Assessment of Fuels for Power Generation by Electric Utility Fuel Cells

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APPENDIX A

FUEL AVAILABILITY AND PRICE FORECAST

1.0 CRUDE PETROLEUM AND REFINED PRODUCTS

1.1 ENERGY AND PETROLEUM

In recent years the U.S. has been facing a radically changing energy environment. Traditionally, the nation has enjoyed a position of relative self-sufficiency in cheap energy, but declining domestic oil and gas reserves and quantum rises in energy prices on the international market have brought this to an end. From 1960 until the OPEC oil price increases in 1973, total U.S. primary energy demand increased at an average annual rate of 4.3% with oil supplying an average 43% of total energy requirements and gas fulfilling another 34%. This same period saw the declining position of coal (whose share of the primary energy demand slipped from 20% in 1960 to under 17% in 1973) and the emergence of nuclear power as an important new source of primary energy.

Since 1973, U.S. energy policy makers have been attempting to steer the country along a path of slower energy demand growth, greater reliance on the more abundant domestic energy sources, and maximum discovery and development of domestic resources (with little success to date). The goal of slower energy demand growth was temporarily realized in the first two years after the OPEC price rises as consumers cut back energy use in response to higher energy prices and to the economic recession. On the supply side, very little improvement has been registered as the prices producers were allowed to charge continued to be held down by either FEA price controls or FPC regulation. In 1976 U.S. energy demand resumed its upward trend, showing a 4.8% increase over 1975 consumption levels. Oil imports reached a high of 7.2 million barrels per day in 1976, reflecting the fact that imported oil was the only fuel available to fill the widening gap between domestic energy supply and demand.

Our forecast of energy demand assumes that U.S. oil prices will rise (in real terms) above current levels so that by 1985 domestic prices will have reached international parity. Similarly, gas prices are not projected to be immediately decontrolled, but will be gradually permitted to approach a premium value above their heat equivalency with oil prices. Government policies in conjunction with rising energy prices are assumed to have moderate success in encouraging conservation by such means as tax credits for energy-saving capital investment, efficiency standards for appliances, more stringent insulation requirements, etc. In the same way the government (aided by higher energy prices) will have modest success in encouraging the development of non-oil energy sources--coal, nuclear, solar, etc. It has been assumed that rising energy prices will be accompanied by a slightly slower than historic rate of economic growth and will average 2.7 to 3.5% per year over the period to 1990.

1.1.1 Energy Demand

Over the period 1960 to 1973, U.S. primary energy demand grew at an average annual rate of 4.3%. This trend was abruptly reversed in 1974 and 1975 in response to higher energy prices and the economic downturn. However, 1976 saw a growth in total energy demand of 4.8% over 1975, partially reflecting the abnormally low levels of 1975, but also signalling consumer acceptance of higher energy prices.

In our forecast, shown in Table A-1-1, future growth in energy demand is expected to be below historic levels and to become slower over the forecast period from the cumulative effect of conservation programs and higher energy prices. Thus, for the period between 1976 and 1980, total energy demand is expected to average 3.3% p.a., declining to 2.9% between 1980 and 1985, and 2.5% p.a. from 1985 to 1990.

Historically, the two fastest growing end-use sectors have been transportation and electric utilities. In the future, growth in transportation demand is forecast to decline dramatically, as mandated efficiency standards are applied to new automobiles and as growth in fuel demand for air travel declines. Utilities will continue to experience moderate growth in fuel demand, but the rate of growth is expected to decline over the forecast period from an average of 5.7% p.a. in the period up to 1980 to 4.3% p.a. in the 1980's. Industrial demand (excluding feedstocks) will increase 2.5% p.a. on the average through 1985 (with more rapid growth in the earlier years and tapering off in the 1980's), reflecting the effects

TABLE A-1-1

U.S. ENERGY DEMAND BY END-USE SECTOR

(Quadrillion Btu's)

	1976	1980	1985	1990
Transportation	19.3	19.6	19.8	20.0
Residential/Commercial	14.7	15.6	16.4	16.5
Industrial	14.5	16.5	18.4	20.4
Utility	22.2	27.7	35.0	42.4
Other	5.1	7.0	10.3	13.5
Total	75.8	86.4	99.9	112.8
Average Annual Demand Growth Rate (% p.a.)		+3.3	+2.9	+2.5

Source: A. D. Little Estimates

of more efficient energy use. The "other" category shown in Table A-1-1 is primarily comprised of petrochemical feedstocks, a market which will continue to grow rapidly.

1.1.2 Energy Supply

The U.S. has entered a period in which non-oil fuels are supply-constrained. These constraints occur in the form of basic resource limitations, production restrictions limits, transportation bottlenecks and environmental regulations. For indigenous natural gas the constraining factor is the basic resource itself; for coal the limitations are production, transportation and environmental regulations; hydroelectricity is resource-limited; and development of nuclear capacity is limited by utility capital availability, demand uncertainties and public opposition to nuclear power plants. In this situation, oil is left to play the balancing role, equating total energy demand with available fuel supplies. Thus, the oil demand shown in Table A-1-2 is derived by subtracting the supplies of non-oil fuels from total projected demand. The assumptions behind the individual supply forecasts are briefly described below.

1.1.2.1 Coal

Forecasted coal production figures are shown in Table A-1-2 in quadrillion Btu's. The coal supply/demand balance and assumptions used in developing these projections are discussed in Section 2.0 of Appendix A.

1.1.2.2 Nuclear

Despite the fact that the United States currently produces half of the total world nuclear electricity output, the development of nuclear capacity in the United States continues to encounter numerous obstacles. These obstacles include technological problems, environmental opposition, cost escalations, lengthening construction times, financing difficulties, and demand uncertainties.

