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EDS COAL LIQUEFACTION PROCESS DEVELOPMENT

Phase V

Quarterly Technical Progress Report January 1—March 31, 1984

July 1984
Date Submitted

Work Performed Under Contract No. FC05-77ET10069

Exxon Research and Engineering Company
Florham Park, New Jersey

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EDS COAL LIQUEFACTION PROCESS DEVELOPMENT

PHASE V

Quarterly Technical Progress Report
for the Period
January 1 - March 31, 1984

W. R. Epperly - Project Director

Exxon Research and Engineering Company
P. O. Box 101
Florham Park, New Jersey 07932

PREPARED FOR
UNITED STATES DEPARTMENT OF ENERGY
UNDER COOPERATIVE AGREEMENT NO. DE-FC05-77ET10069

Submitted - July, 1984

A Project Jointly Sponsored by
United States Department of Energy
Exxon Company, U.S.A.
Electric Power Research Institute
Japan Coal Liquefaction Development Company
Phillips Coal Company
Anaconda Minerals Company
Ruhrkohle AG
ENI

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EDS COAL LIQUEFACTION PROCESS DEVELOPMENT - PHASE V
QUARTERLY TECHNICAL PROGRESS REPORT FOR THE PERIOD
JANUARY 1-MARCH 31, 1984

Abstract

This report is the twenty-first Quarterly Technical Progress Report for U.S. Department of Energy Cooperative Agreement No. DE-FC05-77ET10069 for EDS Coal Liquefaction Process Development Phase V. This report covers the period January 1 through March 31, 1984. Funding is shared by U.S. Department of Energy, Exxon Company, U.S.A. (a division of Exxon Corporation), Electric Power Research Institute, Japan Coal Liquefaction Development Company, Phillips Coal Company, Anaconda Minerals Company, Ruhrkohle AG, and ENI. The agreement covers the period January 1, 1977 through December 31, 1985. The Laboratory Process Research and Development studies were conducted at various Exxon Research and Engineering Co. (ER&E) facilities: Research and Development Division at Baytown, Texas; Products Research Division at Linden, New Jersey; and the Exxon Research and Development Laboratories at Baton Rouge, Louisiana. The Engineering Research and Development studies were performed at the Exxon Engineering Petroleum and Synthetic Fuels and Exxon Engineering Technology Departments of ER&E at Florham Park, New Jersey. The information dealing with operation of the 250 T/D EDS Coal Liquefaction Pilot Plant (ECLP) was generated at Exxon Company, U.S.A., Houston, Texas.

Highlights from this report are summarized as follows. They are grouped according to their corresponding reporting categories in the report.

ENGINEERING RESEARCH AND DEVELOPMENT
PROCESS ENGINEERING AND COST EVALUATIONS

1. Cost Reduction and Laboratory Guidance Studies

- An engineering screening evaluation examining the relative economics of drying coal feed to an EDS liquefaction plant with gas-swept mill and slurry drying technologies was completed. The major conclusion from this work is that slurry drying is the preferred drying option for all EDS sponsor coals.
- Economic screenings of Texas lignite were completed on a preliminary basis and results are presented. The lowest cost plant configuration for this coal was identified, and it was also determined that incentives exist to maximize conversion. Results for several sensitivities are also presented.
- A detailed comparison of RCLU, CLPP, and ECLP yields has been initiated. This study builds off previous yield modeling results, which found that RCLU, CLPP, and ECLP yields were generally consistent given the scatter of the data, although some differences were noted. These pilot unit yield differences have now been quantified, and operating/configurational differences which account

for some of them have been identified. Preliminary yield comparison results after correcting for these known process differences between the pilot plants indicate that:

- RCLU and CLPP yields are generally consistent.
- ECLP's conversion is about 5 lb/100 lb DAF coal lower than RCLU/CLPP at comparable operating conditions.
- Work has been initiated to define the EDS slurry preheater feed system design. The study is based on slurry distributor manifold guidelines and coking correlation predictions, which influence furnace pass control issues such as slurry flow measurement.

EDS TOXICOLOGY PROGRAM

2. Analytical Characterization and Bioassay

- EDS hydrotreated naphtha showed a low level of systemic toxicity to rats exposed to the vapor six hours per day, five days per week for thirteen weeks.

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EDS COAL LIQUEFACTION PROCESS DEVELOPMENT - PHASE V
QUARTERLY TECHNICAL PROGRESS REPORT FOR THE PERIOD
JANUARY 1 - MARCH 31, 1983

This report is the twenty-first Quarterly Technical Progress Report for U.S. Department of Energy Cooperative Agreement No. DE-FC05-77ET10069 for the EDS Coal Liquefaction Process Development - Phase V. This agreement covers the period January 1, 1977 to December 31, 1985, and is a continuation of DOE Contract No. E(49-18)-2353-Phase IIIA covering the period January 1, 1976 to June 30, 1977. The six-month overlap in funding in the first part of 1977 applies to activities related to the 250 T/D EDS Coal Liquefaction Pilot Plant (ECLP), constructed during Phase IV of the project.

Prior development of the EDS Process was conducted during the ten-year period 1966 through 1975. During this period, the EDS process was selected over several other options as the preferred route to the production of coal liquids. This work was funded entirely by Exxon and was divided into Phase I, the Predevelopment Phase, and Phase II, the Planning and Design Phase. A summary of this work is given in the report, FE-2893-16 "Summary of EDS Predevelopment (1966-1975)." Phase IIIA was a continuation of the Phase I and II programs, and was divided into the following two tasks:

- Laboratory Process Research and Development, and
- Engineering Research and Development

A description of the work performed in Phase IIIA can be found in the reports cited.^(1,2) Phase IIIB which began on July 1, 1977 and Phase IV which began on January 1, 1978, include the above two tasks, as well as tasks relating to project management, detailed engineering, procurement, and construction of ECLP. A final Phase IV report covering the project management, engineering/procurement, construction, and preparation for operations of the large EDS pilot plant (ECLP) at Baytown, Texas, has been issued⁽³⁾. Phase V began on April 1, 1980, upon mechanical completion of ECLP and start of operations.

- (1) January-December 1976, EDS Annual Technical Progress Report (FE-2353-9)
- (2) January 1976-June 1977, EDS Final Technical Progress Report (FE-2353-20)
- (3) Construction of Exxon Coal Liquefaction Pilot Plant, Final Report, March 1981 (FE-2893-66)

ENGINEERING RESEARCH AND DEVELOPMENT
PROCESS ENGINEERING AND COST EVALUATIONS

1. Cost Reduction and Laboratory Guidance Studies

1.1 Comparison of Gas-Swept Mill with Slurry Dryer for Coal Drying

An engineering screening evaluation which examined the relative economics of drying coal feed to an EDS liquefaction plant with gas-swept mill (GSM) and slurry drying technologies was completed. This study used the EDS Process Alternatives LP Model (PAM II) to compare the two coal drying systems in several plant configurations using bottoms recycle operations. EDS plants feeding Illinois No. 6 (Monterey mine), Wyoming (Wyodak mine) and Texas lignite (Martin Lake mine) were studied.

Early results for Illinois and Wyoming coals were reported in the July 1, 1981-June 30, 1982 EDS Annual Technical Progress Report (FE-2893-93). The recent work, wrapping up this effort, involved reexamining both drying schemes in light of ECLP learnings, reevaluating a Wyoming POX/HB configuration, and screening a Texas lignite (30 wt% moisture) POX/C₂ configuration. Results show that the relative economics of slurry drying have improved significantly since the earlier screenings. The major conclusion from this work is that slurry drying is the preferred drying option for EDS sponsor coals which all contain less than about 30 wt% moisture.

Gas-Swept Mill/Slurry Dryer Designs Modified to Reflect ECLP Learnings

The early GSM slurry drying screenings assumed that processing coal in a GSM which used flue gases as hot as 900°F to dry the coal did not adversely affect liquefaction yield. However, extensive analyses of ECLP operations, plus smaller pilot unit testing, indicated that yield losses could occur with Wyoming coal and Texas lignite, and that flue gas inlet temperatures should be reduced to 550°F to prevent these yield losses.

The GSM design was modified in accordance with this lower flue gas-inlet temperature in both of the updated screenings of Wyoming coal and Texas lignite. The original design for Wyoming required eight GSM units to process liquefaction feed coal, however, due to the increased volume of flue gas now required with the lower inlet temperature, and a better understanding of GSM maximum capacities, the number of GSM units has more than doubled to 20 (16N, 4S). The overall investment associated with the GSM doubled; partly because of the increase in number of units and partly because of refinements in cost estimating practices. Flue gas desulfurization was also included as part of the coal-fired design. The GSM system for drying Texas lignite is identical to that used for Wyoming since both have similar moisture levels.

A vapor disengagement vessel was added to the slurry dryer design at minimal added cost. Demonstrations at ECLP showed slurry dryer performance to improve dramatically after the installation of this vessel. This vessel reduces slurry pumparound pump cavitation by allowing the water vapor entrained in the slurry to escape and by minimizing the evolution of additional water vapor in the suction line to the pump.

