202	5730.2	40.28	837.38	38.24	305644.6
HTD COKER DIESEL	0.8994	26.3	0.18	3.75	0.17	1370.1
HTD CO-PROCESSING GAS OI	0.9498	8671.1	60.95	1307.91	59.73	477388.6
TOTAL...	.	14427.6	101.42	2189.82	100.00	799282.9

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DETAILED MATERIAL BALANCE

FLUID CATALYTIC CRACKER CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HTD MIXED GAS OILS	0.9202	5730.2	20.02	837.38	20.05	305644.6
VACUUM GAS OIL IPL	0.8996	14187.7	49.58	2026.91	48.54	739822.9
HTD CO-PROCESSING GAS OI	0.9498	8671.1	30.30	1307.91	31.32	477388.6
HTD COKER DIESEL	0.8994	26.3	0.09	3.75	0.09	1370.1
TOTAL...	.	28615.3	100.00	4175.96	100.00	1524226.3

PRODUCTS

HYDROGEN SULFIDE	.	.	.	3.76	0.09	1371.8
UNSATURATE REFINERY GAS	.	.	.	157.85	3.78	57615.8
PROPYLENE	0.5220	2191.3	7.66	181.65	4.35	66303.8
PROPANE	0.5077	854.6	2.99	68.90	1.65	25149.7
BUTYLENE	0.6100	2336.4	8.16	226.34	5.42	82613.1
ISO BUTANE	0.5626	1533.1	5.36	136.97	3.28	49994.6
NORMAL BUTANE	0.5835	459.7	1.61	42.59	1.02	15547.1
FCC GASOLINE	0.7599	15492.2	54.14	1869.58	44.77	682396.1
LIGHT CYCLE OIL	0.9554	4706.5	16.45	714.09	17.10	260642.7
SLURRY OIL	1.0600	3274.5	11.44	551.23	13.20	201197.9
FCC COKE	.	.	.	223.00	5.34	81393.7
TOTAL...	.	30848.3	107.80	4175.96	100.00	1524226.3

HYDROGEN PLANT

CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
SATURATE REFINERY GAS	.	.	.	248.81	100.00	90815.7
TOTAL...	.	.	.	248.81	100.00	90815.7

PRODUCTS

HYDROGEN	.	.	.	120.33	48.36	43919.0
PLANT LOSSES	.	.	.	128.48	51.64	46896.7
TOTAL...	.	.	.	248.81	100.00	90815.7

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DETAILED MATERIAL BALANCE

ALKYLATION UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
PROPYLENE		0.5220	2302.0	21.49	190.83	19.86	69652.6
BUTYLENE		0.6100	2463.1	22.99	238.61	24.83	87091.7
ISO BUTANE		0.5626	5949.1	55.53	531.53	55.31	194007.7
TOTAL...		.	10714.2	100.00	960.96	100.00	350752.0

PRODUCTS

PROPANE	0.5077	229.5	2.14	18.51	1.93	6755.4
NORMAL BUTANE	0.5835	48.9	0.46	4.53	0.47	1654.7
ALKYLATE FROM PROPYLENE	0.7036	4020.3	37.52	449.22	46.75	163965.5
ALKYLATE FROM BUTYLENE	0.7036	4362.5	40.72	487.45	50.73	177919.8
TAR FROM HF ALKY UNIT	.	.	.	1.25	0.13	456.7
TOTAL...	.	8661.3	80.84	960.96	100.00	350752.0

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DETAILED MATERIAL BALANCE

NAPHTHA HYDROTREATER CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
NAPHTHA	IPL	0.7389	10238.4	37.31	1201.41	37.18	438514.8
NAPHTHA	LLOYD	0.7459	6814.5	24.84	807.21	24.98	294630.5
LT ST RUN	IPL	0.6542	2928.5	10.67	304.25	9.41	111051.0
LT ST RUN	LLOYD	0.6542	318.9	1.16	33.13	1.03	12094.0
COKER NAPHTHA		0.7424	694.1	2.53	81.84	2.53	29871.4
CO-PROC NAPHTHA		0.7800	6444.5	23.49	798.28	24.70	291371.4
HYDROGEN		.	.	.	5.53	0.17	2016.8
TOTAL...		.	27438.9	100.00	3231.64	100.00	1179549.9

PRODUCTS

HYDROGEN SULFIDE	.	.	.	3.92	0.12	1431.8
SATURATE REFINERY GAS	.	.	.	0.91	0.03	331.9
PROPANE	0.5077	7.3	0.03	0.59	0.02	214.1
NORMAL BUTANE	0.5835	7.3	0.03	0.67	0.02	246.3
LNA	0.6540	3397.1	12.38	352.83	10.92	128781.7
HHN	0.7247	24961.0	90.97	2872.72	88.89	1048544.2
TOTAL...	.	28372.7	103.40	3231.64	100.00	1179549.9

C5/C6 PENEX UNIT CHARGE

		SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
LNA		0.6540	3397.1	100.00	352.83	99.13	128781.7
HYDROGEN		.	.	.	3.10	0.87	1133.3
TOTAL...		.	3397.1	100.00	355.93	100.00	129915.0

PRODUCTS

SATURATE REFINERY GAS	.	.	.	17.29	4.86	6310.3
C5/C6 PENEXATE	0.6440	3311.2	97.47	338.64	95.14	123604.7
TOTAL...	.	3311.2	97.47	355.93	100.00	129915.0

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DETAILED MATERIAL BALANCE

PLATFORMING UNIT	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HHN		0.7247	24961.0	100.00	2872.72	100.00	1048544.2
TOTAL...		.	24961.0	100.00	2872.72	100.00	1048544.2

PRODUCTS

HYDROGEN	.	.	.	88.72	3.09	32381.8
SATURATE REFINERY GAS	.	.	.	29.84	1.04	10891.3
PROPANE	0.5077	281.1	1.13	22.66	0.79	8271.7
ISO BUTANE	0.5626	125.2	0.50	11.19	0.39	4083.4
NORMAL BUTANE	0.5835	179.5	0.72	16.64	0.58	6072.7
98 RON PLATFORMATE	0.8058	21007.9	84.16	2688.33	93.58	981240.3
94 RON PLATFORMATE	0.7909	122.2	0.49	15.35	0.53	5602.9
TOTAL...	.	21715.9	87.00	2872.72	100.00	1048544.2

AMINE TR- SULFUR PLANT CHARGE

	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
HYDROGEN SULFIDE	.	.	.	236.43	100.00	86297.0
TOTAL...	.	.	.	236.43	100.00	86297.0

