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TECHNOLOGY ASSESSMENT GUIDE

NO. 9b

EXXON DONOR SOLVENT DIRECT LIQUEFACTION

CHAPTER ONE: EXECUTIVE SUMMARY

1.1   OVERALL PROSPECTS FOR THE TECHNOLOGY

The concept of using a hydrogen-donor solvent for coal liquefaction has been under development by Exxon Research and Engineering Company for 15 years. Essential to the process is the use of an externally hydrogenated recycle solvent which acts as a hydrogen donor in the tubular flow reactor. This approach is distinct from most other direct liquefaction processes which utilize a solvent oil without significant hydrogen donor properties. The EDS system does not employ an added catalyst in the process scheme.

The process offers high liquid yields (up to 2.6 bbl per ton of dry feed for C_4-1000°F liquids), and considerable flexibility to accept a wide range of coal feedstocks or produce a variable product state. These attributes are important in circumstances where coal price or availability changes or where shifts in product demand are likely.

Solidification of heavy materials and solids plugging have been the major operating problems. Overall, erosion and corrosion have not been extensive, but were significant where it occurred. Each of these problems can have impacts on process financial success, and it is assumed that a commercial EDS would not experience these difficulties beyond a manageable point.

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Although technically the EDS process appears to offer some attractive advantages, the projected product cost (including) is nearly $10/10^6$ Btu ($55/barrel), which is somewhat out of range of current imported oil prices, and is marginally higher than other competing direct liquefaction processes.

1.2   ENGINEERING ASPECTS[1-1]

The EDS process contains many process elements typical of liquefaction plants such as slurry preparation, hydrogen injection, reaction and product separation, but each step makes its own distinct impact on process operation and is therefore worthy of discussion.

The slurry preparation step combines slurry mixing and coal drying in one operation. Drying is achieved when the crushed feed coal is contacted with hot recycle donor-solvent, raising its temperature sufficiently to vaporize virtually all surface and in-situ water. This step allows some pre-reaction between the coal and donor solvent, and effectively prevents any deactivation of the coal with respect to its liquefaction reactivity which is sometimes caused by air drying, especially in the case of high moisture (low-rank) coals.

Following hydrogen injection, the slurry enters the tubular upward flow liquefaction reactor, operating at 800-900°F and 2000-2500 psi. The reactor contains no catalyst or internals, and simply allows sufficient residence time for reaction. Reactor effluent is separated in conventional separation and fractionation steps to produce light hydrocarbon gases, a C_3-1000°F distillate and vacuum bottoms containing residual material (>1000°F liquids), unconverted coal and mineral matter.

The process is designed to be self-sufficient in donor solvent, and a portion of the C_3-1000°F liquids are separated for this purpose. This material is deficient in hydrogen at this point in the process, and is hydrotreated in a conventional fixed bed hydrotreating unit with commercially available hydrotreating catalysts.

Exxon has discovered that recycling the liquefaction bottoms stream results in higher overall process yields, primarily by allowing heavy coal molecules a longer reaction residence time. This process option is therefore most likely to be part of any commercial plant based on this technology. Light hydrocarbon gases are steam reformed and subsequently shift converted to produce the hydrogen used in hydrotreating the recycle donor solvent. Vacuum bottoms are treated in Exxon's "Flexicoking" unit, a system originally conceived for coking and gasifying heavy petroleum residuum. The Flexicoker employs low-pressure circulating fluidized beds, operating at <50 psi and 800-1200°F in the coker and 1500-1800°F in the gasifier to convert essentially all of its feed carbon to either liquid products or gas. Gases produced are suitable for use as a fuel or as a source of hydrogen. Only a small amount of carbon escapes unreacted in the mineral matter cash stream.

Several alternate process schemes can be considered for the EDS process. Partial oxidation (gasification) of vacuum bottoms can produce a medium-Btu gas suitable for use as fuel or for hydrogen generation. This approach frees the light hydrocarbon gas stream to be sold. Hydrogen needs can also be met by direct coal gasification, as is practiced in some other processes. This alternative can divert vacuum bottoms for steam generation and process heat.

Flexibility in product state depends on process operating conditions and configuration. The single most important factor in determining liquid yield and quality is the extent of bottoms recycle. For example, in the original configuration (once-through liquefaction), bottoms Flexicoking and light gas reforming), naphtha and 400-850°F distillate yields were each 35 percent, with 25 percent heavy fuel oil. A change to bottms recycle, recycle Flexicoking and coal gasification for hydrogen generation resulted in a naphtha yield of 46 percent (although naphtha plus middle distillate was unchanged), no fuel oil and 21 percent C_1/C_2 gas. Higher naphtha yields are possible with bottoms recycle at very high conversions.