Bases for Study

The recent slurry drying and GSM cases were compared using PAM II. The major basis items for this study include:

- One liquefaction operating condition for each coal:

	<u>Wyoming*</u>	<u>Texas Lignite</u>
Solvent/Coal/Bottoms/VGO	1.6/1.0/0.5/0.2	1.2/1.0/1.0/0.0
Average Temperature	840°F	820°F
Nominal Residence Time	60 min	40 min
H ₂ Partial Pressure	2000 psi	2000 psi

* Wyoming Coal Bottoms Recycle Study Design conditions

- One plant configuration for each coal:

	<u>H₂ Source</u>	<u>Fuel Source</u>
Wyoming*	VB water slurry POX	Hybrid Boiler
Texas Lignite	VB water slurry POX	C ₂

* Wyoming Coal Bottoms Recycle Study Design configuration

- Assumed coal properties:

	<u>Wyoming</u>	<u>Texas Lignite</u>
As-received coal moisture	31 wt%	30 wt%
Dry coal feed ash level	7.35 wt%	14.5 wt%

- Economic bases where Required Initial Selling Price (RISP) was calculated with C₁/C₂ in Btu parity with coal liquids (i.e., high gas value). Product recovery factors were 85% for the Wyoming cases and 87% for the Texas lignite cases.
- Coal feed rate to liquefaction was set at 25 kST/SD (dry basis), and supplemental coal was used if needed.
- Sensitivities of plant economics to GSM combustion fuel (gaseous versus coal).
- Sensitivities of plant economics to GSM heat integration with the liquefaction block.
- Ranking of drying alternatives by the overall delta between GSM and slurry drying cases regardless of configuration or coal type.

Revised Drying Schemes for Wyoming Are Economic Standoff

For the EDS Wyoming Coal Bottoms Recycle Study Design configuration of slurry bottoms POX for hydrogen/solid bottoms plus coal-fired hybrid boiler for process heat and steam, the slurry drying and GSM schemes were an economic standoff.

The GSM is competitive with slurry drying only if the GSM is fired with coal. Firing the GSM with C_2^- , produced during liquefaction, increases RISP by 8% over the coal-fired case because a smaller quantity of potentially valuable C_2^- is available for sale.

Overall, the plant investments for a coal-fired GSM case are about 3% lower (SDU basis) than those for the slurry drying base case. While the investments for the GSM units are more than twice that of slurry dryers, large savings are realized in smaller boiler and wastewater treating facilities. The boiler size is reduced substantially because GSM drying does not consume any steam, and wastewater treating is reduced because the evaporated coal moisture is typically vented to the atmosphere instead of being condensed as in the slurry drying base case. Burning coal in the GSM combustor does increase the liquefaction coal ash level slightly; however, the effect on overall plant investment is negligible.

To better understand the relationship between GSM and slurry drying, Figure 1-1 was developed to show drying costs associated with a Wyoming-type coal as a function of coal moisture level. Note that different moisture levels were assumed in order to determine the effect on overall plant costs. It was also assumed that coal cost, ash level, heating value and sulfur content did not change with varying moisture levels. Cost lines for both coal-fired GSM and slurry drying schemes are shown. At the lower coal moisture levels, slurry drying is clearly the preferred drying method; however, as coal moisture level is increased above 30 wt%, the GSM system may become the lower cost drying method.

At the lower coal moisture levels, slurry drying is less costly because most of the energy for drying is contained in the hot recycle solvent and bottoms streams. Little supplemental heat in the form of condensing steam in the slurry pumparound loop is needed. The costs for slurry drying increase at a faster rate with increasing coal moisture level than those for GSM drying primarily because larger boilers are needed to supply the additional heat needed in the pumparound loop. The incremental costs of larger boilers are greater than those for larger GSM units. As the coal moisture level is increased further, an additional factor affects the relative costs of these drying methods. The stripping action of water vapor leaving the slurry dryers vaporize increasing amounts of recycle solvent. For example, a slurry dryer operating with 45 wt% moisture coal strips out about 34% of the solvent as compared to 23% in the Wyoming Coal Bottoms Recycle Study Design using 31 wt% moisture coal. This increased solvent vaporization not only increases the size and cost of the dryer overhead system, but also increases the cost associated with the additional steam needed in the pumparound loop to make up for the "lost" heat of vaporization of the solvent. This is shown by the slightly increasing slope of the slurry dryer cost curve.

Figure 1-1

DRYING COSTS DEPEND ON COAL MOISTURE LEVEL*

*Wyoming Wyodak Coal Properties and Cost Used With Assumed Moisture Level Variations
RISP = Required Initial Selling Price

This graph can be used to generalize that slurry drying is the preferred method of drying other EDS sponsor coals. The respective moisture levels for these coals are 17 wt% for Illinois No. 6 Monterey, 10 wt% for Illinois No. 6 Burning Star, 4 wt% for West Virginia Pittsburgh No. 8 Ireland, and 10 wt% for Queensland, Australia Wandoan.

Texas Lignite Results Consistent with Those for Wyoming

The substantial reduction in plant steam demand associated with the GSM case (see Wyoming results discussion) has a major impact in the GSM/slurry drying Texas lignite comparison. The POX/C₂ configuration studied for the lignite can only utilize the solidified vacuum bottoms to produce plant hydrogen and steam. The steam requirements for the GSM case were reduced to the point where net vacuum bottoms production exceeded overall plant needs, and since selling a bottoms product is not a reasonable alternative for EDS, the only feasible route was to send the excess bottoms to the boilers to produce excess steam. This steam (65% excess) was used in condensing turbines to provide plant power. Past studies have shown that producing excess steam to supply plant power is not economical compared with purchasing power. Correspondingly, the Texas lignite GSM/slurry drying comparison showed that slurry drying had a 2% incentive over GSM drying.

The excess steam situation masks the actual GSM/slurry drying comparison, therefore, a sensitivity case was developed which studied a GSM case that did not produce excess steam. This case showed that both drying schemes were an economic standoff, and this is consistent with the results from Wyoming comparisons. There are several ways in which the excess bottoms can be utilized. A change to a plant configuration, such as POX/POX or POX/HB will provide additional uses for the bottoms, or overall plant conversion can be increased to reduce bottoms make and increase hydrogen requirements. Bottoms may also be used as the fuel supply for the GSM combustors.

The base GSM case assumed that excess heat contained in the recycle solvent stream could not be used as an additional slurry preheat prior to liquefaction. This assumption was based on design considerations which identified potential design concerns that would be associated with the required heat exchanger. A sensitivity case was developed which assumed that this heat exchanger could be designed, and the resulting thermal efficiency credits gave the GSM case a 2% incentive over the slurry dryer case. Earlier Wyoming findings showed similar incentives.

Limitations of Present Study

As with any study, concurrent or future learnings in related areas could affect results and conclusions. This section discusses some of the known areas which could have an impact on drying coal for an EDS plant.

A slurry drying design study is currently planned as part of the EDS Consolidation Program. This detailed work could lead to increased costs for slurry drying primarily in the area of equipment costs. This work could also identify areas where savings can be realized. For instance, the recent Wyoming Coal Bottoms Recycle Study Design Addendum used a higher boiling point solvent than previous studies. This could result in less solvent vaporization in the dryer vessel, and a corresponding decrease in dryer overhead

investments and operating costs. If slurry dryer costs do increase, it is estimated that the RISP for this case, at the extreme, could increase by about 2.5%.

The costs associated with the modified GSM design can increase further if it is judged that additional modifications are required to safely operate a GSM system. Some potential problem areas include; flue gas inlet temperature control, flue gas condensation in GSM ducting, and coal dust explosion prevention.

Overall, the GSM system is judged to have a higher upside cost potential than slurry drying.

Conclusions

This engineering screening study indicates that for coals with low moisture levels (e.g., Illinois #6 Monterey mine with 17 wt% moisture), incentives exist for slurry drying relative to GSM because the heat contained in the hot recycle streams is effectively utilized for coal drying. As the coal moisture levels increase, the incentive for slurry drying decreases to the point where the GSM and slurry drying options become an economic standoff. This point occurs at about the moisture levels of Wyoming Wyodak (31 wt%) and Texas lignite (30 wt%). A variety of uncertainties associated with a GSM system could further increase its cost relative to slurry drying.

The overall conclusions for this GSM versus slurry drying engineering/economic evaluation are as follows:

- Slurry drying is the preferred coal drying method for all EDS Sponsor coals.
- There may be an incentive to use coal fired GSM drying for coals with greater than about 35 wt% moisture.
- Plant balances, especially the net bottoms balance, are major factors in determining overall plant economics.

This concludes the sponsor engineering/economic evaluation of the GSM and slurry drying technologies. Learnings from this work are being incorporated into the ongoing EDS Consolidation Program economic studies. No additional work is planned for the GSM drying scheme, while design studies are planned to further define the slurry dryer design based on ECLP operating data.

1.2 Economic Studies - Texas Lignite Screening

The EDS consolidation economic screening studies are underway to update five of the six Sponsor coal screening studies that have been carried out over the course of the EDS project. These studies will use the updated EDS Process Alternatives LP Model (PAM III) which reflects a state-of-the-art EDS design to compare economics with different coals at various liquefaction conditions in a commercial size EDS plant. The coals are bituminous coals from Illinois (No. 6 Monterey mine) and West Virginia (Pittsburgh No. 8 Ireland mine), sub-bituminous coals from Wyoming (Wyodak) and Queensland,

Australia (Wandoan) and lignite from Texas (Martin Lake mine). The result of these screenings will be a set of consistent cases that will allow inter-coal comparisons.

Screenings with Texas lignite have been completed and preliminary results for all cases are presented here. A plant configuration which uses water slurried coal and solidified vacuum bottoms partial oxidation for the production of plant hydrogen and fuel demonstrated the best overall plant economics. It was also determined that incentives exist to maximize liquefaction conversion. Along with cases examining configurational and conversion level effects, the impact of reforming naphtha onsite and the effect of feed ash level on economics were also determined. Finally, a comparison between the Texas and Wyoming locations shows that inter-coal case comparisons can be very sensitive to the assumed coal prices and relative costs for building plants in both locations.

BASES FOR STUDY

- RCLU model based yields, adjusted for expected commercial yields for three (reference, light product slate and high distillate) liquefaction operating conditions as shown in Table 1-1.
- Three configurations examined:
 - + Bottoms water slurry partial oxidation (POX) for H_2 , C_2^- for fuel.
 - + Bottoms and coal POX for H_2 , bottoms POX for intermediate Btu gas for fuel.
 - + Bottoms POX for H_2 , hybrid boiler (HB) for process heat and steam.
- All other basis items as published in October 1-December 31, 1983 Quarterly Technical Progress Report, FE-2893-129.