PRODUCTS

SULFUR	.	.	.	234.07	99.00	85434.0
SOX EMISSIONS	.	.	.	2.36	1.00	863.0
TOTAL...	.	.	.	236.43	100.00	86297.0

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DETAILED MATERIAL BALANCE

OFFSITE ALLOWANCE	CHARGE	SPECIFIC GRAVITY	VOLUME BBL/DAY	VOLUME PERCENT	WEIGHT TONS/DAY	WEIGHT PERCENT	METRIC TON/YEAR
	IPL CRUDE	0.8265	47500.0	50.00	6234.62	40.33	2275634.6
	LLOYDMINISTER CRUDE	0.9297	47500.0	50.00	7013.09	45.36	2559779.2
	ILLINOIS NO. 6 COAL	.	.	.	2212.39	14.31	807523.4
	TOTAL...	.	95000.0	100.00	15460.10	100.00	5642937.3

PRODUCTS

IPL CRUDE	0.8265	47500.0	50.00	6234.62	40.33	2275634.6
LLOYDMINISTER CRUDE	0.9297	47500.0	50.00	7013.09	45.36	2559779.2
ILLINOIS NO. 6 COAL	.	.	.	2212.39	14.31	807523.4
TOTAL...	.	95000.0	100.00	15460.10	100.00	5642937.3

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UTILITY CONSUMPTION SUMMARY

PROCESS UNITS	COMBINED MODE	LABOR MAINTNCE INSURANCE DOLLARS	POWER CONSUMED KWH	FUEL CONSUMED MM BTU	HIGH PRESSURE STEAM M LBS	MEDIUM PRESSURE STEAM M LBS	LOW PRESSURE STEAM M LBS	COOLING WATER M GAL	CATALYST CONSUMED DOLLARS
CRUDE UNIT		7471.71	58952.30	6557.62	.	543.16	.	168.25	.
VACUUM UNIT		4225.51	13029.28	4000.35	.	825.59	112.58	5208.71	.
CO-PROCESSING UNIT		28760.31	585563.13	4725.14	624.60	.	-4138.78	6472.66	76521.55
DELAYED COKER		1407.14	8062.94	675.43	.	-25.33	9.68	262.01	.
KEROSENE HYDROTREATER		2780.09	7821.43	297.59	66.38	.	.	27.66	.
DIESEL HYDROTREATER		6015.02	31488.66	965.99	215.46	.	.	89.78	.
GAS OIL HYDROTREATER		2188.75	24077.17	585.84	130.67	.	.	54.45	.
FLUID CATALYTIC CRACKER		12147.87	34447.51	.	2238.32	-1181.80	-2238.32	2873.06	3804.30
HYDROGEN PLANT		13753.27	28517.28	6756.31	-593.21	.	.	105.47	.
ALKYLATION UNIT		5754.91	16559.43	1875.22	.	.	749.34	15520.66	532.03
NAPHTHA HYDROTREATER		4064.91	17812.13	2165.69	.	.	.	1689.20	.
C5/C6 PENEX UNIT		2226.34	5107.52	.	.	.	219.46	238.51	354.59
PLATFORMING UNIT		4021.81	35860.49	5683.63	-786.64	.	.	1417.58	1004.65
AMINE TR- SULFUR PLANT		10728.49	24434.34	694.87	.	-347.32	1418.58	416.83	236.43
OFFSITE ALLOWANCE		15181.82	25339.11	.	.	2396.32	.	.	.
TOTAL...		120727.95	917072.71	34983.68	1895.58	2210.62	-3867.45	34544.80	82453.55

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UTILITY CONSUMPTION SUMMARY

PROCESS UNITS	COMBINED MODE	
	I	M

CRUDE UNIT	.	
VACUUM UNIT	.	
CO-PROCESSING UNIT	.	
DELAYED COKER	.	
KEROSENE HYDROTREATER	.	
DIESEL HYDROTREATER	.	
GAS OIL HYDROTREATER	.	
FLUID CATALYTIC CRACKER	.	
HYDROGEN PLANT	.	
ALKYLATION UNIT	.	
NAPHTHA HYDROTREATER	.	
C5/C6 PENEX UNIT	.	
PLATFORMING UNIT	.	
AMINE TR- SULFUR PLANT	.	
OFFSITE ALLOWANCE	.	
TOTAL...	.	

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UTILITY COST

		COST
LABOR	MAINTNCEINSURNCEDOLLARS	1.00
CATALYST	CONSUMED DOLLARS	1.00
POWER	CONSUMED KWH	0.06
FUEL	CONSUMED MM BTU	3.75
HIGH	PRESSURESTEAM M LBS	7.20
MEDIUM	PRESSURESTEAM M LBS	5.25
LOW	PRESSURESTEAM M LBS	4.80
COOLING	WATER M GAL	0.10

UTILITY CONSUMPTION COST SUMMARY

	LABOR MAINTNCE INSURNC DOLLARS	CATALYST CONSUMED DOLLARS	POWER CONSUMED DOLLARS	FUEL CONSUMED DOLLARS	HIGH PRESSURE STEAM DOLLARS	MEDIUM PRESSURE STEAM DOLLARS	LOW PRESSURE STEAM DOLLARS	COOLING WATER DOLLARS
CRUDE UNIT	7471.71	.	3537.14	24591.06	.	2851.57	.	16.82
VACUUM UNIT	4225.51	.	781.76	15001.31	.	4334.34	540.38	520.87
CO-PROCESSING UNIT	28760.31	76521.55	35133.79	17719.26	4497.12	.	-19866.13	647.27
DELAYED COKER	1407.14	.	483.78	2532.86	.	-132.98	46.47	26.20
KEROSENE HYDROTREATER	2780.09	.	469.29	1115.96	477.91	.	.	2.77
DIESEL HYDROTREATER	6015.02	.	1889.32	3622.48	1551.34	.	.	8.98
GAS OIL HYDROTREATER	2188.75	.	1444.63	2196.92	940.84	.	.	5.44
FLUID CATALYTIC CRACKER	12147.87	3804.30	2066.85	.	16115.87	-6204.44	-10743.92	287.31
HYDROGEN PLANT	13753.27	.	1711.04	25336.16	-4271.09	.	.	10.55
ALKYLATION UNIT	5754.91	532.03	993.57	7032.07	.	.	3596.82	1552.07
NAPHTHA HYDROTREATER	4064.91	.	1068.73	8121.35	.	.	.	168.92
C5/C6 PENEX UNIT	2226.34	354.59	306.45	.	.	.	1053.40	23.85
PLATFORMING UNIT	4021.81	1004.65	2151.63	21313.63	-5663.82	.	.	141.76
AMINE TR- SULFUR PLANT	10728.49	236.43	1466.06	2605.75	.	-1823.41	6809.19	41.68
OFFSITE ALLOWANCE	15181.82	.	1520.35	.	.	12580.66	.	.
TOTAL COST (BY UTILITY)	120727.95	82453.55	55024.36	131188.80	13648.17	11605.75	-18563.78	3454.48