Bottoms recycle is effective in increasing the total liquid yield and the amount of lighter products. The increase in lighter products is approximately equal to the liquid yield from the Flexicoking unit. These higher liquid yields increase the number of options for assuring solvent balance (the plant must be self-sufficient in donor solvent production) and reducing the amount of excess liquefaction bottoms. Bottoms recycle has also been shown effective in improving pilot unit service factor. This is primarily due to the reduction in bottoms viscosity which accompanies bottoms recycle operation.

1.3   CURRENT COSTS

The total capital requirement for this 125 trillion-Btu per year liquefaction plant is $3.5 billion, which is dominated by a total plant investment of $2.2 billion and interest during construction of $1 billion.

Annual operating and maintenance costs (at a 90 percent plant capacity factor) total $177 million. Operating and maintenance materials consume the largest single share of this budget (37.2%) followed by local taxes and insurance (31.1%) and labor (26.5%). By-product credits given for sulfur and ammonia total $15.4 million, offsetting the total O&M costs to an annual net of $161 million.

Taken together with a 20 percent capital charge, these operating costs result in a product cost of $7.59/$10^6$ Btu, which is exclusive of coal costs. Assuming coal at $1.50/$10^6$ Btu, the total product cost is $9.96/$10^6$ Btu, or approximately $55/barrel.

1.4 RESEARCH AND DEVELOPMENT DIRECTIONS[1-1]

To achieve Exxon's goal of commercial readiness, the EDS program integrates all phases of process development. Bench scale research, small pilot unit operation and engineering design and technology studies support operation of a 250 TPD coal liquefaction pilot plant and a 70 TPD Flexicoking prototype program. Work is also in progress to evaluate the use of either a bottoms gasification process for generation of hydrogen or fuel gas, or direct combustion of bottoms for fuel.

Design data from the large pilot plants will be collected through extensive tests. The liquefaction plant was planned to include facilities for 115 individual tests which are expected to provide a comprehensive basis for a commercial plant design. Parameters to be tested include materials evaluation, erosion and corrosion, coal slurry preheat furnaces, slurry pumping, high pressure letdown valves and environmental monitoring and control.

A similar program is being planned for the flexi-coking program. Critical areas of investigation will focus on fluid bed operation, product quality and environmental control. Also included will be materials evaluation, high pressure vacuum bottoms pumping and coke particle integrity and gasification activity.

References

1-1. Epperly, W.R., D.T. Wade, and K.W. Plumlee. "Donor Solvent Coal Liquefaction," Chemical Engineering Progress, May 1981.

CHAPTER TWO: ENGINEERING SPECIFICATIONS

2.1   GENERAL DESCRIPTION OF THE TECHNOLOGY

In the Exxon Donor Solvent (EDS) process, coal is
liquefied in a pressurized vessel in the presence of hydrogen
and a hydrogen donor solvent. While the coal liquefaction
itself is non-catalytic, spent donor solvent is catalytically
re-hydrogenated in a separate vessel. Liquefaction products
are separated to recover LPG, naphtha, and low-sulfur fuel
oil products. Liquefaction bottoms consisting primarily
of unreacted coal and mineral matter are processed further
in a flexicoker unit, which generates plant products and
fuel gas for internal use. Waste streams from the
liquefaction and flexicoker units are treated together
to produce by-product ammonia, phenols, and sulfur.

The EDS process has been under development since
1966 by the Exxon Research and Engineering Co. (ER&E).
The most recent milestone in the development of the
process was the successful startup in 1980 of a 250-TPD
plant in Baytown, Texas.

2.2   PROCESS FLOW, ENERGY AND MATERIAL BALANCES

Plant area numbers which are relevant to the EDS process are listed in Table 2-1. The interaction among these units is shown by the conceptualized process flow diagram of Figure 2-1. In Table 2-2, mass flow rates for the major material streams given in the flow diagram are presented. Processing of these streams throughout the plant is described in detail below.

Cleaned coal from a nearby mine is received and stored (if necessary) prior to entering impact-mill coal crushers. The coal is reduced in size to 95 percent minus 8 mesh and fed via enclosed belt conveyors to slurry driers, where it is mixed with hydrogen donor recycle solvent.

Dried coal slurry (4% moisture by weight) is heated and mixed with hydrogen treat gas and is fed to liquefaction reactors, which operate at 840°F and 2000 psig. Coal is liquefied in these non-catalytic, tubular flow reactors in the presence of H_2 and the hydrogen donor solvent.