POX/POX PREFERRED EDS CONFIGURATION

The relative Required Initial Selling Prices (RISP) for three EDS plant configurations at the reference liquefaction operation are presented in Figure 1-2. When valuing the HBG product in Btu parity with coal liquids, the POX/POX configuration has the best overall economics followed by the POX/HB (+7.8%) and POX/ C_2^- (+10.5%) configurations, respectively. The major factor leading to a difference between the POX/ C_2^- and POX/POX configurations is the total amount of product available for sales. Potentially valuable product is being used as plant fuel in the POX/ C_2^- case, and since there is less product to spread the costs over, the RISP is higher. The POX/POX case actually has higher plant investment and operating costs (coal and power requirements as shown in Table 1-2) since it is a more complicated plant.

The POX/POX and POX/HB cases show very similar total costs since the slightly higher hybrid boiler plant investment is offset by lower operating costs (coal and electric requirements) relative to the all POX case as shown in Table 1-2. The major factor debiting the HB case is that its overall plant product recovery factor (PRF) is 84.4% as compared to the POX/POX and POX/ C_2^- PRFs of 90%. These PRFs were determined from detailed analyses of plant reliability, and the lower hybrid boiler PRF reflects the fact that a hybrid boiler outage will cause an entire liquefaction line to come down. Note that even with the debit, the POX/HB configuration still has better economics than

TABLE 1-1TEXAS LIGNITE OPERATING CONDITIONS/YIELDS

	<u>REFERENCE</u>	<u>LIGHTER SLATE</u>	<u>HIGH DISTILLATE</u>
Conditions			
Temp. (°F)	820	820	800
NRT (Min)	60	100	60
H ₂ Press (PSI)	2000	2000	2000
S/C/B/HVGO	1.4/1/1/0.2	1.8/1/1/0.2	1.4/1/1/0.2
Yields (wt% DAF Coal)			
H ₂	-4.5	-5.2	-3.9
Chem Gases	22.5	23.2	22.0
C ₁ /C ₂	6.2	7.1	3.4
C ₃ /C ₄	4.3	7.0	3.6
C ₅ /350°F	16.0	19.4	14.2
350/850°F	22.4	19.9	25.1
C ₄ /850°F	40.4	42.9	41.1
Conversion	67	71	64

FIGURE 1-2

RELATIVE RANKING OF EDS PLANT CONFIGURATIONS FOR TEXAS LIGNITE

TABLE 1-2
PAN REPORT SUMMARY

Liquefaction Operation:

	REFERENCE			LIGHTER PRODUCT SLATE		HIGH DISTILLATE REFERENCE		LIGHTER PRODUCT SLATE	
h ₂ /Fuel:	POX/C ₂ -	POX/POX	POX/HF	POX/C ₂ -	POX/POX	POX/C ₂ -	POX/C ₂ -	POX/C ₂ -	POX/POX
Special Notes:							18% Ash Feed	No Gas Sales	Naphtha to Powerformer
	TX2	TX2A	TX2B	TX3	TX3A	TX4	TX2ASH10	TX3NG5	TX3APP
MATERIAL BALANCE									
DRY COAL, KST/SD									
LIQUEFACTION	28.00	28.00	28.00	28.00	25.00	28.00	28.00	25.00	25.00
HYDROGEN GENERATOR		2.47		1.25	8.08			0.63	5.47
FUEL GENERATOR									
BOILERS	2.93	3.28	5.77	4.43	2.62	1.78	1.88	4.24	2.03
TOTAL	33.93	33.75	39.77	33.67	35.61	26.78	26.88	29.87	32.51
CONTAINED H ₂ , MSCF/SD									
HYDROGEN GENERATOR	385.71	389.71	389.71	453.72	453.72	340.71	353.85	453.72	399.31
CRYOGENIC RECOVERY	10.82	13.82	13.82	11.04	11.04	15.37	7.89	11.04	10.83
PRODUCT UPGRADING									
TOTAL	403.53	403.53	403.53	464.76	464.76	356.08	361.84	464.76	410.14
VACUUM ROTATIONS, KST/SD									
HYDROGEN GENERATOR	8.57	6.17	8.57	9.28	4.42	7.30	8.02	9.29	3.68
FUEL GENERATOR		4.13			4.87	1.44			5.62
BOILERS/FBC	1.73		1.73			2.11	2.56		
DISPOSAL									
TOTAL	10.30	10.30	10.30	9.28	9.29	10.85	11.58	9.29	9.29
LIQUID PRODUCTS, KB/SD									
C3 LPG	4.76	4.76	4.76	7.27	7.27	3.61	4.41	7.27	7.67
C4 LPG	4.76	4.76	4.76	8.01	8.01	4.28	4.19	8.01	8.57
C5/350 F. Naphtha	26.29	26.29	26.29	32.48	32.48	23.14	23.75	32.48	28.20
350 + Distillate	29.36	29.36	29.36	25.80	25.50	33.22	26.39	25.80	25.80
350 F. Hvy/Bunker Fuel									
Phenolics	0.67	0.67	0.67	0.67	0.67	0.67	0.67	0.67	0.67
TOTAL	65.85	65.85	65.85	73.93	73.93	64.92	59.42	73.83	70.61
LIQUID PRODUCT, KFOEB/SD *									
C3 LPG	2.86	2.86	2.86	4.37	4.37	2.17	2.65	4.37	4.60
C4 LPG	3.22	3.22	3.22	5.41	5.41	2.89	2.83	5.41	6.79
C5/350 F. Naphtha	21.67	21.67	21.67	26.77	26.77	19.07	19.58	26.77	23.50
350 + Distillate	28.00	28.00	28.00	24.32	24.32	31.68	25.16	24.32	24.32
350 F. Hvy/Bunker Fuel									
Phenolics	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57	0.57
TOTAL	56.32	56.32	56.32	61.44	61.44	56.38	50.79	61.44	58.78
HBG PRODUCT, KFOEB/SD *	0.95	11.58	8.70	1.78	13.55	0.0	0.09	0.0	13.95
TOTAL PRODUCT, KFOEB/SD *	57.27	67.87	65.02	63.22	75.00	56.38	50.88	61.44	72.73
AVE C3+ HT CONT, F0EB/B	0.86	0.85	0.86	0.83	0.83	0.87	0.85	0.83	0.81
BYPRODUCTS									
AMMONIA, ST/SD	249.67	249.67	249.67	273.47	273.47	235.55	222.18	278.82	273.47
SULFUR, LT/SD	153.21	194.08	163.21	190.03	222.07	159.99	148.58	185.88	218.63

* FOEB AT HT HEAT VALUE

TABLE 1-2 (Cont'd.)

	TX2	TX2A	TX2B	TX3	TX3A	TX4	TX2ASH18	TX3NGS	TX3APF
POWER, MW									
PURCHASED	407.70	521.87	418.46	475.18	614.73	404.66	390.23	471.11	615.89
GENERATED									
TURBINE	69.06	38.43	70.34	69.96	29.16	52.86	60.96	65.78	18.14
FUEL REQUIRED, KFOEB/SD #									
LIQUEFACTION	7.43	7.46		8.47	8.51	7.44	7.37	8.47	8.51
FRACTIONATION	1.31	1.31	1.31	1.27	1.27	1.33	1.35	1.27	1.27
SOLVENT HYDROGENATION	1.37	1.37	1.37	1.50	1.50	1.39	1.35	1.60	1.50
FLEXICOKING									
STEAM REFORMING									
PRODUCT UPGRADING	0.10	0.10	0.10	0.13	0.13	0.09	0.09	0.13	1.87
SULFUR RECOVERY									
BOILERS	16.28	11.56	5.05	15.60	8.25	13.71	13.79	14.94	7.16
HYBRID BOILERS/FBC			21.67						
OTHERS								0.01	
FLARE/VENT									
TOTAL	26.49	21.80	29.51	26.97	20.66	23.95	23.96	26.33	20.31
FUEL SUPPLY, KFOEB/SD #									
HIGH BTU GAS	10.21		2.78	11.37		6.89	10.17	11.39	
PURGE GAS									
INTERMEDIATE BTU GAS		10.24			11.41	3.65			13.15
LOW BTU GAS									
350 F+ FUEL OIL									
COAL	10.32	11.56	21.75	15.60	9.25	6.28	6.17	14.94	7.16
VACUUM BOTTOMS/COKE	5.96		4.98			7.42	7.62		
TOTAL	26.49	21.80	29.51	26.97	20.66	23.95	23.96	26.33	20.31
THERMAL EFFICIENCY, %	52.15	54.98	54.69	52.06	55.02	53.29	51.27	51.87	56.29
STM-1250 PRODUCTN, KLB/HR									
ON-SITE WASTE HEAT BOILER									
BOILERS	2832.31	1984.02	866.85	2676.31	1586.71	2400.72	2416.04	2564.01	1228.13
HYBRID BOILERS/FBC			2102.72						
PURCHASED									
TOTAL	2832.31	1984.02	2969.57	2676.31	1586.71	2400.72	2416.04	2564.01	1228.13
STM-1250 REQUIREMT, KLB/HR									
PROCESS REQUIREMENT	162.32	117.37	252.80	47.84	22.76	179.08		47.84	18.82
EXTRACT TURBINES/TURBOGEN	2645.55	1849.04	2678.84	2621.30	1560.54	2194.78	2416.04	2509.00	1206.37
CONDENSING TURBINES/TURBOGEN									
THROTTLED									
LOAD GROWTH	24.44	17.61	37.82	7.18	3.41	26.86		7.18	2.84
VENT									
TOTAL	2832.31	1984.02	2969.57	2676.31	1586.71	2400.72	2416.04	2564.01	1228.13
# FUEL AT LOW HEAT VALUE									