UTILITY CONSUMPTION COST SUMMARY

	TOTAL COST (BY UNIT)
CRUDE UNIT	38468.30
VACUUM UNIT	25404.17
CO-PROCESSING UNIT	143413.17
DELAYED COKER	4363.47
KEROSENE HYDROTREATER	4846.02
DIESEL HYDROTREATER	13087.13
GAS OIL HYDROTREATER	6776.58
FLUID CATALYTIC CRACKER	17473.86
HYDROGEN PLANT	36539.91
ALKYLATION UNIT	19461.46
NAPHTHA HYDROTREATER	13423.91
C5/C6 PENEX UNIT	3964.63
PLATFORMING UNIT	22969.65
AMINE TR- SULFUR PLANT	20064.19
OFFSITE ALLOWANCE	29282.82
TOTAL COST (BY UTILITY)	399539.28

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PLANT CAPACITY SUMMARY II
--- BY CAPACITY TYPES ---

PROCESS UNIT	UNITS OF CAPACITY	MAXIMUM	MINIMUM	CALENDAR DAY BASIS	STREAM DAY BASIS	INVSTMNT	PER UNIT OPERATIN COST
CRUDE UNIT	BBLS CAPACITY	100000.0	.	95000.0	100000.0	40680000.	0.40
VACUUM UNIT	BBLS CAPACITY	.	.	48570.1	51126.4	14322398.	0.52
CO-PROCESSING UNIT	BBLS CAPACITY	.	25683.0	24461.9	27059.6	183564703.	5.86
DELAYED COKER	BBLS CAPACITY	.	.	3181.0	3348.4	23202738.	1.37
KEROSENE HYDROTREATER	BBLS CAPACITY	.	.	8511.3	8959.3	6214466.	0.57
DIESEL HYDROTREATER	BBLS CAPACITY	.	.	25921.5	27285.8	18694196.	0.50
GAS OIL HYDROTREATER	BBLS CAPACITY	.	.	14225.6	14974.3	15708011.	0.48
FLUID CATALYTIC CRACKER	BBLS CAPACITY	.	10000.0	28615.3	30121.3	67986422.	0.61
ALKYLATION UNIT	BBLS CAPACITY	.	.	8382.8	8824.0	35906187.	2.32
NAPHTHA HYDROTREATER	BBLS CAPACITY	.	.	27488.2	28934.9	7789679.	0.49
C5/C6 PENEX UNIT	BBLS CAPACITY	.	.	3397.1	3575.9	7663919.	1.17
PLATFORMING UNIT	BBLS CAPACITY	.	.	24382.3	25665.6	36808091.	0.94
OFFSITE ALLOWANCE	BBLS CAPACITY	.	.	103223.5	103223.5	391326199.	0.28
HYDROGEN PLANT	TONS CAPACITY	.	.	120.3	133.7	31166819.	303.67
AMINE TR- SULFUR PLANT	TONS CAPACITY	.	.	236.4	248.9	29524162.	84.86

TOTAL INVESTMENT

910557988.

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RAW MATERIAL PURCHASE SUMMARY

TIME 1 BASE MODEL

PURCHASED MATERIALS

		MAXIMUM	MINIMUM	QUANTITY BOUGHT	PRICE	TOTAL \$ PER DAY
IPL CRUDE	BBLS	47500.00	47500.00	47500.00	38.36	1822100.00
LLOYDMINISTER CRUDE	BBLS	47500.00	47500.00	47500.00	33.00	1567500.00
ILL. NO. 6 COAL	TONS	.	.	2212.39	35.00	77433.75
ISOBUTANE	BBLS	.	.	3999.09	28.70	114773.89
NORMAL BUTANE	BBLS	.	.	1750.06	24.70	43226.41

TOTAL RAW MATERIAL PURCHASE3625034.05

RAW MATERIAL PURCHASE RESTRICTIONS:

Appendix E

Definitions and Calculation Procedures

FEED CALCULATIONS

Coal

The coal is analyzed by U866 for LECO C, H, N; U864 for LECO S; U384 for Kjeldahl N; and D3172 for proximate analysis. In addition, a sample is ashed by the technician, and the ash is submitted for metals analysis by U714, and wt-% Si and Al by AAS.

The "proximate analysis" provides an as-received (AR) breakdown of the coal into moisture, volatiles, ash, and fixed carbon fractions. This method uses a controlled heat-up in an oven to drive off the moisture at 104-110°C. The sample is then gradually heated further to reach 500°C in one hour and 750°C after two hours. The "ash" is the material remaining when the sample is heated at 750°C to constant weight. A separate sample is heated rapidly for seven minutes in an oven preheated to 950°C to drive off the volatiles (which are corrected for moisture). The fixed carbon is determined by difference between the total sample and the percentages of moisture, volatiles, and ash.

The "ultimate" analysis provides an elemental breakdown of the coal. The hydrogen (measured by U866H) must be corrected for the H that is present as water. Oxygen is determined by difference.

$$\begin{aligned} W\%COALA &= (\text{wt-\% coal ash, by D3174}) \\ W\%COALM &= (\text{wt-\% coal moisture, by D3173}) \\ W\%COALC &= (\text{wt-\% coal C, by U866C}) \\ W\%COALH &= (\text{wt-\% coal H, by U866H}) - 2/18 \cdot W\%COALM \\ W\%COALS &= (\text{wt-\% coal S, by U864S}) \\ W\%COALN &= (\text{wt-\% coal N, by U866N}) \\ W\%COALO &= 100\% - W\%COALA - W\%COALM - W\%COALC - \\ &\quad W\%COALH - W\%COALS - W\%COALN \end{aligned}$$

The ultimate analysis may also be represented on a "moisture free" (MF) or "moisture and ash free" (MAF) basis by normalizing out the water, or water and ash, respectively. We use the MF basis for calculating the total co-processing feed composition because the feed is dried in the slurry tank before being transferred to the feed day tank. The MAF basis is used for reporting results (to eliminate differences in the ash content of similar coals, and to keep the catalyst separate from the feed basis).