The liquid/solids liquefaction product stream is distilled into a number of cuts, including naphtha, low-sulfur fuel oil, spent solvent, and a vacuum bottoms slurry. Spent solvent is hydrotreated in a fixed-bed catalytic reactor to restore the donatable hydrogen to the recycle solvent, which is sent to the slurry driers. The vacuum bottoms slurry

Table 2-1

Relevant EDS Process Plant Area Numbers

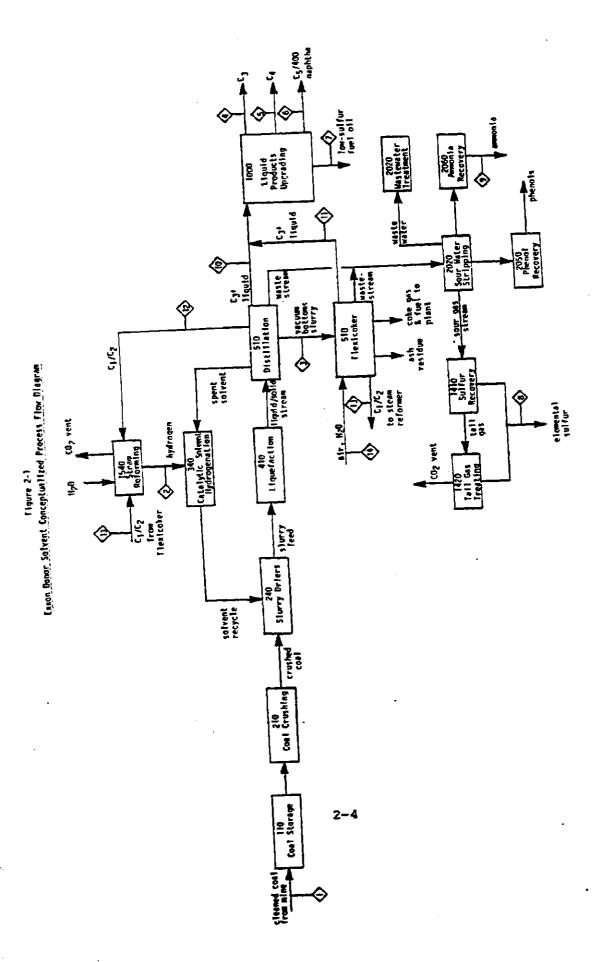|	
100	COAL STORAGE AND HANDLING
	110 Coal Storage
200	COAL PREPARATION
	210 Crushing and Grinding
	240 Drying
300	GASIFICATION
	340 Catalyst Recovery
400	HYDROGENATION
	410 Liquefaction Reaction
500	PRODUCT SEPARATION AND PROCESSING
	510 Fractionation (Distillation and Flexicoking)
1000	LIQUID PRODUCTS UPGRADING
1400	SULFUR RECOVERY AND TAIL GAS TREATING
	1410 Sulfur Recovery
	1420 Tail Gas Treating
1500	HYDROGEN PLANT
	1540 Steam Reforming
2000	UTILITIES AND SUPPORT SYSTEMS
	2020 Sour Water Stripping and Wastewater Treatment
	2050 Aqueous Phenol Recovery
	2060 Aqueous Ammonia Recovery

Figure 2-1

Exxon Donor Solvent Conceptualized Process Flow Diagram

cleaned coal from mine — ◇1 — 110 Coal Storage — crushed coal — 210 Coal Crushing — slurry feed — 240 Slurry Driers — solvent recycle

H₂O — CO₂ vent — 1540 Steam Reforming — hydrogen — 340 Catalytic Solvent Hydrogenation — spent solvent
◇11 — C₁/C₂ from flexicoker — ◇2
C₁/C₂ — ◇12

410 Liquefaction — liquid/solid stream

510 Distillation — vacuum bottoms slurry — waste stream
C₃⁺ liquid — ◇10 — C₃⁺ liquid — ◇11
◇3

510 Flexicoker — coke gas & fuel to plant — ash residue — waste stream
air, H₂O — ◇14 — C₁/C₂ to steam reformer — ◇13
sour gas stream

1000 Liquid Products Upgrading — ◇4 C₃, ◇5 C₄, ◇6 C₅/400 naphtha — ◇7 low-sulfur fuel oil

2020 Wastewater Treatment — waste water
2050 Ammonia Recovery — ammonia — ◇9
2020 Sour Water Stripping — 2050 Phenol Recovery — phenols