	TX2	TX2A	TX2B	TX3	TX3A	TX4	TX2ASH18	TX3NOS	TX3APF
STM-600 PRODUCTN, KLB/HR									
STEAM REFORMING									
FLEXICOKING									
PARTIAL OXIDATION									
OFFSITE BOILERS									
TURBINE DRIVES/TURBOGEN	1278.68	1278.53	1278.68	1207.29	1206.98	1234.11	1250.23	1201.70	1105.11
THROTTLED STM-250									
PURCHASED									
TOTAL	1278.68	1278.53	1278.68	1207.29	1206.98	1234.11	1250.23	1201.70	1105.11
STM-600 REOURMENT, KLE/HR									
LIQUEFACTION	849.58	849.58	849.58	715.00	715.00	849.58	849.58	715.00	715.00
FLEXICOKING									
STEAM REFORMING									
PARTIAL OXIDAT ON									
OTHER ONSITES	230.37	230.37	230.37	297.23	297.23	204.61	208.08	297.23	301.35
OFFSITE UNITS	29.34			32.68		18.85	28.23	32.72	
EXTRACT TURBINES/TURBOGEN									
CONDNG TURBINES/TURBOGEN									
THROTTLED									
LOAD GROWTH	166.78	166.76	166.78	187.47	187.43	160.97	163.07	156.74	158.21
VENT									
TOTAL	1276.07	1246.71	1246.73	1202.39	1169.66	1234.10	1249.97	1201.69	1174.56
STM-125 PRODUCTN, KLB/HR									
LIQUEFACTION/FRACTIONATN									
FLEXICOKING									
PARTIAL OXIDATION	1341.81	2193.08	1341.81	1607.36	2718.51	1443.23	1348.17	1562.05	2635.21
OTHER ONSITES/OFFSITES	220.38	220.35		208.07	208.02	212.70	215.47	207.11	240.25
TURBINE DRIVES/TURBOGEN	1366.87	570.51	1400.16	1414.00	353.55	960.67	1165.81	1307.30	101.26
THROTTLED STEAM									
PURCHASED									
TOTAL	2929.06	2983.93	2741.97	3229.44	3280.08	2616.60	2730.45	3076.46	2976.72
STM-125 REOURMENT, KLB/HR									
LIQUEFACTION	39.37	39.37	39.37	48.33	48.33	39.79	44.37	48.33	48.33
FRACTIONATION	55.01	55.01		53.44	53.44	55.86	57.01	53.44	53.44
SOLVENT HYDROGENATION	14.54	14.54	14.54	15.98	15.98	14.73	14.38	15.98	15.88
FLEXICOKING									
STEAM REFORMING									
PARTIAL OXIDATION	1709.14	1747.01	1709.14	2048.67	2100.52	1469.58	1557.79	1920.43	1850.88
OTHER ONSITES	647.80	672.81	361.69	552.78	581.21	631.76	609.29	551.72	578.98
OFFSITE UNITS	81.15	65.98	110.10	89.00	52.77	63.58	91.46	85.27	40.84
TURBINE DRIVES									
LOAD GROWTH	382.04	389.20	386.39	421.22	427.83	341.29	356.14	401.27	388.26
VENT									
TOTAL	2929.05	2983.93	2621.22	3229.43	3280.07	2616.59	2730.44	3076.45	2976.71

TABLE 1-2 (Cont'd.)

INVESTMENT, M\$(1985)*	TX2	TX2A	TX2B	TX3	TX3A	TX4	TX2ASH18	TX3NGS	TX3APF
LIQUEFACTION/FRACTIONATN	1458.39	1474.82	1282.44	1711.69	1728.48	1468.47	1369.73	1711.69	1728.46
SOLVENT HYDROGENATION	396.15	396.15	396.15	417.12	417.12	398.89	393.78	417.12	417.12
COMMON FACILITIES	333.88	333.88	333.88	333.88	333.88	333.88	333.88	333.88	333.88
CRYOGENIC H2 RECOVERY	122.62	122.62	122.62	140.03	140.03	102.67	117.66	140.03	143.16
PARTIAL OXIDATION for fuel		184.34			202.62				223.85
PARTIAL OXIDATION for H2	559.81	572.23	559.81	627.14	646.53	567.88	620.54	619.25	597.67
OXYGEN PLANT	351.63	445.17	351.63	392.54	494.82	358.16	333.50	386.16	484.69
PRODUCT UPGRADING	75.54	75.54	75.54	83.46	83.46	71.15	72.03	83.46	183.31
OTHER ONSITES	625.94	642.87	1363.61	646.51	663.78	611.01	621.51	642.31	663.10
COAL PREPARATION	130.61	130.23	37.23	138.02	145.67	127.44	127.71	135.89	142.83
OFFSITE BOILER/BFW TREAT	465.73	363.20	205.45	446.96	310.31	415.06	415.82	433.11	259.58
HYBRID BOILER/FBC									
WASTEWATER TREATING	259.85	270.84	201.99	260.57	274.28	260.81	250.65	257.28	272.30
UTILITIES	604.27	649.69	617.13	634.48	680.68	586.93	577.70	624.69	665.91
TANKAGE	200.15	201.23	200.15	245.73	246.89	183.98	185.73	246.93	251.57
OTHER OFFSITES	658.84	638.71	692.04	664.75	623.57	642.74	669.13	656.54	599.99
TOTAL	6243.42	6509.52	6339.67	6742.88	6993.08	6129.16	5989.38	6688.34	6968.43
Relative RISP (C3+ Basis, 1985)	+10.5	Base T	+7.8	+6.2	-4.1	+11.5	+17.7	+8.0	-1.7

*Assumes WBKSD Addendum basis item of 7.5% yearly investment escalation. This will be adjusted to reflect more accurate rate as part of consistency check.

the POX/C₂⁻ configuration when high Btu gas (HBG) is valued in Btu parity with coal liquids, primarily because the hybrid boiler frees up C₂⁻ for sales.

Figure 1-2 also shows the impact of valuing the HBG product at 50% Btu parity with coal liquids. The relative ranking of the cases has now changed with the POX/HB case now showing the highest overall RISP. For these Texas lignite cases, the gas value would have to drop to about 35% before the POX/POX and POX/C₂⁻ cases become economic standoffs.

Cases at the lighter product slate operation also show that the POX/POX configuration has a large incentive over the POX/C₂⁻ configuration (see Table 1-2).

INCENTIVES EXIST TO MAXIMIZE CONVERSION

The response of Texas lignite economics to conversion level is presented in Figure 1-3. Overall, the economics for the three scenarios of POX/POX and POX/C₂⁻ with and without gas sales are shown to improve with increasing conversion. This result that conversion should be maximized is consistent with results for the last sponsor coal screening of Pittsburgh (Ireland mine) coal (refer to July 1, 1982-June 30, 1983 Annual Technical Progress Report, FE-2893-116). The maximum conversion case corresponds to the lighter product slate operation. Conversion could be increased further by increasing average reactor temperature to 840°F. However, yield model predictions indicate that only the C₄⁻ yield would increase, by about 40%, with a corresponding increase in overall plant hydrogen requirements of 18%. Since the preferred plant products are liquids, a higher conversion case has not been simulated.

As conversion decreases, Figure 1-3 shows that the economics for the two POX/C₂⁻ configurations with and without gas sales approach each other. Basically, as conversion decreases, overall gas production falls faster than overall plant fuel requirements. At about 63% conversion, this configuration would fall short of plant fuel and would have to start consuming C₃⁺ product to satisfy plant needs. This would debit this configuration further. Figure 1-3 also shows that the RISP impact of limiting the plant to a no gas sales mode of operation for the POX/C₂⁻ configuration has little impact on overall RISP. This result was expected since this configuration typically does not have a large HBG product rate.

A sensitivity case examined the impact of catalytic reforming of the EDS naphtha material to produce a mogas blendstock. The configuration for this case was one that used partial oxidation to produce plant hydrogen and fuel and operated under the lighter product slate conditions. Results for this sensitivity are shown in Table 1-2. The RISP for this case is approximately 2.5% higher than the corresponding POX/POX case without reforming. This RISP increase is chiefly the result of an overall reduction in volume of product. Note that RISP does not distinguish between the value of alternative products. Thus, a premium value for the reformed naphtha could make this the preferred operation for the plant.

The RISP for the POX/C₂⁻ case at the reference liquefaction operation increased by almost 7% in a sensitivity case examining the impact of a higher ash feed coal to the plant. All other cases assumed a feed ash level

FIGURE 1-3

TEXAS LIGNITE ECONOMICS IMPROVE WITH CONVERSION

0484-Q-157

of 12 wt.% (dry basis) and the sensitivity case used 18 wt.%. Overall total plant costs are reduced in the high ash sensitivity with savings in overall plant investment and plant feed coal costs as shown in Table 1-2. The reduced plant product rate in this case, however, more than offsets those savings, resulting in the higher RISP. It is interesting to note that while overall plant C_2 make decreased, plant fuel requirements did not change with the result that the high ash case has virtually no HBG product sales.

COAL AND LOCATION COSTS AFFECT WYOMING/TEXAS COMPARISON

These Texas lignite results can be compared with the recently completed Wyoming Coal Bottoms Recycle Study Design (WBRSD) Addendum case as shown in Figure 1-4. This plot not only shows how the economics for the two coals compare in a POX/HB configuration, but it also shows how the relative economics can vary depending upon coal cost. Assuming 7.25 and 20.80 \$/ST as received coal costs for Wyoming Wyodak (WBRSD Addendum basis item) and Texas lignite respectively, the Texas lignite case has a 5% incentive over the Wyoming Addendum case. Even though the lignite's cost is almost triple that of Wyoming coal, the incentive for a Texas plant can be explained by examining the relative cost of building plants in the two locations. A Texas location has a relative total erected cost that is roughly 75% of that for the Wyoming location. This reflects differences in indirect costs, labor rates, productivity and site specific items such as winterization.