$$W\%MFCA = (\text{wt-\% MF coal Ash}) = W\%COALA*100\%/(100\%-W\%COALM)$$

$$W\%MFCC = (\text{wt-\% MF coal C}) = W\%COALC*100\%/(100\%-W\%COALM)$$

$$W\%MFCH = (\text{wt-\% MF coal H}) = W\%COALH*100\%/(100\%-W\%COALM)$$

$$W\%MFCS = (\text{wt-\% MF coal S}) = W\%COALS*100\%/(100\%-W\%COALM)$$

$$W\%MFCN = (\text{wt-\% MF coal N}) = W\%COALN*100\%/(100\%-W\%COALM)$$

$$W\%MAFCC = (\text{wt-\% MAF coal C}) = W\%MFCC*100\%/(100\%-W\%MFCA)$$

$$W\%MAFCH = (\text{wt-\% MAF coal H}) = W\%MFCH*100\%/(100\%-W\%MFCA)$$

$$W\%MAFCS = (\text{wt-\% MAF coal S}) = W\%MFCS*100\%/(100\%-W\%MFCA)$$

$$W\%MAFCN = (\text{wt-\% MAF coal N}) = W\%MFCN*100\%/(100\%-W\%MFCA)$$

Resid

The resid feedstocks are analyzed by U866 for wt-% carbon, hydrogen, and nitrogen, U864 for wt-% S, U614 for wt-% heptane insolubles, D4530 for MCRT, and AAS for metals.

$$W\%RESC = (\text{wt-\% resid C, by U866C})$$

$$W\%RESH = (\text{wt-\% resid H, by U866H})$$

$$W\%RESN = (\text{wt-\% resid N, by U384})$$

$$W\%RESS = (\text{wt-\% resid S, by U864})$$

$$W\%RESHI = (\text{wt-\% resid heptane insolubles, by U614})$$

W%RESMC = (wt-% resid micro-carbon residue, by D4350)

The resid is also distilled by D1160 into overhead and bottoms fractions, with the temperature raised from the IBP to at least 510°C. The D1160 overhead fraction is then further analyzed by GLC (D2887) to give a boiling range breakdown at the desired yield points (typically 177, 343, 371 and 510°C).

GRO = Grams (or wt%) of Resid taken overhead in D1160

GRB = Grams (or wt%) of Resid remaining undistilled in D1160.

W%RES177+= (100% - wt-% over at 177°C in D1160 OH GLC)*
GRO/(GRO+GRB)

W%RES343+= (100% - wt-% over at 343°C in D1160 OH GLC)*
GRO/(GRO+GRB)

W%RES371+= (100% - wt-% over at 371°C in D1160 OH GLC)*
GRO/(GRO+GRB)

W%RES510+= (100% - wt-% over at 510°C in D1160 OH GLC)*
GRO/(GRO+GRB)

Feed Blend Procedures

The coal is weighed out, sealed under nitrogen pressure in a plastic bag, and transported by bucket to the plant site, where it is temporarily stored out in the shed. A large batch of feed is then prepared in the plant slurry tank. First, the resid is weighed and added to the tank. Then the heater is turned on and the mixer is started. Coal is added, little by little. A slight vacuum is pulled on the slurry tank to "suck" in the coal. It takes about three hours to add the full load of coal for a 2:1 resid:coal blend. The catalyst is added last, and usually goes in quite easily relative to the coal.

After the slurry is warmed up a little, the circulation pump is started, and the temperature is raised to 150°C, driving off any moisture in the feed. The water coming off is trapped and measured. It takes about two days to get off all of the moisture in a typical run. Greater than 99% moisture removal is required to avoid charge pump cavitation problems.

One slurry tank of feed lasts about three weeks at the base space velocity. A new batch of feed is usually prepared when the tank level gets down to about one third. The slurry tank is analyzed for catalytic metals, ash (by MAM), API, and road oil viscosity (usually at 350 and 450 °F). One sample of each new feed is also run at reference conditions in the autoclave.

The feed analyses in the data base are calculated from the weights and analyses of each individual component added to the slurry tank (resid, coal, and catalyst). The total is normalized to eliminate the moisture removed in drying. The ash and metals analyses of the total feed are used as a check of the blend calculation. The water removed in drying is also checked vs. the water calculated from the coal and catalyst analyses.

Each feed component is weighed as it is added to the feed slurry tank:

WCOAL = weight of AR coal, kg.

WRESID = weight of resid, kg.

WCAT = weight of catalyst, kg.

Then:

$$WMFC = (100\% - W\%COALM) / 100\% * WCOAL$$

$$WMAFC = (100\% - W\%MFCA) / 100\% * WMFC$$

$$W\%MFC = WMFC / (WMFC + WRESID + WCAT) * 100\%$$

$$W\%MAFC = WMAFC/(WMFC+WRESID+WCAT)*100\%$$

$$W\%RES = WRESID/(WMFC+WRESID+WCAT)*100\%$$

$$W\%CAT = WCAT/(WMFC+WRESID+WCAT)*100\%$$

Total Feed

The feed charge tank is mounted on a scale. The total weight of feed (= GFEEED) used during the test period is measured by difference in scale readings, and is corrected for the weight of any material collected from feed pump leaks that never makes it to the plant. A scaling factor is also used to correct for any drift in the balance readings.

Then:

$$GMAFC = W\%MAFC/100\%*GFEEED$$

$$GRESID = W\%RESID/100\%*GFEEED$$

$$GMAFF = GMAFC + GRESID = (\text{grams MAF feed})$$

$$GFEEEDC = (W\%MAFCC*GMAFC+W\%RESC*GRESID)/100\%$$

$$GFEEEDH = (W\%MAFCH*GMAFC+W\%RESH*GRESID)/100\%$$

$$GFEEEDS = (W\%MAFCS*GMAFC+W\%RESS*GRESID)/100\%$$

$$GFEEEDN = (W\%MAFCN*GMAFC+W\%RESN*GRESID)/100\%$$

$$GFEEEDHI = GMAFC+W\%RESHI*GRESID/100\%$$

$$GFEEEDMC = GMAFC+W\%RESMC*GRESID/100\%$$

$$GFEEED177+ = GMAFC+(W\%RES177+)*GRESID/100\%$$

$$GFEEED343+ = GMAFC+(W\%RES343+)*GRESID/100\%$$

$$GFEEED371+ = GMAFC+(W\%RES371+)*GRESID/100\%$$

$$GFEEED510+ = GMAFC+(W\%RES510+)*GRESID/100\%$$

The Feed Distribution on an MAF basis is:

$$W\%FEEDNAPH = [(GRESID)-(GFEED177+)]/GMAFF*100\%$$

$$W\%FEEDDIST = [(GFEED177+)-(GFEED343+)]/GMAFF*100\%$$

$$W\%FEEDVGO = [(GFEED343+)-(GFEED510+)]/GMAFF*100\%$$

$$W\%FEEDRESID = [(GFEED510+)]/GMAFF*100\%$$

$$W\%FEEDMAFC = GMAFC/GMAFF*100\%$$

$$TOTALFEED = 100\%$$

GAS CALCULATIONS

There are two net gas streams, the net recycle gas (RG), and the debutanizer offgas (DG), which are scrubbed and saturated in caustic bubblers, sampled for GC analysis, and metered by wet test meter (WTM).