1810 Sulfur Recovery — tail gas — elemental sulfur — ◇8
1720 Tail Gas Treating — CO₂ vent

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Table 2-2

Major EDS Plant Material Streams

Stream No.	Composition	Mass Flow Rate (klb/hr)
1	Cleaned coal	2060
2	Hydrogen	90.64
3	Vacuum bottoms slurry	1067
4	C_3 product	31.3
5	C_4 product	34.0
6	C_5/400 naphtha	267.5
7	Low-sulfur fuel oil	471.1
8	Sulfur by-product	80.34
9	NH_3 by-product	12.4
10	C_3 + liquids	672.6
11	C_3 + liquids	290.5
12	C_1/C_2 gases from distillation	133.9
13	C_1/C_2 gases from flexicoker	37.08
14	Air/H_2O input to flexicoker	675.68

is conveyed to a flexicoker which generates C_3 and liquid product, C_1/C_2 gas products, and low-Btu fuel gas for plant consumption. The flexicoker employs an integrated coking/gasification sequence in circulating fluidized beds operating at low pressure (50 psi) and intermediate temperatures (900-1200°F in the coker and 1500-1800°F in the gasifier).

Light product gases (C_1/C_2) from the liquefaction products distillation unit and the flexicoker are used to produce hydrogen in a steam reforming unit. In this unit, the product gas is cryogenically purified and compressed for use in solvent hydrogenation. C_3 and liquid products from the liquefaction and flexicoking processes are delivered to a liquid products recovery unit where C_3/C_4 LPG, C_5-naphtha, and low-sulfur fuel oil are separated out for export.

Waste streams from the liquefaction and flexicoker units are conveyed to a sour water stripper which produces a wastewater stream for treatment and by-product streams for ammonia and sulfur recovery.

An overall EDS plant material and energy balance is shown in Table 2-3. A total coal input of approximately 24,720 tons per day (HHV = 12,663 Btu/lb) would be required to produce 125×10^{12} Btu/year of products.

Table 2-3

Overall Material and Energy Balance

Input	HHV Btu/lb	Mass Flow Rate klb/hr	Gross Heating Value MMBtu/hr
Coal	12,663	2,060	<u>26,086</u>
TOTAL ENERGY INPUT			26,086
Products			
C3 LPG	21,651	31.3	678
C4 LPG	21,263	34.0	723
Low-Sulfur Fuel Oil	16,930	471.1	7,976
Naphtha	18,342	267.5	<u>4,906</u>
PRODUCT ENERGY INPUT			14,283
Ammonia By-product	9,800	12.4	120
Sulfur By-product	4,000	80.3	<u>322</u>
TOTAL ENERGY OUTPUT			14,725
OVERALL PLANT EFFICIENCY			56%

2.3   PLANT SIZING AND SITING ISSUES AND CONSTRAINTS

The EDS plant examined in this study is designed to produce the equivalent of 125 x 10^{12} Btu/year of LPG, naphtha, and low-sulfur fuel oil products. The plant would need to be adjacent to a mine or in a location suitable for the receipt of the required amounts of coal feed — either by train, barge, or slurry pipeline. If the plant were to be located near a mine, the resources of the area would have to be capable of supplying approximately 9 million tons of cleaned coal per year for the estimated plant operating lifetime of 20 years. Sufficient water and land area (approximately 45 acres) requirements for such a plant would also be required.

2.4   RAW MATERIAL AND SUPPORT SYSTEM REQUIREMENTS

The coal assumed for use in this plant design is
Illinois No. 6 Coal, with a HHV (dry) of 12,663 Btu/lb, as-
received moisture of 15 percent, and an elemental analysis
as follows:[2-1]

Constituent	Wt % (dry basis)
C	70.15
H	5.07
O	10.24
N	1.19
S	4.14
Cl	0.11
Ash	9.10
	100.00

Approximately nine million tons per year of dry, cleaned coal
of this type would be required for the EDS plant under con-
sideration.

Steam, process and cooling water, instrument and
process air, and fuel gas requirements estimated for a plant
of 125 x 10^{12} Btu/year product capacity are shown in Table
2-4. Catalyst and chemical requirements for such a plant are
listed in Table 2-5. Of course, water, air, and chemical
supply and distribution systems capable of handling these
volumes of material would be required.

Table 2-5

Catalyst and Chemical Requirements

Unit	Catalyst	Amount
Solvent Hydrogenation	Ni-Mo, 1/16" extrudate	n.a.
	Inert balls (mullite, 3/8" diameter)	1905 ft^3
	Coarse gravel	3090 ft^3
Liquefaction Bottoms Flexicoker	Coker gas hydrotreater catalyst	1308 ft^3
	Driers (alumina)	180 ft^3
	H_2S removal unit solution makeup	2885 gal/day
Hydrogen Generation Unit (units: ft^3)	Co-Moly	654
	Zinc Oxide	3492
	Nickel Urania	1772
	Iron Oxide	2340
	Copper Oxide	3883
	Solution	8601
	Inhibitor	2101
	Nickel Oxide	2560