The delta between the two locations can change dramatically if the assumed coal costs are varied. For instance, as shown in Figure 1-4, a 30% increase in lignite cost would result in both locations being a standoff. Alternatively, if the Wyoming cost were to drop 30%, the delta between the locations would be only 3%. This plot can be used to examine the effects of varying both coal costs.

CONCLUSIONS/FUTURE WORK

The preliminary conclusions for the Texas lignite portion of the EDS consolidation economic studies are as follows:

- The EDS plant configuration that demonstrates the best overall economics is one that uses partial oxidation of water slurried coal and solidified vacuum bottoms for the production of plant hydrogen and fuel.
- The relative ranking of the plant configurations can vary depending on the value of the HBG product in relation to the other plant products.
- Incentives exist to maximize liquefaction conversion.
- Limiting the plant to a no gas sales mode of operations has little impact on RISP for a POX/ C_2 configuration.
- On a RISP basis, it is not preferred to produce a mogas blendstock by reforming naphtha. RISP, however, does not distinguish between the relative values of plant products and when this is considered the reforming option may be economically preferred.

FIGURE 1-4

RELATIVE ECONOMICS OF TEXAS AND
WYOMING PLANTS DEPEND ON COAL COST

- For every percentage point increase in coal feed ash level from 12 to 18 wt%, RISP also increases by about one percent.
- Comparisons between Texas and Wyoming locations can be sensitive to assumed coal costs.

These results and conclusions are preliminary in nature in that a comprehensive consistency check still needs to be performed. This consistency check will not be initiated until most of the coals have been screened, and once this check is completed absolute RISP values will be published.

Work is continuing on the Wyoming segment of the consolidation economic screening. All of the liquefaction operations and yields have been developed as have most of the bottoms processing vectors. Actual case screenings will be initiated shortly.

1.3 EDS Yield Modeling Efforts

As part of the EDS reactor modeling effort, a detailed comparison of RCLU, CLPP, and ECLP yields has been initiated. This study builds off previous yield modeling results, which found that RCLU, CLPP, and ECLP yields were generally consistent given the scatter of the data, although some small pilot unit yield differences were noted. These pilot unit yield differences have now been quantified. Also, operating and configurational differences which account for some of these yield differences have been identified. Preliminary results comparing the pilot plant yields, after correcting for these known process differences between the plants, are summarized below:

- RCLU and CLPP yields are generally consistent.
- ECLP's conversion is about 5 lb/100 lb DAF coal lower than RCLU/CLPP at comparable operating conditions.
- Most of this lower conversion is reflected in lower C_4 -1000°F yields.

Further evaluation of possible causes for these yield differences are continuing.

Background

In 1983, liquefaction yield scale-up correlations were developed to allow consistent prediction of the process yields for any size EDS plant. These correlations are based on RCLU data and use a model format nearly identical to the semi-empirical model developed for RCLU. The primary difference between the two models is the use of actual reactor residence time determined from tracer studies for the yield scale-up model, instead of nominal residence time (space time). This difference allows the scale-up correlations to predict the complete process yields for RCLU, CLPP, and ECLP on a consistent basis.

By using the yield scale-up model, yields for all three pilot units can be predicted within the same confidence interval, thus implying that RCLU, CLPP, and ECLP yields are generally consistent given the scatter of the

data. However, some small differences are evident within this data scatter. In particular, CLPP conversions tend to be higher than predicted and ECLP conversions tend to be lower. This is illustrated in Figure 1-5, which is a parity plot of predicted versus observed conversion for the three pilot units processing the three coals studied. For more information, see the July 1, 1982-June 30, 1983 Annual Technical Progress Report, FE-2893-116.

Pilot Unit Yield Differences Quantified

The observed yield differences between the pilot units have been quantified. These yield differences are small relative to the data scatter, generally being less than or equal to one standard deviation. Nonetheless, many of them are statistically significant because they occur consistently in a relatively large number of yield periods. Average pilot unit yield differences, their statistical significance, and typical data scatter limits are given in Table 1-3 (includes data on Illinois coal, Wyoming coal, and Texas lignite).

Some Yield Differences Explained by Operating and Configurational Differences

A number of operating, configurational, and analytical differences between the three pilot units have been identified which can affect process yields. These differences and their effect on pilot unit yields and yield comparisons, are described below:

Reactor Holdups - Due to differences in reactor gas/liquid velocities and diameter, the liquefaction reactors in RCLU, CLPP, and ECLP have different gas and liquid holdups. These differences were quantified using tracer tests and included in estimating the reactor slurry phase residence time at reaction conditions (actual residence time). Since the yield scale-up model is based on actual residence time, the yield differences listed in Table 1-3 have been corrected for this effect. For more information, see the July 1, 1982-June 30, 1983 Annual Technical Progress Report, FE-2893-116.

Reactor Backmixing - The RCLU and CLPP liquefaction reactors are nearly plug flow, whereas the ECLP reactors have a moderate amount of backmixing. The extent of this backmixing was determined from tracer tests. Its effect on ECLP's conversion was estimated as less than 1 lb/100 lb DAF coal debit, and the yield differences listed in Table 1-3 were corrected for this effect. See the October 1, 1982-December 31, 1982 Quarterly Technical Progress Report, FE-2893-103, for more information.

Coal Drying - Feed coal for both RCLU and CLPP is dried at moderate temperatures under nitrogen. The ECLP feed coal was usually dried in a gas-swept mill (GSM) which can expose the coal to an oxygen containing gas at higher temperatures. This use of GSM dried coal was shown to produce liquefaction conversion debits (mostly from the distillate yield) in laboratory tests using low rank coals. ECLP also showed GSM-dried coal conversion debits during some Wyodak coal operations. However, during most Wyodak coal operations, and all Monterey and Martin Lake coal operations, the GSM temperature was controlled to prevent conversion debits. Where required, the Wyodak yield periods used in developing Table 1-3 were corrected for this GSM debit for consistency.

FIGURE 1-5

RCLU/CLPP/ECLP 1000F+ CONVERSION PREDICTIONS
FALL WITHIN SAME DATA SCATTER BAND

TABLE 1-3

YIELD DIFFERENCES BETWEEN CLPP/ECLP AND RCLU
ARE SMALL BUT SIGNIFICANT

<u>Species</u>	Typical RCLU Data Scatter (1 Standard Deviation)	Average Difference Compared With RCLU	
		<u>CLPP</u>	<u>ECLP</u>
H ₂ Consumption	0.5	-0.2	-0.4
H ₂ O + Chemical Gases	1.1	-0.2	-1.2
C ₁ -C ₃	1.8	-1.7	-1.1
C ₄ -400°F	2.9	+2.2	-1.7
400-1000°F	3.8	+2.3	+2.1
Conversion	3.1	+2.8	-1.5
C ₄ -1000°F	2.7	+4.5	+0.4

Notes:

- (1) Values in boxes are statistically significant (less than 5% chance that the yield difference is zero)
- (2) All values are in lb/100 lb DAF coal
- (3) Results are based on 241 RCLU, 61 CLPP, and 12 ECLP bottoms recycle yield periods corrected for reactor holdup and backmixing differences processing three coals: Monterey, Wyodak, and Texas lignite.
- (4) Where applicable, ECLP Wyodak coal yield periods were corrected for gas-swept mill coal drying conversion debits

For more information, see the October 1-December 31, 1982 Quarterly Technical Progress Report, FE-2893-103.

Naphtha Hydrotreating - In RCLU, all liquefaction naphtha is hydrotreated along with the recycle solvent in the solvent hydrogenation unit. Since neither CLPP nor ECLP hydrotreat liquefaction naphtha, they should have slightly lower H_2 consumption, gas yields, and chemical gas yields, as well as higher naphtha yields. This effect is corrected for in study designs and economic comparisons, but not in the pilot unit yield comparisons listed in Table 1-3. Figures 1-6, 1-7, and 1-8 show the effect of naphtha hydrotreating on the naphtha heteroatom content for a range of coal feeds.

Temperature Profile - The liquefaction reactor in RCLU is isothermal, whereas both CLPP and ECLP have temperature profiles. The effect this has on yields is uncertain, however, model predictions indicate that increasing temperature profiles tend to decrease gas yields and increase liquid yields relative to isothermal operation. This, along with the lack of naphtha hydrotreating, tends to explain the lower gas yields at CLPP and ECLP. See the July 1, 1981-June 30, 1982 Annual Technical Progress Report, FE-2893-93, for more information.

Bottoms Distillation - Each pilot unit uses a different bottoms fractionation method. RCLU uses a hydrogen stripper, CLPP uses a vacuum flash, and ECLP has a vacuum tower. Consequently, the pilot units generally have different bottoms cut points. Since the $400^{\circ}F$ - $1000^{\circ}F$ yield includes the $1000^{\circ}F$ minus liquids contained in the bottoms, consistent measurement of the bottoms $1000^{\circ}F$ minus content is needed for accurate yield determination.

The historical basis for calculating yields in all three pilot units is by Microlube distillation, although $1000^{\circ}F$ minus content is often measured using a modified Hivac C distillation as well. Each of these distillation techniques is an in-house modification to the ASTM D-1160 standard procedure. (The Hivac distillation is used as the basis for determining the 15/5 "true boiling point" curve used in study designs). A comparison of these two methods is given in Figure 1-9, which plots Hivac versus Microlube $1000^{\circ}F$ content for Illinois coal (Monterey mine) bottoms recycle operations at RCLU, CLPP, and ECLP. These data show that for RCLU and ECLP, Hivac C and Microlube distillations are linearly related. However, this relationship is different for each of the three pilot units. RCLU has the highest Hivac $1000^{\circ}F$ minus content for a given Microlube, while ECLP has the lowest, and CLPP is in between. Data from Wyoming coal and Texas lignite bottoms recycle operations follow these same relationships.