The caustic bubblers are weighed at the beginning and end of the test period and analyzed for sulfur (wt-% as H₂S and ppm as RSH) by a modified U163 method. The samples are potentiometrically titrated with a silver nitrate solution. The plotted curve has two inflection points, one for H₂S, and a second for RSH. The results are reported as ppm S (not H₂S or RSH). The debutanizer bubbler sometimes acts as a knock-out pot for light hydrocarbons in the off-gas stream. The amount of hydrocarbon can be calculated from the difference in bubbler weights during the test period after adjusting for the sulfur scrubbed out in the bubbler.

Let:

GRGSi = initial weight of recycle gas bubbler

GRGSf = final weight of recycle gas bubbler

GDGSi = initial weight of debutanizer gas bubbler

GDGSf = final weight of debutanizer gas bubbler

Then:

GRGSS = $GRGSf * [(wt\% \text{ S as } H_2S \text{ in RG})/100\% + (ppm \text{ S as RSH in RG})/10^6]$
= grams S in recycle net gas

GDGSS = $GDGSf * [(wt\% \text{ S as } H_2S \text{ in DG})/100\% + (ppm \text{ S as RSH in DG})/10^6]$
= grams S in debut. off-gas

GDGHC = $(GDGSf - GDGSI) - GDGSS$ = grams hydrocarbon in debutanizer offgas scrubber.

The net gas rates are computed in grams/period from the wet test meter readings and the volume of samples sent for analysis (since the samples are taken off before the wet test meter). The gas measured on the WTM is water saturated, because the gas goes through a scrubber prior to metering. The vapor pressure must therefore be corrected to remove the water of saturation. The partial pressure of water for a given ambient temperature may be calculated by the Antoine equation:

$$H_2O \text{ Vapor Pressure Correction} = e^{(16.85 - 5700 / (T^{\circ}F + 340))}$$

Then:

MRG or MDG = $(\text{cubic feet of gas/period on WTM}) * (\text{WTM factor}) * (453.6 \text{ gmol/lbmol}) / (379.431 \text{ SCF/lbmol})$ = moles/period of recycle gas or debutanizer gas

where,

WTM Factor = $[(\text{barometric pressure, mmHg} - \text{vapor P of } H_2O, \text{ mmHg})/760]$
 $* [273^{\circ}C / (273 + \text{gas temperature, } ^{\circ}C)]$

Note: The cf from the WTM is hand-corrected by the plant operator for the volume of the

gas sampling apparatus before the data is input to the spreadsheet. No additional correction is therefore required in the spreadsheet.

The gas streams are analyzed by GC to give the mole-% of the following components: O₂, N₂, H₂, CO, CO₂, CH₄, C₂, C₂=, C₃, C₃=, iC₄, nC₄, C₄=, iC₅, nC₅ and C₆+.

The gas samples should contain no oxygen. It is assumed that any O₂ present came from air that leaked into the sample during handling. Oxygen and a stoichiometric amount of nitrogen (based on the average composition of air) are normalized out of the analyses. In the case of the debutanizer gas, there is a nitrogen purge on the debutanizer column, so only that part of the nitrogen attributed to air is normalized out:

$$(\text{normalized moles component } i)_{RG} = [(\text{mole\% } i)_{RG} / (100\% - \text{mole\% O}_2 - \text{mole\% N}_2)] * \text{MRG}$$

$$(\text{normalized moles component } i)_{DG} = [(\text{mole\% } i)_{DG} / (100\% - \text{mole\% O}_2 - 3.563 * \text{mole\% O}_2)] * \text{MDG}$$

where

$$3.563 = \frac{(78.084 \% \text{ Nitrogen})}{(20.946\% \text{ O}_2 + 0.934\% \text{ Ar} + 0.033\% \text{ CO}_2)}, \text{ in air}$$

The GC analyses do not show NH₃, although it is present in the total gas that is measured on the wet test meter. (Ammonia is not appreciably absorbed in the caustic scrubber, but may form (NH₄)₂S, which has a slight solubility in the water of the three-phase separator. We have assumed that the NH₃ stays in the off-gases and comes out in the GC column. The total moles of gas must, therefore, be normalized down by the amount of NH₃, which is calculated by mass balance from disappearance of N in the other streams:

$$\text{GGASN} = (\text{grams N in gas}) = (\text{grams N in feed}) - (\text{grams N in overhead liquid}) - (\text{grams N in TSB}) - (\text{grams N in TIB})$$

$$(\text{Gas adj. factor}) = 1 - [(\text{g N in gas}) / (14 \text{ g/mole N})] / (\text{total gas moles})$$

$$(\text{grams NH}_3 \text{ in gas}) = (17.031 \text{ g/mole NH}_3) / (14.007 \text{ g/mole N}) * (\text{grams N in gas})$$

The other hetero-gas yields are based on measured values. The H₂S yield is based on the S scrubbed out in the off-gas bubblers.

$$\begin{aligned} (\text{grams H}_2\text{S in gas}) = & (34.016 \text{ g/mole H}_2\text{S}) / (32 \text{ g/mole S}) * \{[(\text{wt-\% S as H}_2\text{S in RG}) / 100\% + (\text{ppm S as RSH in RG}) / 10^6] * \text{GRGSf} \\ & + [(\text{wt-\% S as H}_2\text{S in DG}) / 100\% + (\text{ppm S as RSH in DG}) / 10^6] * \text{GDGSf} \} \end{aligned}$$

Sulfur balances that include only the S recovered in the gas bubblers have historically been low (~60-70%). Starting with Run 21, we began analyzing the water drained from the three-phase separator on a spot basis. This stream was found to contain most of the "missing" sulfur. Although this sulfur does not show up as gaseous H₂S, it should still be included in the total H₂S yield when comparing our results with those from other processes where the H₂S yields are calculated by disappearance of sulfur from the feed and liquid products.

$$\text{GH2OS} = (\text{grams S with net H}_2\text{O}) = [(\text{wt-\% S as H}_2\text{S in H}_2\text{O}) / 100\% + (\text{ppm S as RSH in H}_2\text{O}) / 10^6] * (\text{grams H}_2\text{O drained})$$

$$\text{GGASS} = (\text{grams S in gases}) = (32 \text{ g/mole S}) / (34.016 \text{ g/mole H}_2\text{S}) * (\text{grams H}_2\text{S in gas})$$