Unit	Chemical	Amount
Gas and Water Treating	Caustic	27 gpm
	DEA	30900 initially, 62 gal/day makeup
	Isopropyl Ether	3914 lb/day
Product Recovery	Calcium Chloride	14 ft^3/day

Source: Reference 2-1

2.5   EFFECT OF COAL TYPE

The EDS process has been successfully employed to
liquefy bituminous, subbituminous, and lignitic coals at
pilot plant scale. Product yields for selected coals are
shown in Table 2-6. The relatively high oxygen content of
low-rank coals causes high H_2O and CO_2 yields, while these
low-sulfur coals produce less H_2S. Low-rank coals have been
more difficult to process than bituminous coals because of
problems with $CaCO_3$ deposits and high viscosity of lique-
faction bottoms. However, process modifications which lessen
the impacts of these operating problems have been demonstrated.
These include periodic withdrawal of agglomerates by acid
washing of wall scale and coal pretreatment with SO_2 to con-
vert calcium to the more stable calcium sulfate. High vis-
cosity bottoms can be controlled by increasing the liquefaction
residence time.

Table 2-6

Liquefaction Product Yields

Liquefaction Yields (Lbs/100 Lbs. DAF Coal)	Illinois No. 6 Bituminous (Monterey No. 1)	Wyoming Subbituminous (Wyodak)	Texas Lignite (Big Brown)
H_2	-4.3*	-4.6*	-3.9*
$H_2O + CO_x$	12.2	22.3	21.7
$H_2S + NH_3$	4.2	0.9	1.7
$C_1 - C_3$ Gas	7.3	9.3	9.1
$C_4 - 1000^oF$ Liquid	38.8	33.3	33.3
Bottoms	41.8	38.8	38.1

*Net Requirement

Source: Reference 2-2

2.6   AIR POLLUTION CONTROL TECHNOLOGY

Pollution control techniques and designs for the EDS system considered in this study were chosen to meet state and federal regulations, including NSPS.

The area of greatest potential air pollution problems in the EDS plant is the coal receipt, storage, and preparation area. Fugitive dust is to be controlled by water sprays, enclosed conveyor systems, and baghouse filters. Dry fines are emitted by the flexicoker also, but these particles are removed in a venturi scrubber prior to gas venting.

Conventional pollution control systems are employed to remove sulfur from gas streams. H_2S and CO_2 are stripped from waste streams and sulfur is recovered by a sulfur plant. Tail gas from the plant is treated to produce additional sulfur and to provide cleanup of vent gases.

2.7   WATER POLLUTION CONTROL TECHNOLOGY

Wastewater treatment system design for the plant under
consideration is based on "Best Available Control Technology"
(BACT) standards for the mid-1980's. Wastewaters produced
by the EDS system include sour and non-sour phenolic streams.
Sour water is stripped of H_2S and CO_2 and is combined with
a non-sour phenolic water stream to be sent to the phenol
extraction plant. NH_3 is recovered in anhydrous form. Crude
phenols are recovered and the remaining wastewater along with
H_2S removal unit purge is treated by dissolved air floatation,
biological oxidation, filtration, and activated carbon adsorption.
Approximately 15 percent of the treated wastewater is reused as
cooling tower makeup, with the remainder being discharged.

The H_2S removal unit purge stream may require additional
treatment to reduce chemical oxygen demand (COD) due to the
presence of thiosulfate. Acidification with sulfuric acid would
convert the sodium thiosulfate to sulfate and promote recovery
of vanadium and anthraquinone disulfuric acid (ADA), which are
also present in the waste stream.

2.8   SOLID WASTE DISPOSAL

Solid wastes produced from an EDS process include digested biological sludge, oily sludge, flexicoker ash, and solids from the boiler feedwater treatment unit. In addition, spent catalysts are disposed of intermittently.

Wastewater treatment sludges are thickened, concentrated, and loaded into trucks for transportation to a remote land farming operation. The sludge could be used in revegetation of mine tailings. Flexicoker ash is slurried with water and disposed of in a lagoon which is covered over and vegetated after several years. Spent catalysts may be disposed of either in a landfill or by regeneration, reuse, or metals reclamation.

2.9   OSHA ISSUES

The EDS process requires pulverized coal. The pulver-
ization process may expose workers to coal dust and noise. How-
ever, safe coal handling procedures are well understood.

Another issue will be that the products of the pro-
cesses, especially the fractions in the higher boiling
temperature ranges, are likely to be high in carcinogenic
polynuclear aromatic hydrocarbons (PAH). Exposure to these
fractions may occur during maintenance and cleaning operations,
or during inadvertent contact with the coal-oil slurry. Ex-
posure can be reduced through the use of protective clothing
and proper personal hygiene.