The nuclear supply forecast shown in Table A-1-2 reflects the most recent appraisal of individual projects, as well as the assumption that

TABLE A-1-2

U.S. ENERGY SUPPLY AND DEMAND TO 1990

(Quadrillion Btu's)

	1976	1980	1985	1990
Oil	35.0	40.0	45.3	45.3
Gas	21.1	20.2	20.0	19.2
Coal	14.6	19.6	24.8	32.9
Nuclear	2.0	3.3	6.5	11.0
Hydro*	3.1	3.3	3.3	4.4
Total Energy Demand	75.8	86.4	99.9	112.8

Source: A. D. Little Estimates

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*Includes geothermal, solar and other unconventional energy sources.

plants will continue to experience delays in start-up and less than complete utilization. Our nuclear forecast is based on the addition of nearly 180 nuclear plants (of 1,000 mw each), so that by 1990 installed generating capacity will rise from 42 gw in 1976 to 220 gw in 1990.

1.1.2.3 Hydro/Other

Hydroelectric power generation has leveled out because the most appropriate dam locations have been utilized. Annual fluctuations in hydroelectric output will occur as a result of varying meteorological conditions, but no significant incremental generating capacity is anticipated over the forecast period.

Other energy forms, such as solar, geothermal, wind, etc., are expected to grow significantly over the next 15 years. Despite this expansion, the contribution of these novel energy forms will be small (about 0.1 quadrillion Btu by 1985 and 1 quadrillion Btu by 1990).

1.1.2.4 Petroleum

U.S. petroleum production peaked in 1972, and declined by 1.6 million bbl/d between 1972 and 1976. Crude oil production from existing reserves in the lower 48 states and southern Alaska is expected to decline by about 7.5% p.a. in the future--7.7 million bbl/d in 1976 to 3.5 million bbl/d in 1990 (Figure A-1-1). Associated with the drop in conventional lower 48 oil and gas production will be a fall in gas liquids output (lease condensate, LPG, ethane and natural gasoline) from 2.0 million bbl/d in 1976 to 1.2 million bbl/d in 1990. These declines will be offset by three new oil sources:

- Prudhoe Bay which is expected to operate at 1.7 million bbl/d capacity in 1980 through 1985, declining slightly by 1990.

- New oilfield discoveries, as well as revisions to existing fields through improved recovery and extensions which will contribute 0.3 million bbl/d to the 1980 oil supply and 3.9 million bbl/d to the 1990 supply. Much of this new oil supply could come from outer continental shelf (OCS) areas and the Naval Petroleum Reserves (NPR). NPR 4 in Alaska will contribute significantly to an expansion of North Slope production.

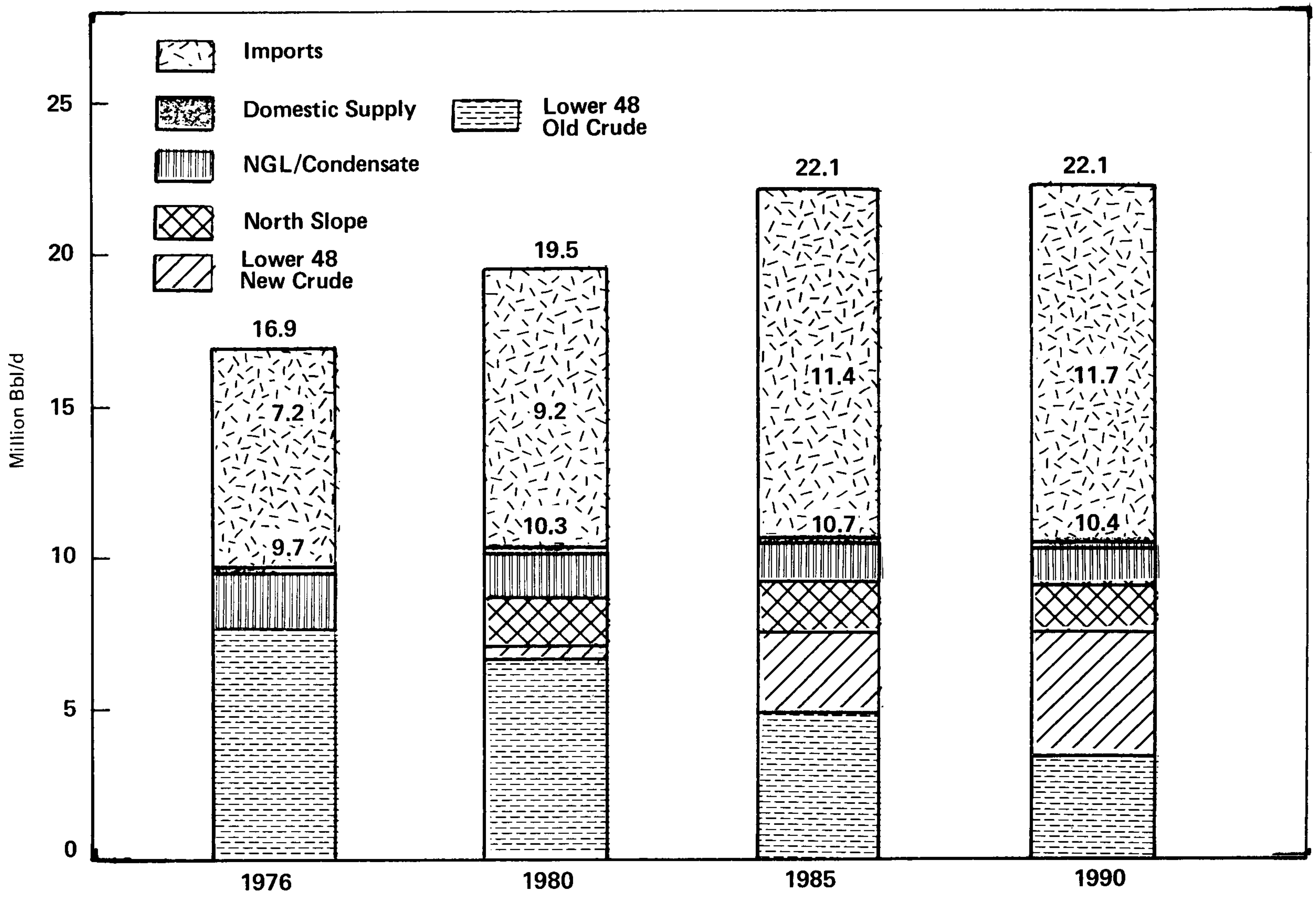

Figure A-1-1 U.S. OIL SUPPLY AND DEMAND 1976-1990

- Syncrude production from oil shales which have been included at 0.2 million bbl/d in 1990, and from coal liquefaction plants (assessed at 0.1 million bbl/d in 1990).