In general using a Hivac basis instead of Microlube increases RCLU's $400^{\circ}F$ - $1000^{\circ}F$ yield and $1000^{\circ}F$ minus conversion more than the other pilot units. Thus, by using a Hivac basis, RCLU's apparently lower $400^{\circ}F$ - $1000^{\circ}F$ yield (seen in Table 1-3) can be reconciled with CLPP and ECLP. This difference is also important when estimating commercial yields, since the

FIGURE 1-6

FIGURE 1-7

FIGURE 1-8

PILOT UNIT DISTILLATION COMPARISON

ILLINOIS COAL BOTTOMS RECYCLE OPERATIONS

liquids recoverable in commercial vacuum towers are calculated based on Hivac distillations.

The cause for this difference between Hivac and Microlube in the three pilot units is being investigated. Preliminary results indicate that it correlates with the bottoms 1000°F plus compositions (i.e., asphaltene and preasphaltene contents) for all three pilot units. Further work is planned to clarify the mechanism.

Revised Yield Comparisons Still Show Some Pilot Unit Differences

Based on the above discussion, the pilot unit yield differences listed in Table 1-3 were corrected to a Hivac distillation basis, and for the effects of naphtha hydrotreating at RCLU. These revised pilot unit yield differences are given in Table 1-4.

The revised comparisons almost completely reconcile RCLU and CLPP yields, with all yield differences being less than 1 lb/100 lb DAF coal. However, ECLP's conversion and C₄-1000°F liquids yield remain about 5 lb/100 lb DAF below RCLU/CLPP.

Possible Causes for Remaining Yield Differences Under Investigation

A number of possible causes for ECLP's apparent lower conversion are being investigated. These causes are in two general categories: measurement differences and process differences.

Differences in the measurement of yields in the three pilot units can affect yield comparisons, as illustrated by the different results obtained using a Microlube or a Hivac distillation basis. Similarly, differences in the measurement of operating conditions affect yield comparisons, since all comparisons are adjusted to common operating conditions using the yield models.

In order to evaluate if ECLP's apparent lower conversion is due to measurement differences, several areas are being investigated. These include the Hivac/Microlube distillation difference, other possible differences in yield work-up, and uncertainties in determining ECLP's operating conditions.

The remaining yield differences may also be due to process/operating differences between the pilot units. Some possible differences which are being investigated include: regressive reactions in ECLP's separation system and the effect of ECLP's larger mean coal particle size.

TABLE 1-4

CORRECTING TO HIVAC BASIS RECONCILES CLPP
WITH RCLU, BUT SOME DIFFERENCES REMAIN

<u>Species</u>	<u>Microtube Based Model⁽¹⁾</u>		<u>Hivac Based Model⁽²⁾</u>	
	<u>CLPP</u>	<u>ECLP</u>	<u>CLPP</u>	<u>ECLP</u>
H ₂ Consumption	-0.2	-0.4	0.0	-0.2
H ₂ O + Chemical Gases	-0.2	-1.2	+0.8	-0.4
C ₁ -C ₃	-1.7	-1.1	-0.3	+0.3
C ₄ -400°F	+2.2	-1.7	+0.4	-3.4
400-1000°F	+2.3	+2.1	0.0	-1.7
Conversion	+2.8	-1.5	+0.9	-5.0
C ₄ -1000°F	+4.5	+0.4	+0.4	-5.1

Notes:

- (1) Same as in Table 1-1.
- (2) Corrected for RCLU naphtha hydrotreating
- (3) Values in boxes are statistically significant (less than 5% chance that the difference is zero)
- (4) All values are in lb/100 lb DAF coal
- (5) Results are based on 61 CLPP and 12 ECLP bottoms recycle yield periods corrected for reactor holdup and backmixing differences processing three coals: Monterey, Wyodak, and Texas lignite.
- (6) Where applicable, ECLP Wyodak coal yield periods were corrected for gas-swept mill coal drying conversion debits

1.4 Furnace Feed System Design

A study was initiated to define the EDS slurry preheater feed system design required for safe operation of a multipass furnace. It is known that safe furnace operation requires accurate pass-to-pass flow balancing, however a tolerable level of slurry feed maldistribution has not been quantified. The objectives of this study include definition of flowmeter accuracy necessary for safe multipass furnace design (limited by furnace coking considerations), and also a comparison of the relative cost and feasibility of alternative feed system designs. The study is based on recent learnings from the slurry manifold technology study and previously developed coking correlations (see July 1 - September 30, 1983 Quarterly Technical Progress Report, FE-2893-122).

The study approach has been to first determine a base coking rate for Wyoming bottoms recycle operation. Sensitivities to reduced flow rate, increased solids concentration, solvent quality and furnace operating conditions were investigated. Preliminary results, based on coking severity, indicated that maintaining a high mass flux will extend furnace runlengths, and coal concentration increases will not significantly increase coking rate. Maintaining solvent quality was also shown to be important. General information on slurry flow measurement is being assessed to determine future flowmeter testing requirements.

In the final portion of the study, three furnace feed systems will be evaluated based on the above information. These include a conventional manifold with spared feed pumps and individual pass slurry flowmeters and control valves, a symmetrical manifold with spared feed pumps and individual pass slurry flowmeters and control valves, and an individual positive displacement pump per pass system (with pump spares). With suitable flowmeter accuracy, either of the first two systems could be selected over the more complex individual pump per pass system. Results of the study will be documented in an interim report.

1.5 Slurry Flow Meter Test Program

A test program designed to determine the performance characteristics of a wedge type slurry flow meter and a venturi was carried out at Forney Engineering, Dallas, Texas between February 20-24, 1984. Equipment used in the tests consisted of a mixing tank with stirred mixer, a Viking positive displacement pump, and a flow loop which contained the meters. Coal in fuel oil slurry of 20, 25, 30, 35, 40, 45 and 50% wt were tested. Coal mixtures were prepared by weighing incremental amounts of coal and adding these to a fixed amount of fuel oil after fuel oil gravity was measured. No. 2 fuel oil was selected as a solvent because its characteristics closely match those of a typical EDS solvent being fed to a process furnace at elevated temperature. A technical objective of the test was to simulate the flow regime that would exist in a commercial unit.

These two flow meters were to be compared with a positive displacement flow meter. However, this objective was not achieved due to problems with the positive displacement meter. During the initial test phase, when a 20% weight slurry was run in the loop, coal particles dropped out of the liquid hydrocarbon as slurry passed through the positive displacement meter. This occurred after half an hour of operation at flow rates of 80 gallons per minute. Disassembly of the meter showed that internal parts of the meter were damaged beyond repair. As a result, a spool piece was placed in the line where the meter had been located, and the tests were run without a proving meter. The objective in doing this, knowing that a precise determination of wedge and venturi meter accuracy could not be made, was to test the directional capability of the meters. That is, to determine that the flow meters would indicate increased flow, when flow was increased by manipulating a pump by-pass valve.

Coal concentrations, on a weight basis, ranged from 20 on up to 50%, to determine if the directional properties of the meter were correct at varying concentrations. In all cases, both meters were found to be directionally correct. That is, as flow was increased by direct manipulation of positive displacement pump and by-pass of the flow loop, the meters always correctly indicated this. A 5 ft section of pipe was also used to produce P's that were measured at the same time as the wedge meter and venturi elements. The intent was to provide some method of checking pipe flow rate (as indicated by the pressure drop) and comparing it to meter performance. However, pressures measured were too low. During each slurry concentration run, a sample of the slurry was taken to permit later confirmation of slurry concentration.

Further action on this issue is being evaluated as part of the Furnace Feed System Design Study.

EDS TOXICOLOGY PROGRAM

1. Sample Acquisition and Storage

Nothing to report for this time period.

2. Analytical Characterization and Bioassay

2.1 Evaluation of the Subchronic Toxicity of EDS Hydrotreated Naphtha (ETP-20).

SUMMARY

An inhalation study was conducted to assess the subchronic toxicity of EDS hydrotreated naphtha (ETP-20). Information on the source⁽¹⁾ of the sample and on analytical characterization^(2,3) can be found in previous EDS Technical Reports. Male and female Sprague-Dawley rats were exposed to graded doses of hydrotreated naphtha vapor 6 hours/day, 5 days/week for 13 weeks. Following a two-week recovery period, the animals were sacrificed and necropsied. Parameters examined included survival, growth and clinical observations during the exposure and recovery periods, urinalysis at intervals throughout the study, blood chemistry at necropsy, and microscopic examination of selected tissues.

There was some evidence of systemic involvement following repeated inhalation exposure to EDS hydrotreated naphtha, although effects were mild and primarily confined to the high exposure groups. One manifestation of systemic toxicity was renal involvement in the male rat as evidenced by increased renal epithelial cell count and decreased urine osmolarity. These findings are consistent with results of inhalation studies conducted with low boiling petroleum and shale derived materials and suggest that light hydrocarbon nephropathy in male rats is a generic effect of inhalation exposure to these materials regardless of source. Other organ systems in which changes were observed included the lungs (both males and females), liver (males only), and adrenals (males only). The clinical significance of these observations is unknown.

CONCLUSIONS

There was some evidence of systemic involvement following repeated inhalation exposure to EDS hydrotreated naphtha, although effects were mild and primarily confined to the high exposure groups.

The only apparent treatment-related effect which was observed during the exposure period was some mild eye irritation in the high exposure group. Eye irritation was also noted in acute studies⁽⁴⁾ with this material. Ocular irritation has been reported in humans exposed to vapor concentrations of gasoline of 270-900 ppm and greater⁽⁵⁾, and the present results are consistent with the observation that vapors of low boiling hydrocarbons are slightly irritating to the eye.