2.10   PROCESS PERFORMANCE FACTORS

2.10.1   Product Characteristics and Marketability

The EDS process produces C_3 and C_4 LPG, C_5 naphtha,
and low-sulfur fuel oil products as well as by-product ammonia
and sulfur. Selected properties of the principal products are
shown in Table 2-7.

While the C_3 and C_4 products would be marketed as
LPG products, the low-sulfur fuel oil and naphtha products
are potentially valuable blending stocks. However, refining
of the naphtha product into gasoline or heating oil would
require hydrotreating to reduce sulfur, nitrogen, and sulfur
levels.

2.10.2   Capacity Factors, Flexibility, Reliability

The EDS plant assessed in this study is designed to
operate with a 90 percent capacity factor (see Chapter 3)
and to produce synthetic fuel products at approximately 60
percent overall thermal efficiency.

The flexibility of the EDS process for operating
on various feedstocks is illustrated in Section 2.5 above.
Yield and operability data have been obtained over a wide
range of operating conditions.

Table 2-7

Selected Product Characteristics

- C_3 LPG

HHV:	21,651 Btu/lb (3.86 MBtu/bbl)
C_2- fraction:	0.7 wt %
C_4+ fraction:	4.3 wt %
Specific gravity:	0.509

- C_4 LPG

HHV:	21,263 Btu/lb (4.33 MBtu/bbl)
C_3- fraction:	2.4 wt %
C_5+ fraction:	1.7 wt %
Specific gravity:	0.582

- Low-sulfur fuel oil

HHV:	16,930 Btu/lb (5.97 MBtu/bbl)

Elemental Composition (wt %):

C	88.91
H	7.29
O	2.56
N	0.71
S	0.53

- Naphtha (C_5/400°F)

HHV:	18,342 Btu/lb (5.20 MBtu/bbl)

Elemental Composition (wt %):

C	87.59
H	10.23
O	1.91
N	0.13
S	0.14

2.11   TECHNOLOGY STATUS AND DEVELOPMENT POTENTIAL

The first phase of Exxon's EDS process research was begun in 1966 in a cost-sharing program with DOE, EPRI and other interests. Through predevelopment, planning and design, and process engineering, a one-ton per day pilot was was completed in 1975. Process performance has also been studied in a 5C lb/day recycle coal liquefaction unit (RCLU). A 250-TPD Exxon Coal Liquefaction Plant (ECLP), under construction since 1978, was successfully started up in the summer of 1980.

Detailed testing on the ELCP has begun on a wide variety of coals for a range of process conditions. Some uncertain technical issues relating to EDS technology which will be studied include slurry coking, ammonium chloride plugging, and slurry handling problems in the liquefaction process and the effects of high ash levels on the flexicoker operation.

2.12   REGIONAL FACTORS INFLUENCING ECONOMICS

A wide variety of regional constraints apply to the
design, construction, and operation of an EDS plant. Siting
and resource requirements for such a complex are described in
Sections 2.3 and 2.4 above. Environmental control constraints
would be determined according to the particular meteorology,
topography, and existing pollution regulations at a proposed
site. Pollutant releases can be either continuous or inter-
mittent, and hazardous materials may be produced in either
mode. The magnitude of release of such materials is de-
pendent on plant size and operating conditions. The types
of pollutants produced by the EDS process and methods for
their control and disposal are described in Sections 2.6-
2.8 above.

References

2-1.   Fant, B.T. EDS Coal Liquefaction Process Development,
       Phase IIIA, U.S. DOE Report FE-2353-13, January 1978.

2-2.   Mitchell, W.N., K.L. Trachte, and S.Z. Zaczepinski.
       Performance of Low-Rank Coals in the Exxon Donor Solvent
       Process.   Presented at the Lignite Symposium, Grand
       Forks, N.D., May 1979.

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SECTION THREE

ECONOMIC ANALYSIS - EXXON DONOR SOLVENT

3.1 Introduction and Methodology

3.1.1 Economic Anlysis Methodology

The economic analysis relies on a conceptual design for
a commercial size coal liquefaction plant using Exxon Donor
Solvent Technology (3-1). The data presented in the report
was scaled to a size of 125 trillion Btu per year and
corrected from 1975 dollars to third quarter 1980 dollars.
As local taxes and insurance were omitted in the reference
document this operating cost was estimated by ERCO at
2.5 percent of the total plant investment.

In order to make the plant self-sufficient in electricity,
ERCO added a cost estimate for a 218.1 MW coal-fired electric
power plant with flue gas desulfurization to the capital
cost estimate. The power plant was sized at 218.1 MW to
supply 180 MW at 90 percent capacity. According to the
reference used (3-2), these power plants have only 74.6
percent availability. The cost estimate in dollars per
kilowatt capacity in (3-2) was updated from 1978 to 1980
dollars when the cost estimate was made. The cost of the
electric power plant is shown as Unit 2050, Electric Power
Plant, in Table 3-1 below.