The net water product is calculated as the difference between the water collected from the three-phase separator and the water injected into the HPS overhead line. The water input is measured by the change in level of a graduate. The water from the separator contains traces of S, so it is collected in a scrubber containing caustic. The water is measured as the change in weight of the scrubber from the beginning to the end of the test period minus the S in the scrubber.

$$(\text{Grams H}_2\text{O}) = [(\text{grams H}_2\text{O drained}) - (\text{grams H}_2\text{O input})] - (\text{grams S with net H}_2\text{O})$$

The grams of each component in the gas streams are calculated from the adjusted moles and molecular weight:

$$(\text{grams component i}) = (\text{normalized moles of i}) * (\text{gas adj. factor}) * (\text{MW of i, g/mole})$$

where the MW's are given in the following table:

<u>Component</u>	<u>MW</u>	<u>wt-% C</u>	<u>wt-% H</u>	<u>wt-% O</u>
O ₂	32.00	0.00	0.00	100.0
N ₂	28.00	0.00	0.00	0.00
H ₂	2.02	0.00	100.0	0.00
CO	28.01	42.84	0.00	57.16
CO ₂	44.01	27.27	0.00	72.73
CH ₄	16.04	74.81	25.19	0.00
C ₂	30.07	79.81	20.19	0.00
C ₂ =	28.05	85.56	14.44	0.00
C ₃	44.09	81.65	18.35	0.00
C ₃ =	42.08	85.56	14.44	0.00
iC ₄	58.12	82.59	17.41	0.00
nC ₄	58.12	82.59	17.41	0.00
C ₄ =	56.10	85.56	14.44	0.00
iC ₅	72.15	83.16	16.84	0.00
nC ₅	72.15	83.16	16.84	0.00
C ₆ +	86.17	83.56	16.44	0.00

For example,

$$\begin{aligned}
 (\text{grams Gas } C_1) = & [(\text{norm. mol\% } C_1 \text{ in RG}) * (\text{moles RG/period}) + (\text{norm.} \\
 & \text{mol\% } C_1 \text{ in DG}) * (\text{moles DG/ period})] * (\text{Gas adj. factor}) \\
 & * (16.04 \text{ g/mole})/100\%
 \end{aligned}$$

The elemental contents of the gas streams are also computed component by component from the %CH information in the table above:

$$GGASC = \text{Sum}_i [(\text{grams component}_i) * (\text{wt-\% C})_i / 100\%]$$

$$GGASH = \text{Sum}_i [(\text{grams component}_i) * (\text{wt-\% H})_i / 100\%]$$

LIQUID PRODUCT WORKUP

Light hydrocarbon product is collected from the bottom of the debutanizer. This stream is typically a small product stream, usually less than 2 wt-% of the MAF feed. To minimize the laboratory analysis, the debutanizer bottoms is combined with the vacuum column overhead before analysis. A number of other very small streams are also collected and weighed. These include any light material caught in the traps or recovered from leaks (e.g., the fractionator hydrocarbon trap, the debutanizer hydrocarbon trap, the three-phase separator knock-out pot, the recycle displacement meter drain, hydrocarbon entrained and later separated out of the water removed from the three-phase separator, hydrocarbon knocked out in the debutanizer scrubber (See above), and hydrocarbon from the vacuum pump drain). These weights are added into the total vac. col. overhead weight, but the samples are not physically added to the overhead sample that goes to the lab for analysis. The only time that this may be a problem is when the plant is being run at very high severities, in which case substantial quantities of light hydrocarbon may end up in the vacuum pump oil. In this case the selective removal of light hydrocarbons may skew the analysis of the total overhead material.

Let:

$$\begin{aligned} \text{GOH} &= \text{grams of total overhead product} = \text{grams adjusted overhead} \\ &\quad (\text{already includes debutanizer bottoms and other corrections}) + \\ &\quad \text{grams scrubber hydrocarbon} \end{aligned}$$

The total overhead product is analyzed for API (D4052), wt-% carbon, hydrogen, and nitrogen (by UOP 866), wt-% sulfur (by U864), and boiling range by GLC (D2887). The samples may be checked occasionally for heptane insolubles, MCRT, or metals, but normally these values are negligible and are assumed to be zero. Because of periodic problems with the nitrogen analysis, Kjeldahl nitrogen analyses (U384) have also been

run as an occasional check on the U866 numbers.

GOHC = GOH * (wt-% C in OH) = grams overhead carbon
GOHH = GOH * (wt-% H in OH) = grams overhead hydrogen
GOHN = GOH * (wt-% N in OH) = grams overhead nitrogen
GOHS = GOH * (wt-% S in OH) = grams overhead sulfur

GOH177+ = GOH * (100% - wt-% over at 177°C)/100% = grams 177°C+
material in overhead product
GOH343+ = GOH * (100% - wt-% over at 343°C)/100% = grams 343°C+
material in overhead product
GOH371+ = GOH * (100% - wt-% over at 371°C)/100% = grams 371°C+
material in overhead product
GOH510+ = GOH * (100% - wt-% over at 510°C)/100% = grams 510°C+
material in overhead product

The total bottoms product from the vacuum column is weighed. In a twelve hour test period, two graduates of product are typically obtained. One portion is worked up by the technician. The sample is separated, using toluene solvent and centrifugation at 3000 rpm, into toluene soluble bottoms (TSB) and toluene insoluble bottoms (TIB) fractions. The toluene is removed from the TSB using a rotary flash evaporator. The TIB fraction is dried in a vacuum oven to remove any residual solvent.

Let:

GTBP = measured grams/period of total vacuum column bottoms product
GBSEP = grams of vacuum bottoms used in solvent separation
step
GTSBSEP = grams of toluene soluble bottoms recovered from separation
step

GTIBSEP = grams of toluene insoluble bottoms recovered from separation
 step

If (GTIBSEP+GTSBSEP) > GBSEP, we assume that not all of the toluene has been removed from the TSB fraction, and decrease the TSB fraction accordingly to give 100% mass balance around the separation step. (If toluene is left in the TSB, you can usually see it in the front end of the GLC, which shows an IBP and several per cent at 110°C, before jumping up 200 or more °C to the vacuum bottoms boiling range.)

If (GTIBSEP+GTSBSEP) < GBSEP, we assume that the losses are due to handling and prorate them over the two separated fractions to give 100% mass balance around the separation step.

Then:

GTSB = (GTSBSEP/(GTSBSEP+GTIBSEP))*GTBP = grams of toluene
 soluble bottoms product

GTIB = (GTIBSEP/(GTSBSEP+GTIBSEP))*GTBP = grams of toluene
 insoluble bottoms product

The toluene soluble bottoms fraction is analyzed by U866 for wt-% carbon, hydrogen, and nitrogen, U864 for wt-% S, U614 for wt-% heptane insolubles, D4530 for MCRT, and AAS for metals.