Projected U.S. oil supply and demand are shown in Table A-1-3.

TABLE A-1-3

U.S. OIL SUPPLY AND DEMAND

(Million Bbl/d)

	1976	1980	1985	1990
Lower 48/Southern Alaska - Old	7.7	6.7	4.9	3.5
New Discoveries/Revisions	-	0.3	2.6	3.9
Prudhoe Bay	-	1.7	1.7	1.5
NGL's	1.6	1.3	1.2	1.0
Lease Condensate	0.4	0.3	0.3	0.2
Shale Oil	-	-	-	0.2
Coal Liquids	-	-	-	0.1
Total Supply	9.7	10.3	10.7	10.4
Crude Oil Demand (excluding processing gain)	16.9	19.5	22.1	22.1
Oil Imports	7.2	9.2	11.4	11.7

Source: A. D. Little Estimates

1.2 PETROLEUM PRODUCTS PRICE FORECASTS

1.2.1 International Crude Oil Prices

Crude oil prices are today the single most important influence on the delivered price of oil products to the consumer. Figure A-1-2 depicts the growth in FOB prices for Arabian light crude oil (currently used by OPEC as a marker for pricing of all its crude oils) FOB Ras Tanura. The price has risen to the current level of $12.70/Bbl from $1.27/Bbl in 1970.

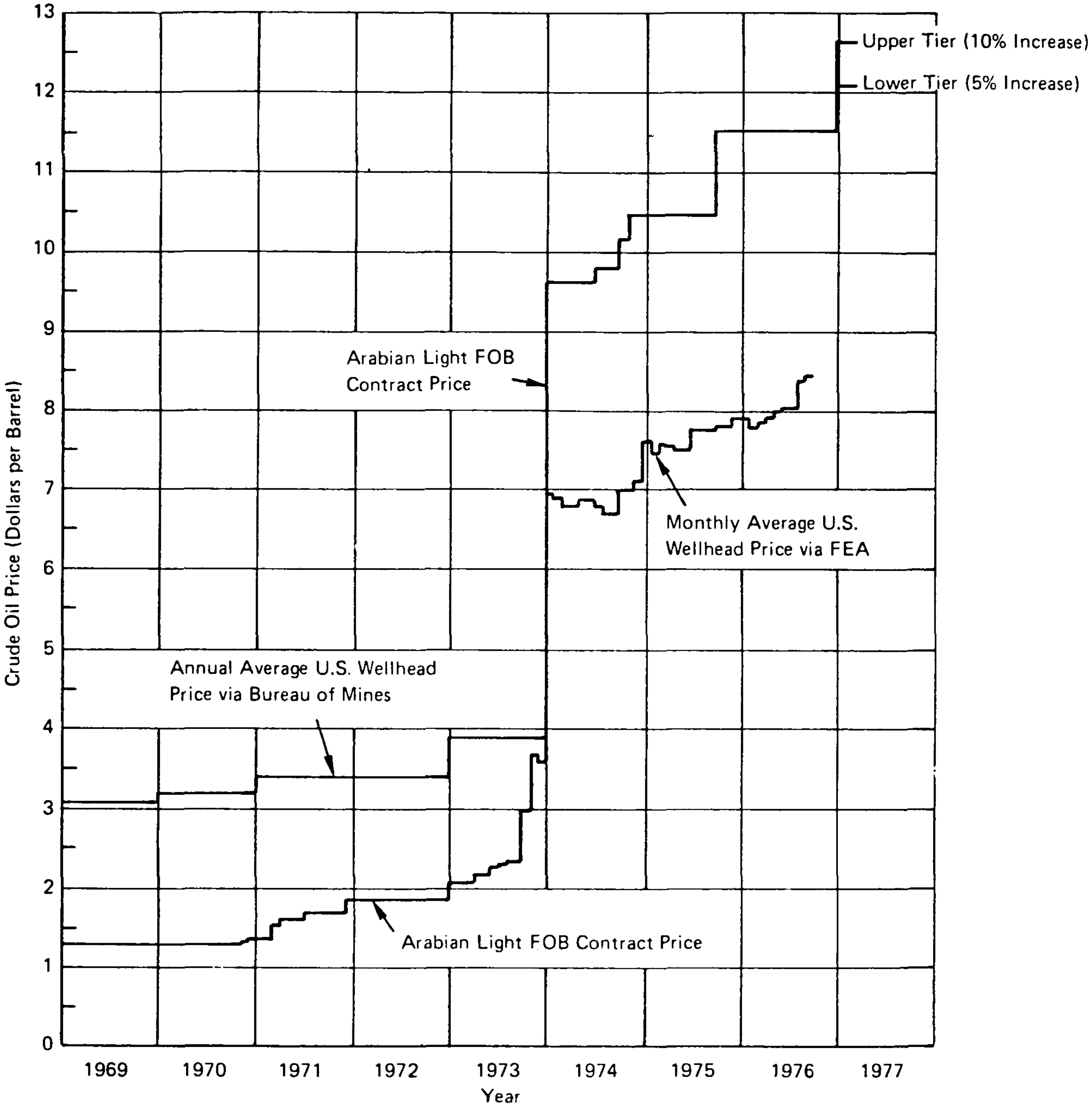

FIGURE A-1-2   EVOLUTION OF DOMESTIC·AND FOREIGN CRUDE OIL PRICES

During that time, international crude oil prices have overtaken U.S. domestic crude prices.

The quantum jump in crude prices which took place in 1973/74 was caused by a number of factors, including:

- Rapid growth in oil demand due to synchronization of the economies of the major developed countries.
- Temporary shortages of crude oil production capacity and shipping and refinery capacity.
- The Arab-Israeli war of June 1973, and the Arab oil embargo.