The adjusted data was then used to estimate product
costs.

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3.1.2 Scaling Exponents

The base plant size was 121.5 x 10^{12} Btu/year (3-1).
This was scaled up to 125 x 10^{12} Btu/year using a scaling
exponent of 1.0. The scaling exponent of 1.0 was based on
the strong assumption that all available scale economies
were captured in the reference plant.

3.1.3 Price Indices

Costs and by-product credits were updated from 1975 to
1980 dollars using the indices and prices discussed in the
Background Section.

3.1.4 Economic Criteria

The standard economic criteria discussed in the Back-
ground Section were used. The schedule of investment
expenditures was 15 percent, 45 percent, 30 percent and 10
percent in years one through four of construction.

3.1.5 Contingencies

A project contingency of 15 percent was applied to the
sum of all area costs to cover unanticipated cost increases
during final design and construction.

A process contingency of 50 percent was applied to the
cost of Area 400, Hydrogenation, because it was not yet at
the pilot plant stage when the report was prepared. Area

500, Product Separation and Processing, was assigned a
contingency of 25 percent to allow for further technical
development, and Area 1500, the Hydrogen Plant, was assigned
a 10 percent process contingency.

3.2 Capital Costs

3.2.1 Itemized Capital Costs

Itemized capital costs for the plant are shown in
Table 3-1. The Total Plant Investment is $2194.7 million.
The Hydrogen Plant, Area 1500, is the most expensive compo-
nent, at $405.4 million. Area 400, Hydrogenation, at $276.3
million, and Unit 2050, Electric Power Plant at $205.6 mil-
lion, are also expensive components.

The total capital requirement is $3466.4 million, as is
also shown in Table 3-1. Besides the total plant investment,
the largest component of the capital requirement is interest
during construction at $1006.1 million. Interest during
construction is large because of the long, four year,
construction period.

3.2.2 Variability of Capital Costs

For areas within the plant using advanced technology,
all equipment was specified, and heat and material balances
were complete. For the conventional technologies, the
capital cost estimate relied on duty specifications. This
level of design detail corresponds to between the "budget
authorization" and the "study estimate" level of accuracy

TABLE 3-1

TOTAL CAPITAL REQUIREMENT: EXXON DONOR SOLVENT[a]

AREA	UNIT	ITEM	COST (10^6 $)	PERCENT OF SUBTOTAL
100		Coal Handling and Storage	39.1	2.3
200		Coal Preparation	192.2	11.3
400		Hydrogenation	276.3	16.3
500		Product Separation and Processing	244.7	14.4
1500		Hydrogen Plant	405.4	23.8
2000		Utilities and Support Systems		
	2020	Wastewater Treating	52.5	3.1
	2030	Solids Disposal	24.1	1.4
2100	2050	Electric Power Plant	205.6	12.1
		Offsites and miscellaneous		
	2120	Tankage, Shipping and Receiving	49.6	2.9
	2130	Other Support Facilities	210.3	12.4
	Subtotal		1699.8	100.0
	Process Contingency		239.9	
	Project Contingency		255.0	
	Total Plant Investment		2194.7	
	Working Capital		133.9	
	Start-Up		131.7	
	Interest During Construction		1006.1	
	Total Capital Requirement		3466.4	

[a]Source: 3-1, updated to 1980 dollars and scaled by ERCO to 125 trillion Btu/yr.

as defined in the Chemical Engineering Handbook (3-3), which
would place the cost estimate within 20-30 percent.
Because the technology had no large pilot plant experience
when the estimate was made, and because very conservative
contingencies were added, the amount of variability in this
estimate is probably in the range of ±35 percent.

3.3 Operating and Maintenance Expenses

3.3.1 Itemized Operating and Maintenance Expenses

Gross operating and maintenance (O&M) expenses total
$176.8 million, as is shown in Table 3-2. These expenses
are dominated by Local Taxes and Insurance (estimated by
ERCO as described in Section 3.1 above) at $64.9 million,
and Materials and Other at $65.8 million. Labor was esti-
mated at $46.9 million.

Gross O&M expenses are partially offset by sulfur
and ammonia by-product credits. Sulfur sales would yield
approximately $9.5 million and ammonia sales $5.9 nmillion
for a total by-product credit of $15.4 million. As
Table 3-2 shows, net O&M costs (gross O&M expenses less
by-product credits) total $161.4 million.