Then:

GTSBC = GTSB * (wt-% C in TSB) = grams TSB carbon

GTSBH = GTSB * (wt-% H in TSB) = grams TSB hydrogen

GTSBN = GTSB * (wt-% N in TSB) = grams TSB nitrogen

GTSBS = GTSB * (wt-% S in TSB) = grams TSB sulfur

GTSBHI = GTSB * (wt-% HI in TSB) = grams heptane insolubles in toluene soluble bottoms

GTSBMC = GTSB * (wt-% MCRT in TSB) = grams micro-carbon in toluene soluble bottoms

The toluene soluble bottoms product is distilled by D1160 into overhead and bottoms fractions, with the temperature raised from the IBP to at least 510°C. The D1160 overhead fraction is then further analyzed by GLC (D2887) to give a boiling range breakdown at the desired yield points (typically 177, 343, 371 and 510°C).

Let:

GDO = Grams (or wt%) of TSB taken overhead in D1160

GDB = Grams (or wt%) of TSB remaining undistilled in D1160.

And:

GTSB177+ = $(100\% - \text{wt-\% over at } 177^{\circ}\text{C in D1160 OH GLC}) / 100\% * \text{GDO} / (\text{GDO} + \text{GDB}) * \text{GTSB}$ = grams 177°C+ material in toluene soluble bottoms product

GTSB343+ = $(100\% - \text{wt-\% over at } 343^{\circ}\text{C in D1160 OH GLC}) / 100\% * \text{GDO} / (\text{GDO} + \text{GDB}) * \text{GTSB}$ = grams 343°C+ material in toluene soluble bottoms product

GTSB371+ = $(100\% - \text{wt-\% over at } 371^{\circ}\text{C in D1160 OH GLC}) / 100\% * \text{GDO} / (\text{GDO} + \text{GDB}) * \text{GTSB}$ = grams 371°C+ material in toluene soluble bottoms product

GTSB510+ = $(100\% - \text{wt-\% over at } 510^{\circ}\text{C in D1160 OH GLC}) / 100\% * \text{GDO} / (\text{GDO} + \text{GDB}) * \text{GTSB}$ = grams 510°C+ material in toluene soluble bottoms product

The Toluene Insoluble Bottoms Product (TIB) is analyzed for elemental breakdown (C/H/N by U866, S by U864, and metals by AAS).

GTIBC = GTIB * (wt-% C in TIB) = grams TIB carbon
 GTIBH = GTIB * (wt-% H in TIB) = grams TIB hydrogen
 GTIBN = GTIB * (wt-% N in TIB) = grams TIB nitrogen
 GTIBS = GTIB * (wt-% S in TIB) = grams TIB sulfur

This fraction contains coal minerals, catalyst, and unconverted MAF coal. The TIB is ashed at 700°C by the lab technician. The ash fraction must be subtracted out to get the amount of unconverted coal.

GTIBA = GTIB * (wt-% Ash in TIB) = grams TIB Ash.
 GUCC = GTIB - GTIBA = grams unconverted coal

HYDROGEN CONSUMPTION

The hydrogen consumption is calculated as the difference between the makeup hydrogen used to maintain plant pressure and the hydrogen in the net gas streams. The input hydrogen is measured by a positive displacement meter, and the output hydrogen is measured by wet test meter.

(g Hydrogen Consumption) = {[(cycles on H₂ meter) * (meter factor, 0.17 nCF/cycle) * (2.016 g/gmole H₂) * (453.6 gmole/lbmole) / (359 nCF/lbmole) } - [(g H₂ in RG) + (g H₂ in DG)]

WEIGHT RECOVERY AND COMPONENT BALANCES

The overall weight balance is used to monitor pilot plant performance and check for leaks.

(Overall Weight Recovery) = $[(GTBP)+(GOH)+(g\ H_2O)+(g\ H_2S)+(g\ NH_3) + (g\ Gas\ CO_x)+(g\ Gas\ C_1\ to\ C_6)]/[(g\ MAF\ Coal)+(g\ Resid)+(g\ Catalyst)+(g\ Hydrogen\ Cons.)]*100\%$.

Weight recoveries and component balances on a moisture and ash free feed (MAFF) basis are calculated as:

(MAF Weight Recovery) = $[(g\ TLP)+(g\ UCC)+(g\ H_2O)+(g\ H_2S)+(g\ NH_3) + (g\ Gas\ CO_x)+(g\ Gas\ C_1\ to\ C_6)]/[(g\ MAF\ Coal)+(g\ Resid)+(g\ Hydrogen\ Cons.)]*100\%$.

(Carbon Balance) = $(GGASC+GOHC+GTSBC+GTIBC)/(GFEEDC)*100\%$
 (Hydrogen Balance) = $[(GGASH+GOHH+GTSBH+GTIBH)/(GFEEDH+GH2CONS)]*100\%$
 (Sulfur Balance) = $[(GGASS+GH2OS+GOHS+GTSBS+GTIBS)/(GFEEDS+GCATS)]*100\%$
 (Nitrogen Balance) = $(GGASN+GOHN+GTSBN+GTIBN)/(GFEEDN)*100\%$
 (Ash Balance) = $100\%*(GTSBA+GTIBA)/(W\%MFCA*GMFC+W\%RESA*GRESID+W\%CATA*GCAT)$
 (Metals Balance) = $[(GTSB*ppm\ metal\ in\ TSB/10^6) + (GTIB*wt-\% metal\ in\ TIB)]/[(GCOAL*ppm\ metal\ in\ coal/10^6)+(GRESID*ppm\ metal\ in\ resid/10^6) + (GCAT*wt-\% metal\ in\ catalyst)]*100\%$

Note that the nitrogen balance should always be 100%. (We made that assumption earlier in order to back out the ammonia production.)

COAL CONVERSION

The unconverted coal (UCC) is measured as the difference between the TIB and TIB ash.

$$g \text{ UCC} = (100\% - \text{TIB wt-\% Ash}) / 100\% * (g. \text{ TIB})$$

Depending on the quality of the ash balance, there can be a great deal of variability in the unconverted coal values calculated by this method. Optionally, the TIB ash may be back-calculated from the feed ash (less TSB ash) by assuming that there is a 100% ash balance. At present, the calculation procedure does not automatically ash balance the plant. The ash balance is reported, and if necessary, the TIB ash can be altered manually to force the balance.