Since 1973/1974, OPEC has been a producing cartel and has succeeded in imposing further price increases, approximately in line with dollar cost inflation. Surplus tanker capacity now exists in the market today, and (as outlined in Section 1.2.2) is expected to continue through the 1980's. Spare refinery capacity is also available--particularly in Europe, less so in the United States and Japan. Oil demand, which grew at 7% p.a. in the 1960's and early 1970's, was temporarily halted by the economic recession of 1974-1976. However, oil demand is expected to resume growth (3.4% p.a. over the next 15 years) as the world accommodates the higher oil prices. Oil demand growth will be depressed below pre-1973 levels. This is the result of efforts to accelerate production of other energy forms such as coal, more efficient use of energy through conservation (particularly in the United States), and through price-induced demand elasticity.

Despite this outlook for lower oil demand growth, OPEC will be able to maintain its grip on world oil prices. Table A-1-4 illustrates A. D. Little's forecast of future world oil supply and demand. It shows that despite the maximization of new non-OPEC production sources (North Sea, Mexico, etc.), the demand for OPEC oil will continue to increase, assuming that oil prices remain at current levels in real terms. Despite the fall-off in its production between 1973 and 1975, OPEC was still able to maintain a cartel price (and increase it) although its spare production capacity almost tripled over that period.

TABLE A-1-4

WORLD OIL* SUPPLY AND DEMAND[†] 1960-1990

(Million Bbl/d)

	1960	1965	1970	1973	1975	1980	1985	1990
World Oil Demand:	22.0	31.7	47.8	57.8	56.0	69.0	81.2	91.3
Non-OPEC Production								
U.S.A.	8.0	9.0	11.3	11.0	10.0	10.3	10.7	10.4
Eastern Bloc	3.4	5.4	7.9	9.6	12.0	15.5	16.8	17.5
Others	1.9	2.8	4.6	5.5	6.5	11.9	14.2	20.7
Subtotal Non-OPEC	13.3	17.2	23.8	26.1	28.5	37.7	41.7	48.6
OPEC Capacity	n.a.	n.a.	n.a.	35.4	37.9	45.1	48.4	49.2
"Surplus"	n.a.	n.a.	n.a.	3.7	10.4	13.8	8.9	6.5
OPEC Production	8.7	14.5	24.0	31.7	27.5	31.3	39.5	42.7

Source: A. D. Little Estimates

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*Includes natural gas liquids and condensates.

[†] Excludes stock changes and unallocated demand.

OPEC oil production is not expected to grow significantly between now and 1980 from its current level of about 31 million bbl/d, due principally to the buildup of North Sea oil output. Although this will result in an increase in OPEC's spare productive capacity, this is not expected to create any problems in terms of cartel management for OPEC. Overall OPEC production (and hence revenues) will increase over 1975 levels. During this period, however, OPEC will be unable to gain significant real price increases, since its production levels will remain more or less constant, and ample spare production capacity will exist. Furthermore, the growth in oil demand to 1980 will still leave spare refining capacity and surplus tanker capacity. Thus, during this period the potential for demand-induced shortages creating a seller's market for oil is considered to be relatively limited.

After 1980, a new picture will emerge. Table A-1-4 shows that OPEC oil production must begin to increase rapidly in response to growing oil demand and, in addition, its spare capacity (even with projected capacity expansions) is expected to decline. This situation will intensify if non-OPEC oil production cannot grow as rapidly as forecast. Furthermore, it is expected that during the 1980's (earlier for refineries, later for tankers) there is a strong probability of a shortage-inspired oil products price increase created by the oil industry's short-term lack of refining or transportation capacity. As in 1973-1974, this will be permanently captured by OPEC in terms of a quantum increase in the FOB price of its crude oil. An in-house poll of energy experts at A. D. Little concerning the expected trends in international crude oil prices showed that the probability of a discontinuous price path is much higher than that for a smooth progression of significant real increases in the price. This is due to the political difficulties both within, and external to, OPEC which would occur if steady, real increases in the price of crude oil were instituted.

After this "hardware shortage" price increment (c. 1985), a continuation of relatively stable oil prices in real terms is projected as the world market adjusts once more to these higher prices. The impact of such a quantum increase (\$5/bbl) on oil demand is difficult to assess, but it is not expected to reduce oil demand growth by much more that 1% p.a. On the supply side, such a price increase is expected (on the basis of current cost projections) to come close to encouraging the development of new resources such as shale oil, tar sands and Middle East LNG (for the U.S. and European markets). The contribution from these new resources to world energy supply before the year 2000 is, however, expected to be relatively minor (no more than 3 million bbl/d of oil equivalent).

As oil demand growth continues into the 1990's at rates between 2 and 4% p.a., the probability of promoting premature resource exhaustion grows steadily. Figure A-1-3 shows the expected impact of different oil demand growth rates on the world oil supply curve. Conservatively, it is expected that continued oil demand growth at 2-4% p.a. will result in a peaking of world oil supply between 2000 and 2020. Prior to that time, however, new energy technologies must be developed (solar, breeder reactors) because oil cannot indefinitely be the world's marginal energy source. Until new technologies in the amounts necessary to fill the "oil supply gap" are developed, a "resource increment" is expected which will increase the oil price in the 1990's or shortly thereafter. This will further depress the demand for oil, thus extending the useful life of the remaining resources. It will also create a greater incentive for more rapid development of major new energy sources. The timing and magnitude of this increment are, of course, difficult to assess. By the mid-1990's however, it is anticipated that oil resources may be strained sufficiently to create the atmosphere for a further quantum jump in prices (again assessed at around $5/bbl in real terms).

An illustration of the projected price trend for Arabian light crude oil (in 1975 dollars) is shown in Figure A-1-4. The $5/bbl "hardware" and "resource" discontinuities in the price are shown as occurring in 1985 and 1995 respectively, although their precise timing and magnitude are uncertain.