Male rats exposed to the same dose regimen for 9 days demonstrated significantly ($p < 0.01$) increased liver weights as compared to controls, whereas in the present study, liver weights were only significantly different from controls when compared on a percent of body weight basis. Liver weights in females were not increased in either study. These results may reflect some compensatory liver growth in the males to cope with an increased metabolic load⁽⁶⁾ or may be simply a statistical anomaly.

Kidney weights in the male rats were elevated in the high dose group as compared to controls. There was also evidence of a dose-response relationship. In the 9-day inhalation study, a similar pattern of kidney enlargement was noted; however, statistical significance was not obtained. No renal effects were noted in the exposed female rats in either study. These results may reflect renal involvement associated with the light hydrocarbon nephropathy phenomenon in male rats^(7,8). The urinalysis findings of increased renal epithelial cell count and decreased urine osmolarity are consistent with this hypothesis. The increase in renal epithelial cell counts in the urine is considered to be a sensitive indicator of renal involvement but is not considered to be indicative of the level of damage sustained⁽⁹⁾. The absence of clear histopathologic findings suggests that the level of renal involvement must be low.

Subacute alveolitis of the lungs was observed in exposed animals of both sexes. The incidence exceeded that in the concurrent control groups. The severity of the change was primarily minimal or mild. It could not be definitively established from these data whether those changes resulted from inhalation exposure to HN or arose incidentally. The finding of increased adrenal weights in the present study is also of uncertain significance.

The absence of gross or histopathologic findings in the sexual organs, coupled with other findings from a reproductive toxicity study⁽¹⁰⁾, suggests that inhalation exposure to EDS hydrotreated naphtha has little effect on the reproductive process in rats.

A series of investigations of the subchronic toxicity of petroleum hydrocarbons are available for comparative purposes. These investigations were all conducted under protocols similar to that of the present study; rats (and other species) were exposed by inhalation 6 hr/day, 5 days/week for 13 weeks to graded doses of test materials. Repeated exposures to low boiling aromatic lean materials including Stoddard Solvent⁽¹¹⁾, unleaded gasoline⁽¹²⁾ and varnish makers and painters naphtha⁽¹³⁾ resulted in slight pathologic changes in the kidneys of male rats. Pathologic changes were not induced in the kidneys of female rats or of other species. Animals exposed to aromatic materials including toluene⁽¹⁴⁾ and mixed xylenes⁽¹⁵⁾ were not statistically different from air-exposed control groups by any of the criteria of injury which were monitored. These results suggest that repeated inhalation exposures to vapors of low boiling hydrocarbon liquids are essentially devoid of subchronic toxic effects other than the finding of light hydrocarbon nephropathy which is associated with exposure to aromatic-lean materials.

EXPERIMENTAL RESULTS

In-Life Observations

There were no spontaneous or treatment-related mortalities during this study; however, one female rat from the low exposure group was euthanized because of a broken leg. In-life observations during the exposure period were unremarkable, other than some evidence of eye irritation in the high dose group. Body weight gain in treatment groups was not significantly different (by Bartlett's Test of equal variances) from controls (Figures 2-1 and 2-2).

Gross Observations

Gross examination indicated little evidence of systemic toxicity. There was some evidence of kidney, and adrenal enlargement in the males (Table 2-1). Kidney weights in the high dose groups were significantly different from controls at the $p < 0.05$ level (Bartlett's Test of equal variances), and showed an increasing trend with dose. Adrenal weights were also significantly ($p < 0.01$) elevated in the high dose group males as compared to the controls. Liver weights were statistically different from controls but only when expressed as organ to body weight ratios (data not shown). No significant differences in organ weights were observed in female rats in any of the exposure groups.

HISTOPATHOLOGY

Microscopic examination identified some abnormal findings in sections of lung and kidney. Seven high dose group males had multifocal subacute alveolitis of the lungs as compared to three control males. Similar findings were noted in four group IV females, but not in any group I females. The microscopic findings in the lungs consisted of foci of mononuclear cells (lymphocytes, plasma cells, and macrophages) in several alveoli surrounding the terminal airways, generally near the pleura of the lungs. The severity of the changes was mostly minimal or mild.

The most common kidney changes included regenerative tubules and tubular protein, the former consisted of small areas of slightly hypertrophic, basophilic tubular epithelial cells in the cortex; tubular protein was characterized by segments of dilated tubules filled with pink staining proteinaceous material. The two changes were often found in the same section of kidney. Of the twelve examples of each in group IV males, eight animals presented with both. Six group I males had regenerative tubules; five of these also had tubular protein. Another four had tubular protein only. Among the group IV females, eleven had tubular protein and four had regenerative tubules; three had both. There were six group I females with tubular protein and one also had regenerative tubules.

Urine Chemistry

All of the investigated parameters were moderately affected by treatment. Urine volume was increased in the high dose group; glucose excretion was increased in the high and middose groups; protein and creatinine excretion were elevated in the high dose group, and urine osmolality was

FIGURE 2-1

MEAN FEMALE BODY WEIGHTS (G)
DURING HN EXPOSURE

FIGURE 2-2

MEAN MALE BODY WEIGHTS (G)
DURING HN EXPOSURE

Table 2-1
Mean Organ Weights (in grams)

Test Group	Sex	Liver	Kidney	Adrenals	Testes	Epi- didimides	Seminal Vesicles	Lungs and Trachea
Control	M	17.2 ^(a) +2.0	3.38 +0.3	0.048 +0.007	3.76 +0.21	1.46 +0.11	2.328 +0.414	1.92 +0.19
Control	F	9.5 +1.1	1.96 +0.21	0.074 +0.01	--	--	--	1.49 +0.10
0.2 mg/l	M	17.0 +1.9	3.39 +0.34	0.048 +0.007	3.77 +0.38	1.41 +0.13	2.419 +0.559	1.98 +0.14
0.2 mg/l	F	9.4 +1.0	2.00 +0.17	0.078 +0.007	--	--	--	1.51 +0.12
1.0 mg/l	M	17.4 +2.2	3.47 +0.29	0.051 +0.007	3.92 +0.38	1.43 +0.09	2.345 +0.564	2.02 +0.14
1.0 mg/l	F	9.3 +0.8	1.94 +0.15	0.079 +0.010	--	--	--	1.52 +0.11
5.0 mg/l	M	18.0 +1.9	3.68* +0.32	0.054* +0.009	3.92 +0.22	1.45 +0.07	2.470 +0.505	1.96 +0.3
5.0 mg/l	F	9.6 +1.1	2.00 +0.18	0.077 +0.008	--	--	--	1.50 +0.08

(a) N = 18 in all cases.

* $p \leq 0.05$ by Bartlett's test of equal variances.

reduced in the high dose group. In all cases, the largest effect was seen at the end of the fourth week. No significant differences among groups were observed at the end of the thirteenth week.

The parameter most affected was renal epithelial cell count (Table 2-2). The effects appeared to be both time and dose dependent. The peak effect was at the end of the fifth exposure week, and by the end of the second post-exposure week renal epithelial cell counts in the exposed groups were close to control values.

HEMATOLOGY

The following hematology measurements were performed on all animals at terminal sacrifice: white blood cell count, red blood cell count, hemoglobin concentrations and hematocrit. No significant differences were observed for any of these parameters (data not shown).

TEST SYSTEM

A total of 170 male and female Sprague-Dawley rats (85/sex), approximately six weeks old, were obtained from Hilltop Laboratory Animals, Scottdale, PA. The animals were randomly sorted, by weight, into four groups (18/sex). The animals were housed in stainless steel cages under controlled atmospheric conditions (20-24°C, 40-70% relative humidity) with a 12-hour light-dark cycle. Rats were housed two per cage during exposures and individually caged during the post-exposure periods. Animals were exposed 6 hours/day, 5 days/week for a total of 13 weeks; test material concentrations were 0 (room air control), 0.2 mg/l, 1.0 mg/l and 5.0 mg/l.

Four times during the study (at the end of first, fourth and thirteenth week of exposure and two weeks post-exposure), 10 male rats from each exposure group were recaged in Nalgene metabolism cages for overnight urine collection. The specific rats studied were randomly selected at the first collection interval and continued to be used throughout the experiment. Over a four-day period during the third post-exposure week, all animals were sacrificed under pentobarbital anesthesia following blood collection from the abdominal aorta. The animals were necropsied, and major organs were grossly examined. Selected tissues were collected and preserved in 10% neutral buffered formalin. Tissues from the high dose group and the controls were sectioned and submitted as hematoxylin and eosin stained slides for histopathologic examination. Tissues examined microscopically included: brain, heart, liver, lungs, trachea, nasal cavity, nasopharynx, kidneys, lymph nodes (cervical, pulmonary, mesenteric), testes, epididimides, prostate, seminal vesicles, ovaries, oviducts, uterus, cervix and vagina.

VAPOR GENERATION SYSTEM

The vapor generation system used in this study is shown in Figure 2-3. Vapors were prepared by heating the test material to approximately 360°F. As test material evaporated, it was mixed with room air and introduced into one cubic meter chambers. The chamber exhaust flow rates were maintained at 200 liters per minute (LPM) providing one air change every 5 minutes and an equilibration time (T₉₉) of 23 minutes.