Coal Conversion is based on the disappearance of moisture and ash free coal (MAFC):

$$(\text{wt-\% Coal Conversion}) = (g \text{ MAFC} - g \text{ UCC}) / (g \text{ MAFC}) * 100\%$$

To eliminate discrepancies in the conversions due to poor weight balances, the unconverted coal yield is normalized by the MAF weight recovery:

$$\text{Normalized wt-\% Coal Conv.} = [(g \text{ MAFC} - (g \text{ UCC}) / (\text{wt-\% MAF recovery}) * 100\%)] / (g \text{ MAFC}) * 100\%$$

HEPTANE INSOLUBLE CONVERSION

The heptane insoluble (HI) conversion is defined as the disappearance of heptane insolubles (including coal):

$$(\text{wt-\% HI Conversion}) = [(G\text{FEEDHI}) - (G\text{TSBHI} + G\text{UCC})] / (G\text{FEEDHI}) * 100\%$$

$$(\text{Normalized wt-\% HI Conv.}) = [(G\text{FEEDHI}) - (G\text{TSBHI} + G\text{UCC}) * 100\% / (\text{wt-\% MAF$$

$$\text{Rec.})]/(\text{GFEE DHI}) * 100\%$$

NONDISTILLABLE CONVERSION

The nondistillable (ND) conversion is defined as the disappearance of material boiling above 510°C (including coal):

$$(\text{wt-\% ND Conversion}) = [(\text{GFEE D 510+}) - ((\text{GOH 510+}) + (\text{GTSB 510+}) + (\text{GUCC}))]/(\text{GFEE D 510+}) * 100\%$$

$$(\text{Normalized wt-\% ND Conversion}) = \frac{\{(\text{GFEE D 510+}) - ((\text{GOH 510+}) + (\text{GTSB 510+}) + (\text{GUCC}))\}}{(\text{wt-\% MAF Rec.}) * 100\%} / (\text{GFEE D 510+}) * 100\%$$

MCRT CONVERSION

The ConCarbon (MCRT) conversion is defined as the disappearance of ConCarbon (including coal):

$$(\text{wt-\% MCRT Conversion}) = [(\text{GFEE DMC}) - (\text{GTSBMC} + \text{GUCC})]/(\text{GFEE DMC}) * 100\%$$

$$(\text{Normalized wt-\% MCRT Conversion}) = \frac{[(\text{GFEE DMC}) - (\text{GTSBMCRT} + \text{GUCC})]}{(\text{wt-\% MAF Rec.}) * 100\%} / (\text{GFEE DMC}) * 100\%$$

PRODUCT YIELDS

All of the product yields are represented on a MAFF (moisture and ash free feed) basis. The total product yield will therefore exceed 100% by the amount of the hydrogen consumption. The C₅ and C₆ components from the gas streams are included in the

naphtha yield. The yields are normalized by the total MAF weight recovery.

$$(\% \text{yield}_i) = 100\% * (\text{g. product}_i) / (\text{GMAFF}) * 100\% / (\text{wt\% MAF recovery})$$

$$(\text{wt-\% H}_2\text{O} + \text{CO}_x \text{ yield}) = (\text{GH}_2\text{O} + \text{GCOX}) / \text{GMAFF} * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% H}_2\text{S yield}) = \text{GH}_2\text{S} / \text{GMAFF} * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% NH}_3 \text{ yield}) = \text{GNH}_3 / \text{GMAFF} * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% LE yield}) = (\text{GGASC1} + \text{GGASC2} + \text{GGASC2}^+ + \text{GGASC3} + \text{GGASC3}^+ + \text{GGA} \\ \text{SiC4} + \text{GGASnC4} + \text{GGASC4}^+) / \text{GMAFF} * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% Naphtha yield}) = [(\text{GGASiC5} + \text{GGASnC5} + \text{GGASC6}^+) + (\text{GOH} - \text{GOH177}^+) + (\text{GTS} \\ \text{B} - \text{GTSB177}^+)] / \text{GMAFF} * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% Dist. yield}) = [(\text{GOH177}^+) - (\text{GOH343}^+) + (\text{GTSB177}^+) - (\text{GTSB343}^+)] / \text{GMAFF} \\ * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% VGO yield}) = [(\text{GOH343}^+) - (\text{GOH510}^+) + (\text{GTSB343}^+) - (\text{GTSB510}^+)] / \text{GMAFF} \\ * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% Resid yield}) = [(\text{GOH510}^+) + (\text{GTSB510}^+)] / \text{GMAFF} * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% UCC yield}) = \text{GUCC} / \text{GMAFF} * 100\% / (\text{MAFREC}) * 100\%$$

$$(\text{wt-\% H}_2 \text{ Consumption}) = \text{GH}_2\text{CONS} / \text{GMAFF} * 100\% / (\text{MAFREC}) * 100\%$$

AVERAGE PRODUCT PROPERTIES

For the purposes of comparing the bench-scale unit results to data from the autoclave or from other processes that use different product cut-points, product properties are presented as weighted averages for the total liquid product:

$$(\text{TLP } ^\circ \text{API}) = (\text{OHAPI} \cdot \text{GOH} + \text{TSBAPI} \cdot \text{GTSB}) / (\text{GOH} + \text{GTSB})$$

$$(\text{TLP wt-\% C}) = (\text{GOHC} + \text{GTSBC}) / (\text{GOH} + \text{GTSB}) * 100\%$$

$$(\text{TLP wt-\% H}) = (\text{GOHH} + \text{GTSBH}) / (\text{GOH} + \text{GTSB}) * 100\%$$

$$(\text{TLP wt-\% S}) = (\text{GOHS} + \text{GTSBS}) / (\text{GOH} + \text{GTSB}) * 100\%$$

$$(\text{TLP wt-\% N}) = (\text{GOHN} + \text{GTSBN}) / (\text{GOH} + \text{GTSB}) * 100\%$$

$$(\text{TLP wt-\% HI}) = (\text{GTSBHI}) / (\text{GOH} + \text{GTSB}) * 100\%$$

$$(\text{TLP wt-\% MC}) = (\text{GTSBMC}) / (\text{GOH} + \text{GTSB}) * 100\%$$

DESULFURIZATION

The Desulfurization is defined as the disappearance of sulfur in the total liquid product relative to the coal and resid feeds:

$$(\text{wt-\% Desulfurization}) = [(\text{GFEEDS}) - (\text{GOHS} + \text{GTSBS})] / (\text{GFEEDS}) * 100\%$$

DENITRIFICATION

The Denitrification is defined as the disappearance of nitrogen in the total liquid product relative to the coal and resid feeds:

$$(\text{wt-\% Denitrification}) = [(\text{GFEEDN}) - (\text{GOHN} + \text{GTSBN})] / (\text{GFEEDN}) * 100\%$$