The actual values used in the base international crude oil price projection was the arithmetic average of the prices defined in two scenarios tested in A. D. Little's in-house poll. Opinion was almost evenly divided on which scenario of price evolution was more probable. The prices were

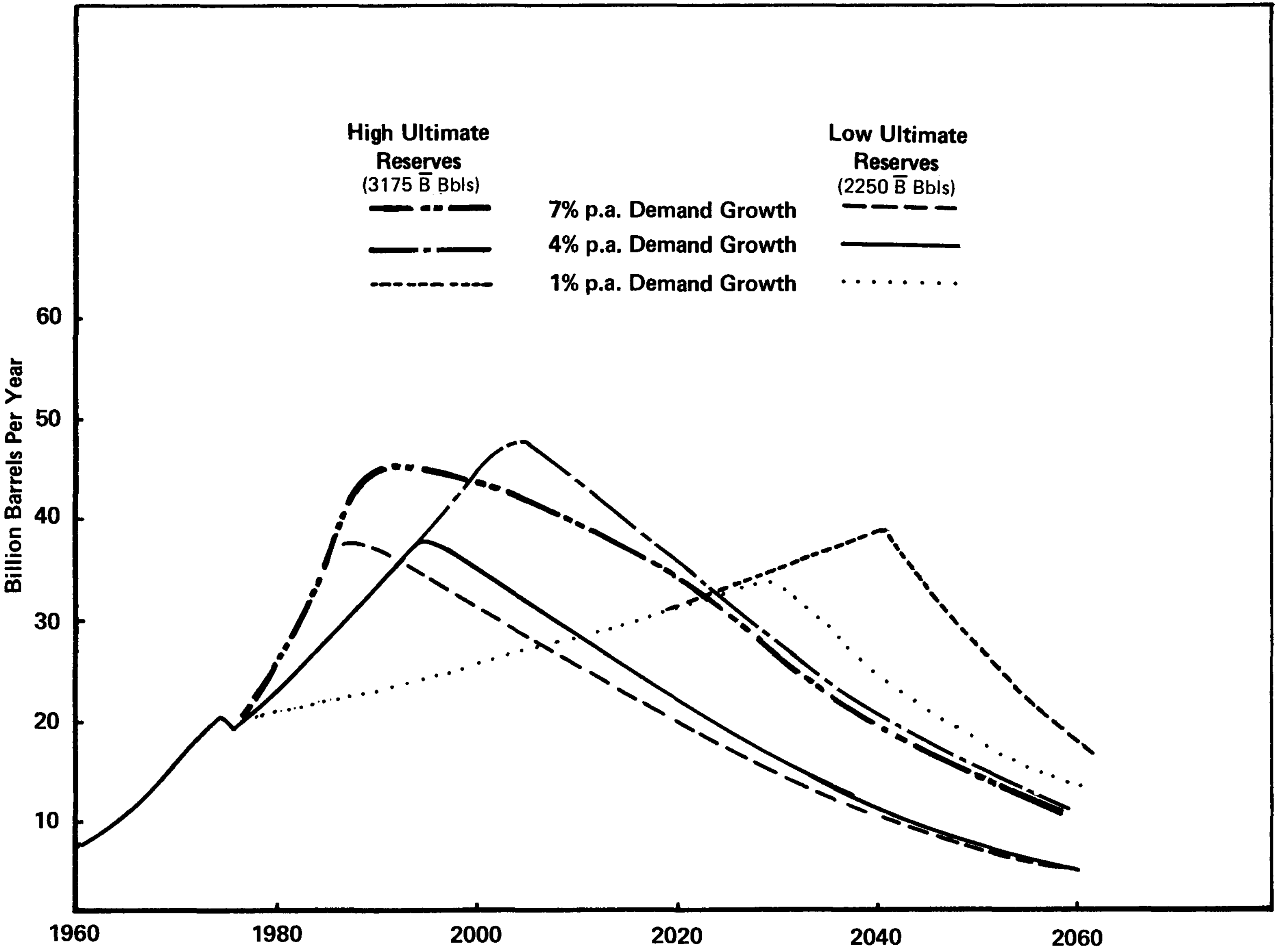

Figure A-1-3 WORLD CRUDE OIL PRODUCTION SCENARIOS 1960-2060

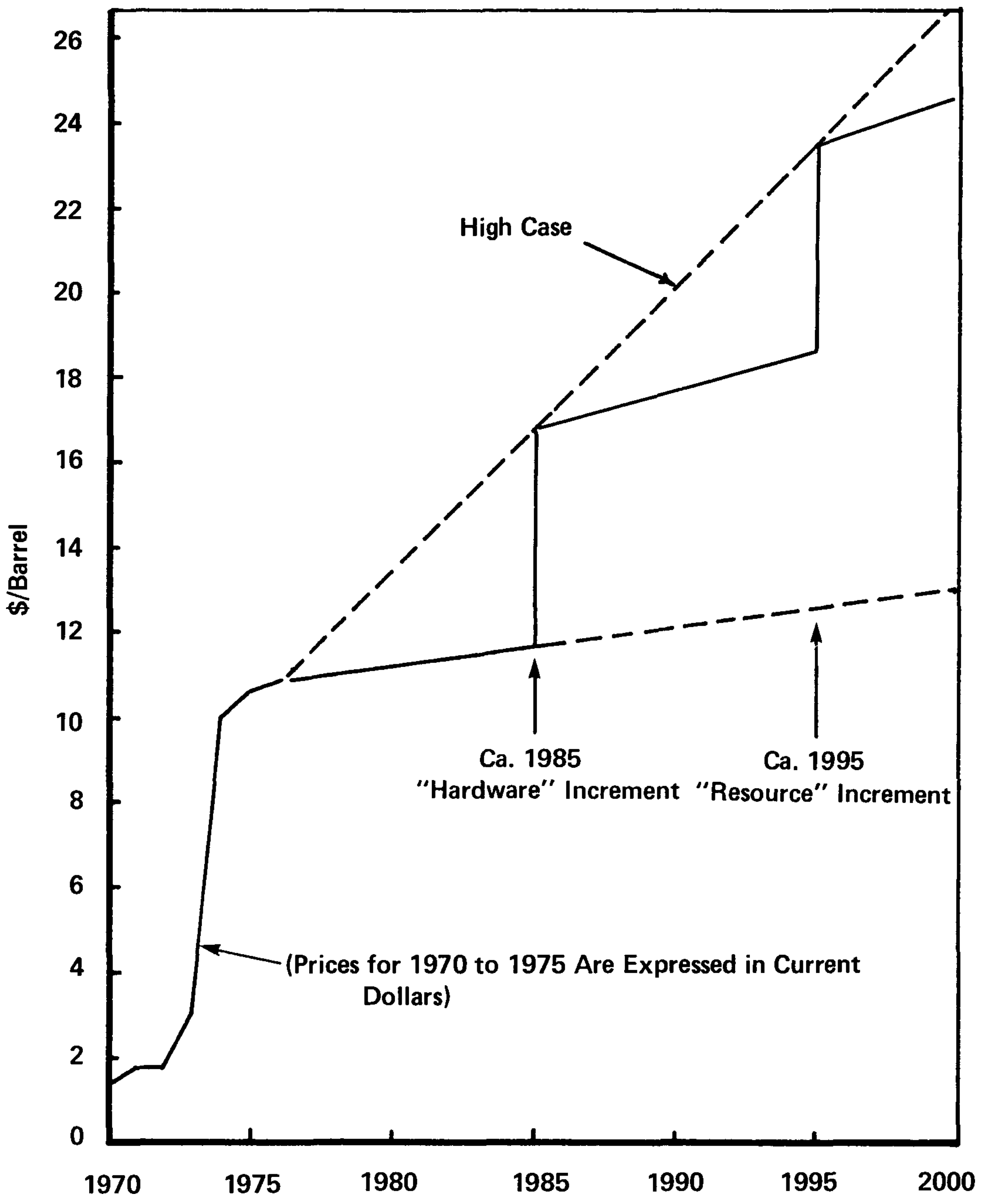

Figure A-1-4 INTERNATIONAL CRUDE OIL PRICE SCENARIOS 1975 $/Barrel FOB Ras Tanura for Arabian Light projected originally in 1976 dollars and for the purposes of this study were de-escalated to 1975 dollars using a GNP deflator of 1.06.* The price scenarios developed for the in-house analysis and the final FOB prices employed in this current study for EPRI are tabulated in Table A-1-5. Implicitly a future inflation rate of 5% p.a. (in U.S. dollars) was assumed. The price projections, if the price scenarios and inflation forecast were correct, would lead to a 1990 "best estimate" FOB crude price of around $37/bbl in current dollars.

A projection of future U.S. domestic crude oil prices was outside the scope of this current study for EPRI. It was mutually agreed that the domestic crude price projections and the domestic-international volume weightings required to calculate the average refinery gate crude price, under a scenario of continued controls on domestic prices, would be obtained from the Foster Associates recent study for EPRI. (8) The 1980 weightings of imported to domestic crude oil were obtained from A. D. Little's projections of domestic crude supply and imports, since the Foster Associates' study did not cover the year 1980.

1.2.2 Ocean Transportation Costs

International crude oil freight rates are primarily affected by the demand and supply balance of the world tanker fleet. Therefore, a forecast of the demand for and supply of large tankers (above 30,000 DWT) which are employed for long-haul crude oil movements, was developed. Next, the representative costs of transporting crude oil on major trade routes were determined. A review of these transportation costs, the present level of cost recovery through freight rates (spot and/or period cover), and the demand-supply situation in the present world fleet provided the guidelines which determined the future levels of cost recovery in the tanker market. These relative levels were determined for the major trade routes (which act as market-setting routes) and were then transformed into the freight rate forecast using the Worldscale index for the various trade