3.3.2 Variability of Operating and Maintenance Costs

Local taxes and insurance, materials, and other expenses
were estimated as a portion of the total plant investment.
They account for 68.3 percent of gross O&M costs. Therefore,
it is reasonable to expect O&M costs to be as variable as
the capital cost estimate, or within ±35 percent.

TABLE 3-2

NET ANNUAL OPERATING AND MAINTENANCE COST:
EXXON DONOR SOLVENT[a]

ITEM	COST (10^6 $)	PERCENT OF SUBTOTAL
Local taxes and insurance	54.9	31.1
Labor	46.9	26.5
Materials and other	65.8	37.2
Catalysts and chemicals	8.7	4.9
Water	0.5	0.3
Total O&M costs	176.8	100.0
By-Products	10^6 $	
Sulfur	(9.5)	
Ammonia	(5.9)	
Total	(15.4)	
Net O&M Costs	10^6 $	
Gross O&M Costs	— 176.8	
By-Product Credits	(15.4)	
Total	161.4	

[a]Source: 3-1, updated to 1980 dollars and
scaled by ERCO to 125 trillion Btu/yr.

3.4 Effect of Technology Development on Costs

The cost estimate presented here is for a pioneer plant. Exxon estimated that a mature plant could be built for up to 40 percent less than this pioneer plant (3-1, p. 25) ten years after the pioneer plant is built, using the experience and technology improvements gained while operating the pioneer plant.

Another means to assess possible cost reduction through technology development is with the experience theory discussed in the Background section. The Hydrogenation area (400), Product Separation and Processing area (500) and Hydrogen Plant area (150) together account for 59.5 percent of the Total Plant Investment if process contingencies on these areas are included. These areas are all comprised of immature components which would be improved by construction of more Exxon Donor Solvent (EDS) plants. With a maximum experience factor of 10 percent on new energy technologies, the experience factor on EDS technology would be 59.4 percent of 10 percent or approximately 6 percent. Each doubling of EDS production capacity could result in a cost reduction of 6 percent in real dollars.

3.4 Product Costs

The coal liquids produced by the plant have three cost components: capital charges, net O&M costs, and fuel costs. A non-fuel product cost can be computed from the capital costs and the net O&M costs using the formula described in the Background section. This non-fuel cost indicates the cost of converting coal into a synthetic fuel independent of the price of coal. From Table 3-1 and 3-2, the

O&M cost is $161.4 million. With a capacity factor of 90 percent and a capacity of 125×10^{12} Btu/yr, the non-fuel product cost is:

$$P = \frac{(\$3,466.4 \times 10^6 \times 20\%) + \$161.4 \times 10^6}{125 \times 10^{12} \text{ Btu} \times 90\%}$$

$$= \begin{array}{cc} \$6.16 /10^6 \text{ Btu} & + \quad \$1.43/10^6 \text{ Btu} \\ \text{(capital costs)} & \text{(O\&M costs)} \end{array}$$

$$= \$7.59/10^6 \text{ Btu}$$
(Total non-fuel product cost)

The total non-fuel product cost is $7.59/10^6$ Btu, with capital costs of $6.16/million Btu, and O&M costs of $1.93/ 10^6$ Btu.

The non-fuel cost can be combined with a cost of coal to yield a total product price using the formula given in the Background. The overall coal to gas efficiency of the process is 63.3 percent, not including sulfur, or ammonia. With a coal cost of $1.50/10^6$ Btu, the product cost can be computed as follows:

$$E = \begin{array}{cc} \$7.59/10^6 \text{ Btu} & + \quad \dfrac{\$1.50/10^6 \text{ Btu}}{0.633 \text{ efficiency}} \\ \text{(capital and} & \text{(coal costs)} \\ \text{O\&M costs)} & \end{array}$$

$$= \begin{array}{cc} \$7.59/10^6 \text{ Btu} & + \quad \$2.37/10^6 \text{ Btu} \\ \text{(capital and} & \text{(coal costs)} \\ \text{O\&M costs)} & \end{array}$$

$$E = \$9.96/10^6 \text{ Btu}$$
(total product cost)

The total product cost would be $9.96/10^6$ Btu. This corresponds to an oil price of approximately $55/barrel.

REFERENCES

3-1. Exxon Research and Engineering Company, "EDS Coal Liquefaction Process Development Phase IIIA," U.S. Department of Energy, FE-2353-13, January 1978.

3-2. Electric Power Research Institute, "Technical Assessment Guide," Special Report, EPRI PS-1201-SR, July 1979.

3-3. Perry, Robert P. and Cecil E. Chilton, eds., Chemical Engineering Handbook, Fifth Edition (New York: McGraw-Hill, 1973), pp. 25-12 to 25-17.