Table 2-2

Effect of Inhalation of EDS Hydrotreated Naphtha on Urinary Excretion of Renal Epithelial Cells in Male Sprague-Dawley Rats

Dose Group	Exposure Day			
	5	26	89	103
0 mg/l				
Mean	44 (a)	153	1,006	603
S.D.	92	222	2,814	1,080
(N)	10	10	10	10
0.2 mg/l				
Mean	256	8,379	2,544	1,194
S.D.	345	4,988	2,117	1,119
(N)	10	10	10	10
39 1.0 mg/l				
Mean	1775* (b)	24,563**	6,908*	955
S.D.	1260	7,827	5,002	1,948
(N)	10	10	10	10
5.0 mg/l				
Mean	6426**	39,499**	10,403**	2,153
S.D.	2712	19,761	5,491	3,681
(N)	10	10	10	10

(a) Cells/hour
 (b) * - $p < 0.05$
 ** $p < 0.01$

Figure 2-3

VAPOR GENERATION AND EXPOSURE SYSTEM

Exposure concentrations were determined on both a nominal and actual basis. Nominal concentrations were calculated by dividing the net change in material weight over the exposure period by the total volume of air passing through each chamber. Actual concentrations were determined three times daily by drawing calibrated volumes of chamber air through charcoal tubes to adsorb test material. Hydrocarbons were eluted from the tubes with carbon disulfide and subjected to gas chromatography (GC). The total hydrocarbon content of each sample was determined by reference to test material standards. Normally the GC was used as a total hydrocarbon analyzer; however, at various times during the study, profiles of the eluted hydrocarbons were produced and compared to test material standards. No changes in marker peaks were observed indicating that the vapor was qualitatively similar in composition to the starting material. Results of chamber analysis and other chamber information are summarized in Table 2-3.

Additional analytical support was provided by continuous monitoring with a Hewlett-Packard 5880A GC modified to serve as a total hydrocarbon analyzer. During the first exposure week, a respirable aerosol mass monitor (Piezobalance Model 3500) was used to determine whether any aerosol was present in the exposure chambers. Total material present as aerosol made up less than 1% of the hydrocarbons in the exposure chamber.

Table 2-3

Summary of Vapor Generation and Concentration Data
Results are Listed as Daily Mean Values

	<u>GROUP I</u> <u>(Control)</u>	<u>GROUP II</u> <u>(0.2 mg/l)</u>	<u>GROUP III</u> <u>(1.0 mg/l)</u>	<u>GROUP IV</u> <u>(5.0 mg/l)</u>
Static Pressure (mm H ₂ O)	-10.0	-10.0	-10.0	-10.0
Average Chamber Airflow (LPM)	200	200	200	200
Exposure Duration (Minutes)	360	360	360	360
Total Flow (L) 72,000	72,000	72,000	72,000	
Nominal Concentration (mg/l)	0	0.26 ± 0.01	1.29 ± 0.04	6.51 ± 0.17
Actual Concentration (mg/l)	*	0.20 ± 0.03	1.08 ± 0.10	5.19 ± 0.42
42 Chamber Aerosol Concentration (mg/m ³)	0.04 ± 0.02	0.05 ± 0.01	0.04 ± 0.02	0.34 ± 0.03

* Below analytical limit of detection (<0.01 mg/l).

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LIST OF ABBREVIATIONS

ACGIH	American Conference of Governmental Industrial Hygienists
AID	Analytical and Information Division
API	American Petroleum Institute
APS	Atmospheric Pipe Still
ART	Actual (Slurry) Residence Time
ASME	American Society of Mechanical Engineering
ASTM	American Society for Testing and Materials
BACT	Best available control technology
BARD	Baytown Research and Development Division (Baytown, TX)
B/CD	Barrels per Calendar Day
BFW	Boiler Feed Water
BR	Bottoms recycle
B/SD	Barrels Per Stream Day
BTX	Benzene Toluene Xylenes
CCLU	Continuous Coking Liquefaction Unit
CD	Calendar day
C-E	Combustion Engineering, Inc.
CLPP	Coal Liquefaction Pilot Plant (One Ton Per Day Unit)
Co/Mo	Cobalt/Molybdenum
COD	Chemical Oxygen Demand
COT	Coil Outlet Temperature
CS	Chao-Seder
CSCU	Continuous Stirred Coking Unit
DAF	Dry, Ash-Free
DBM	Design Basis Memorandum

LIST OF ABBREVIATIONS (Continued)

DC	Direct Combustion
DCF	Discounted Cash Flow
DEA	Diethanolamine
DMEG	Discharge Multimedia Environmental Goal
DMMF	Dry Mineral Matter-Free
DOE	Department of Energy
DRS	Data Reconciliation System
DSC	Differential Scanning Calorimetry
ECD	Environmental Control Division
ECLP	EDS Coal Liquefaction Pilot Plant (250 Tons Per Day)
EDS-FPU	EDS - FLEXICOKING Prototype-Unit
EEPM	Exxon Engineering Project Management
EETD	Exxon Engineering Technology Department (Florham Park, NJ)
EOR	End-of-Run
EP	Extraction Procedure
EPA	Environmental Protection Agency
EPRI	Electric Power Research Institute
ERDL	Exxon Research and Development Laboratories (Baton Rouge, LA)
ER&E	Exxon Research and Engineering Company
ESP	Electrostatic Precipitator
FBG	Fluid Bed Gasifier
FBU	Fluidized Bed Unit
FCBU	Fluid Coker Bogging Unit
FCCU	Fluid Catalytic Cracking Unit

LIST OF ABBREVIATIONS (Continued)

FGDS	Flue Gas Desulfurization
FOEB	Fuel Oil Equivalent Barrels
FPTF	Fireside Performance Test Facility
FSBF	Full-Scale Burner Facility
FTIR	Fourier Transform Infra Red
GC	Gas Chromatography
GC/FID	Gas Chromatograph/Flame Ionization Detection
GPM	Gallons Per Minute
G.S.D.	Geometric Standard Deviation
GSM	Gas Swept Mill
HB	Hybrid Boiler
HBG	High Btu Gas
HC	Hydrocarbon
HCO	Hydrogenated Cresote Oil
HDN	Hydrodenitrogenation
HDS	Hydrodesulfurization
HPLC/FSUV	High Performance Liquid Chromatograph/Fast Scan Ultraviolet
HVGO	Heavy Vacuum Gas Oil
IBG	Intermediate Btu Gas
ICPES	Inductivity Coupled Plasma Emission Spectroscopy
ID	Inside diameter
IOL	Imperial Oil Limited
IKG	Integrated Coking/Gasification Pilot Plant
k	Thousand

LIST OF ABBREVIATIONS (Continued)

LBG	Low Btu Gas
LDU	Liquefaction/Distillation Unit
LGF	Load Growth Factors
LHV	Lower Heating Value
LP	Linear Program
LPG	Liquefied Petroleum Gas
LSCU	Large Stirred Coker Unit
LSFO	Low Sulfur Fuel Oil
M	Million
MATE	Minimum Acute Toxicity of Effluent
MBG	Moving-Bed Coal Gasification
M.M.D.	Mass Median Diameter
MPSS	Multi-Pass Spent Solvent
NDE	Non-Destructive Evaluation
NDIR	Non-Dispersive Infrared
NDT	Non-Destructive Testing
NESHAPS	National Emissions Standards for Hazardous Air Pollutants
Ni/Mo	Nickel Molybdenum
NIOSH	National Institute of Occupational Safety and Health
N-PNA	Nitrogen (containing) Polynuclear Aromatic
NPSH	Net Positive Suction Head
NRT	Nominal Residence Time
NSPS	New Source Performance Standards
OTCLU	Once-Through Coal Liquefaction Unit

LIST OF ABBREVIATIONS (Continued)

PACT	Powered Activated Carbon Treatment
PAM	Process Alternatives LP Model
PCTM	Pollution Control Technical Manual
PDA	Process Development Allowance
P&ID	Piping and Instrument Diagram
PNA	Polynuclear Aromatics
POX	Partial Oxidation
PR	Pressure Relief
PRD	Product Research Division (Linden, NJ)
PSD	Prevention of Significant Deterioration
psig	pounds per square inch, gauge
RCLU-1	Recycle Coal Liquefaction Unit with Nominal Coal Feed Rates of 75 Pounds-Per-Day
RCLU-2	Recycle Coal Liquefaction Unit with Nominal Coal Feed Rates of 75 Pounds-Per-Day
RCO	Raw Creosote Oil
RCRA	Resource Conservation and Recovery Act
RISP	Required Initial Selling Price
RKJZ	Redlich-Kwong-Joffe-Zudkevitch
RV	Reference Value - Code to prevent disclosure of non-program data.
SASS	Source Assessment Sampling System
S/C	Solvent/Coal
S/C/B	Solvent/Coal/Bottoms
SCF	Standard Cubic Feet

LIST OF ABBREVIATIONS (Continued)

SD	Stream Day
SDU	Study Design Update
SEM	Scanning Electron Microscope
SFEU	Solid Fuels Evaluation Unit
SFU	Small Fluidized Bed Unit (for coke gasification)
SHU	Solvent Hydrogenation Unit
SOR	Start-of-Run
SQI	Solvent Quality Index
SRC	Solvent refined coal
TACB	Texas Air Control Board
TDS	Total Dissolved Solids
T/D	Tons Per Day
TEC	Total Erected Cost
TGA	Thermal Gravimetric Analysis
TGP	Texaco Gasification Process
TIR	Temperature Increase Required
TMA	Thermomechanical Analysis
TOC	Total Organic Carbon
TRI	Thermal Resistance Increase
T/SD	Tons Per Stream Day
TSP	Total Suspended Particulate
UOP, Inc	Union Oil Products, Inc.
VB	Vacuum Bottoms
VGO	Vacuum Gas Oil
VLE	Vapor Liquid Equilibrium

LIST OF ABBREVIATIONS (Continued)

VPS	Vacuum Pipe Still
V/Hr/V	Volumetric Hourly Space Velocity, ft^3 feed per hour per ft^3 of reactor volume.
WBRSD	Wyoming Bottoms Recycle Study Design
W-N	Watson-Nelson
WWT	Wastewater Treatment
YP	Yield Period