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*Bureau of Labor Statistics.

TABLE A-1-5

PROJECTED FUTURE INTERNATIONAL CRUDE OIL PRICES

34° API Arabian Light FOB Ras Tanura

	1976 $/Bbl			1975 $/Bbl
	SCENARIO I $5 Quantum Increases in 1985 and 1995	SCENARIO II 2.5% per year Increase plus $5 Quantum Increases in 1985 and 1995	Average	Average
1976	11.51	11.51	11.51	10.86
1980	11.51	12.70	12.10	11.42
1985	16.51	19.38	17.94	16.93
1990	16.51	21.28	18.90	17.83
1995	21.51	28.40	24.96	23.55
2000	21.51	30.82	26.17	24.69

routes as a point of reference for the actual costs of transportation for non-U.S. flag vessels.

1.2.2.1 Tanker Demand

A matrix of the major crude oil movements by maritime routes was developed by breaking down crude oil imports into the main geographic regions. Using OECD statistics, the imports were broken down by areas of production. It was assumed that the relative weights of the crude oil sources will remain about the same for each importing region. In the case of the United States, the impact of the Alaskan crude oil and declining Canadian imports was reflected in the relative importance of the future crude oil sources. The resulting matrix is presented in Table A-1-6.

Based on the operational characteristics of very large crude carriers (VLCC's), the tanker tonnage that would be required to transport a given volume of crude oil over a range of one-way voyage distances between the producing regions and the consuming areas was calculated. Due to the current world oversupply of tankers, the fleet has averaged service speeds well below average design speeds. The owners' desire to reduce bunker fuel costs is one of the main reasons behind this slow-steaming. Widespread slow-steaming has the effect of increasing the required ship tonnage to transport the same volume of crude. In-house forecasts of worldwide oil demand and supply over the period 1975-1990 were used to forecast the matrix of future crude oil movements. This matrix provided the volumes (in million bbl/d) of crude oil that each importing region required from the major producing areas. Using a specific tonnage requirement for each major trade route (DWT per bbl/d shipped) and the matrix of the crude oil movements, the tanker tonnage requirements of each trade route for each year under consideration were then calculated. The aggregate of the requirements represents the estimated demand for crude oil tanker tonnage. Tables A-1-7 to 10 show the projected crude oil volume and the tanker tonnage requirements for the years 1975, 1980, 1985, and 1990.

The present situation of oversupply of tanker tonnage, slow-steaming and other forms of sub-optimum utilization of tankers (longer port times,

TABLE A-1-6

CRUDE OIL IMPORTS IN 1975

(% of Total Imports of Importing Area)

FROM/TO:	U.S.	West Europe	Japan	Latin America	Asia/ Australia
Canada	(12)*	-	-	-	-
Venezuela	13	2	-	95	-
E. Europe	-	3	-	-	-
N. Africa	10	7	3	-	-
W. Africa	18	13	-	-	-
Middle East	34	75	78	5	90
Far East	8	-	19	-	10
Other	5	-	-	-	-
Total	100	100	100	100	100

Source: OECD, A. D. Little Estimates.

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*Canadian exports to the United States are almost exclusively by pipeline and do not influence tanker supply/demand.

TABLE A-1-7

CRUDE OIL IMPORTS BY TANKERS AND TANKER DEMAND IN 1975*

(Tanker Capacity in Million DWT)

FROM/TO:	U.S.		West Europe		Japan		Latin America		Other	
	$\overline{M}$Bbl/d	$\overline{M}$ DWT	$\overline{M}$Bbl/d	$\overline{M}$ DWT	$\overline{M}$Bbl/d	$\overline{M}$ DWT	$\overline{M}$Bbl/d	$\overline{M}$ DWT	$\overline{M}$Bbl/d	$\overline{M}$ DWT
Venezuela	0.85	1.7	0.25	1.1	--	--	0.78	3.0	--	--
N. Africa	0.65	3.8	0.90	2.7	0.15	1.2	--	--	--	--
W. Africa	1.17	6.5	1.67	6.0	--	--	--	--	--	--
Middle East	2.22	21.3	9.65	96.5	3.88	24.5	0.04	0.3	0.55	2.0
Far East	0.52	3.7	--	--	0.95	3.1	--	--	0.06	0.1
Other	0.31	0.6	--	--	--	--	--	--	--	--
Total		37.6		106.3		28.8		3.3		2.1

GRAND TOTAL = 178.1

Source: A. D. Little Estimates

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*Estimate of tanker demand assumes 15 knot service speed (no slow-steaming).

TABLE A-1-8

CRUDE OIL IMPORTS BY SEA AND TANKER DEMAND IN 1980

(Tanker Capacity in Million DWT)

FROM/TO:	U.S.		West Europe		Japan		Latin America		Other	
	MBbl/d	M DWT	MBbl/d	M DWT	MBbl/d	M DWT	MBbl/d	M DWT	MBbl/d	M DWT
Venezuela	1.05	2.5	0.24	1.0	--	--	0.76	--	--	--
N. Africa	0.81	4.8	0.85	2.6	0.19	1.5	--	3.0	--	--
W. Africa	1.46	8.2	1.57	5.7	--	--	--	--	--	--
Middle East	2.75	26.4	9.08	90.8	4.99	31.4	0.04	--	--	--
Far East	0.65	4.6	--	--	1.22	4.0	--	0.3	0.36	1.3
Other	0.41	0.8	--	--	--	--	--	--	0.04	0.1
Total		47.3		100.1		36.9		3.3		1.4

GRAND TOTAL = 189

Source: A. D. Little Estimates

TABLE A-1-9

CRUDE OIL IMPORTS BY SEA AND TANKER DEMAND IN 1985

(Tanker Capacity in Million DWT)

FROM/TO:	U.S.		West Europe		Japan		Latin America		Other	
	M̄Bbl/d	M̄ DWT	M̄Bbl/d	M̄ DWT	M̄Bbl/d	M̄ DWT	M̄Bbl/d	M̄ DWT	M̄Bbl/d	M̄ DWT
Venezuela	1.41	2.9	0.26	1.1	--	--	0.86	3.4	--	--
N. Africa	1.08	6.4	0.92	2.8	0.21	1.6	--	--	--	--
W. Africa	1.94	10.9	1.70	6.1	--	--	--	--	--	--
Middle East	3.67	35.2	9.83	98.3	5.38	33.9	0.04	0.3	0.45	1.7
Far East	0.86	6.1	--	--	1.31	4.3	--	--	0.05	0.1
Other	0.54	1.0	--	--	--	--	--	--	--	5.4*
Total		62.5		108.3		39.8		3.7		7.2

GRAND TOTAL = 221.5

Source: A. D. Little Estimates

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*Includes 3.0 M̄ DWT for imports from the Middle East to China and 2.4 M̄ DWT for imports from the Middle East to Eastern Europe.

TABLE A-1-10

CRUDE OIL IMPORTS BY SEA AND TANKER DEMAND IN 1990

(Tanker Capacity in Million DWT)

FROM/TO:	U.S.		West Europe		Japan		Latin America		Other	
	M̄Bbl/d	M̄ DWT	M̄Bbl/d	M̄ DWT	M̄Bbl/d	M̄ DWT	M̄Bbl/d	M̄ DWT	M̄Bbl/d	M̄ DWT
Venezuela	1.53	3.1	0.30	1.3	--	--	0.86	3.4	--	--
N. Africa	1.18	7.0	1.04	3.1	0.23	1.8	--	--	--	--
W. Africa	2.12	11.9	1.94	7.0	--	--	--	--	--	--
Middle East	4.00	38.4	11.18	111.8	6.01	37.9	0.04	0.3	0.36	1.3
Far East	0.94	6.7	--	--	1.46	4.8	--	--	0.04	0.1
Other	0.59	1.1	--	--	--	--	--	--	--	15.7
	2.50	--	--	--	--	--	--	--	--	--
Total		68.2		123.2		44.5		3.7		17.1*

GRAND TOTAL = 257

Source: A. D. Little Estimates

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*Includes 6.1 M̄ DWT for imports from the Middle East to China and 9.6 M̄ DWT for imports from the Middle East to Eastern Europe.

multiport loading/discharging, etc.) has resulted in a higher level of employment of the world fleet than would be the case at maximum efficiency. Part of the world fleet has been removed from active tonnage by laying-up. These inefficiencies are the result of the current oversupply of tonnage. Since the demand-supply balance influences the freight levels, the projections of the breakdown of the inefficiencies in slow-steaming and lay-up are restricted to 1980. It is difficult to estimate the future size of the laid-up fleet and it will probably have little effect on the oversupply of tonnage that can be used.

These inefficiencies, considered in terms of the effective reduction of the oversupply of tanker tonnage, are shown in Table A-1-11.

1.2.2.2 Tanker Supply and Costs

An analysis of the composition of the present world fleet by age and size was carried out to determine the total tonnage of tankers that can be used in long haul crude oil trades. The economies of scale in transportation indicate that tankers under 30,000 are very unlikely to be used in these trades. Tankers registered under the U.S. flag are generally not used in international trade and thus have no impact on the size of the fleet considered here. Hence, these two categories of tankers were eliminated to arrive at the aggregate supply of tanker tonnage available to world trade. Estimates of the size of this fleet are presented in Table A-1-12.

From in-house data on the capital costs and operating costs of tankers we calculated the costs of transporting crude oil and products on two selected trade routes--for crude oil, Ras Tanura, Saudi Arabia to Houston (SEADOCK) and for clean products from Houston to New Haven. The voyage characteristics and transportation costs are presented in Tables A-1-13 and 14.

These costs are grouped under the following categories:

- Voyage Costs
  - bunker costs
  - port charges

TABLE A-1-11

TANKER AND COMBINED CARRIER SUPPLY

(Over 30,000 DWT)

	1975 (M DWT)	1980 (M DWT)
Total Supply	263	315*
Less U.S. Flag	6	8
	257	307
Less Laid-Up	46	40
	211	266
Less Slow-Steaming	18	30
	193	236
Demand	178	189
Other Inefficiency	15	47

---

*Assumes 50% of orders not presently under construction will be cancelled and all tankers under 50,000 DWT and built before 1960 are scrapped.

Sources: H.P. Drewry and A.D. Little Estimates

TABLE A-1-12

TANKER WORLD FLEET SUPPLY - BEST CASE

(30,000 DWT and Over)

	1980 (M̄ DWT)	1985 (M̄ DWT)	1990 (M̄ DWT)
Fleet	257	307	287
New Buildings	50	8*	15*
Scrappage[†]		28	60
Total	307	287	242

---

*Estimates. No firm orders known for deliveries in 1985, 1990.

[†] All tankers 20 years and older scrapped.

Source: A.D. Little Estimates

TABLE A-1-13

TANKER VOYAGE CHARACTERISTICS

	VLCC	CLEAN PRODUCTS TANKER
Loading Port	Ras Tanura	Houston
Discharge Port	Houston (SEADOCK)	New Haven
One Way Voyage Distance (NM)	12,502	2,018
Tanker Size (1000 DWT)	260	35
Service Speed (Knots)	15	14
Days at Sea per Voyage	69.43	12.01
Bad Weather Margin	3.47	0.57
Days in Port	4.0	2.5
Total Days per Voyage	76.90	15.08
Operating Days/Year	340	340
Number of Voyages/Year	4.42	22.55
Bunker Consumption at Sea (LT/Day)	174	60
Bunker Consumption at Port (LT/Day)	77	30
Bunker Consumption per Voyage LT	12,993	830
Bunker Costs per Voyage (1000 1975 $)	895	54
Port Charges per Voyage (1000 1975 $)	45	8
Cargo Carried per Voyage (1000 LT)	246.7	33.8
Cargo Delivered per Year (1000 LT)	1,090	703

Source: A.D. Little Estimates

TABLE A-1-14

CURRENT TRANSPORTATION COSTS

(1975 Dollars)

	Ras Tanura - Houston (SEADOCK) 260,000 DWT Non - U.S. Flag Tanker	Houston - New Haven 35,000 DWT Product Tanker U.S. Flag
Capital Cost ($/DWT)	144	410
($ Million)	37.53	14.35
Capital Charges (1000 $/yr)	5,254	1,865
Fixed Operating Costs (1000 $/yr)		
• Crew	526	1,134
• Insurance	892	378
• Repair & Maintenance	670	227
• Miscellaneous	264	75
Voyage Costs (1000 $/yr)		
• Bunker	3,958	1,218
• Port Charges	200	180
Cargo Transported (1000 LT/yr)	1,090	703
Worldscale Index, $/LT	16.08	--
Voyage Costs, $/LT	3.81 (WS 24)	1.99
Cash Costs, $/LT	5.97 (WS 37)	4.57
Fully Built-Up Costs, $/LT	10.79 (WS 67)	7.22

Source: A.D. Little Estimates

- Cash Costs
  - voyage costs
  - fixed operating costs
- Fully Built-Up Costs
  - cash costs
  - capital charges

In the international marketplace freight rates are generally quoted in terms of the Worldscale index which represents the costs of transporting oil by a hypothetical tanker on a given trade route. Market freight rates are quoted in terms of a percentage of the Worldscale index for any given trade route. Presently, spot freight rates for VLCC's on voyages from the Arabian Gulf to all major consuming areas are about WS 27. The estimates of transportation costs indicate that at this level of freight rates owners are recovering slightly more than their voyage costs (Table A-1-14).

Present charter rates for periods of 2-5 years for VLCC's are approximately WS 50. At this level owners will probably recover all of their cash costs and make some contribution to capital costs. At WS 67 the freight rates of VLCC's will allow the owners to recover their fully built-up costs.

Since the estimate of the demand and supply of tanker tonnage shown in Figure A-1-5 indicates these to be in balance no earlier than 1990, we estimate that freight rates are unlikely to reach the levels of fully built-up costs (WS 67) before then. We believe that in the period between 1977 and 1990 VLCC freight rates will gradually increase from their present level of WS 27 to reach WS 67 by 1990. The projected freight rate levels are also shown in Figure A-1-5. In the case of U.S. flag tankers, the freight rates are assumed to be close to fully built-up costs because the U.S. flag tanker fleet is expected to remain in balance with demand in U.S. domestic trade.

The projected freight rates on all the routes under consideration here are presented in Table A-1-15. In the case of Arabian light crude oil moving to the U.S. Gulf Coast, a direct VLCC shipment to a major offshore

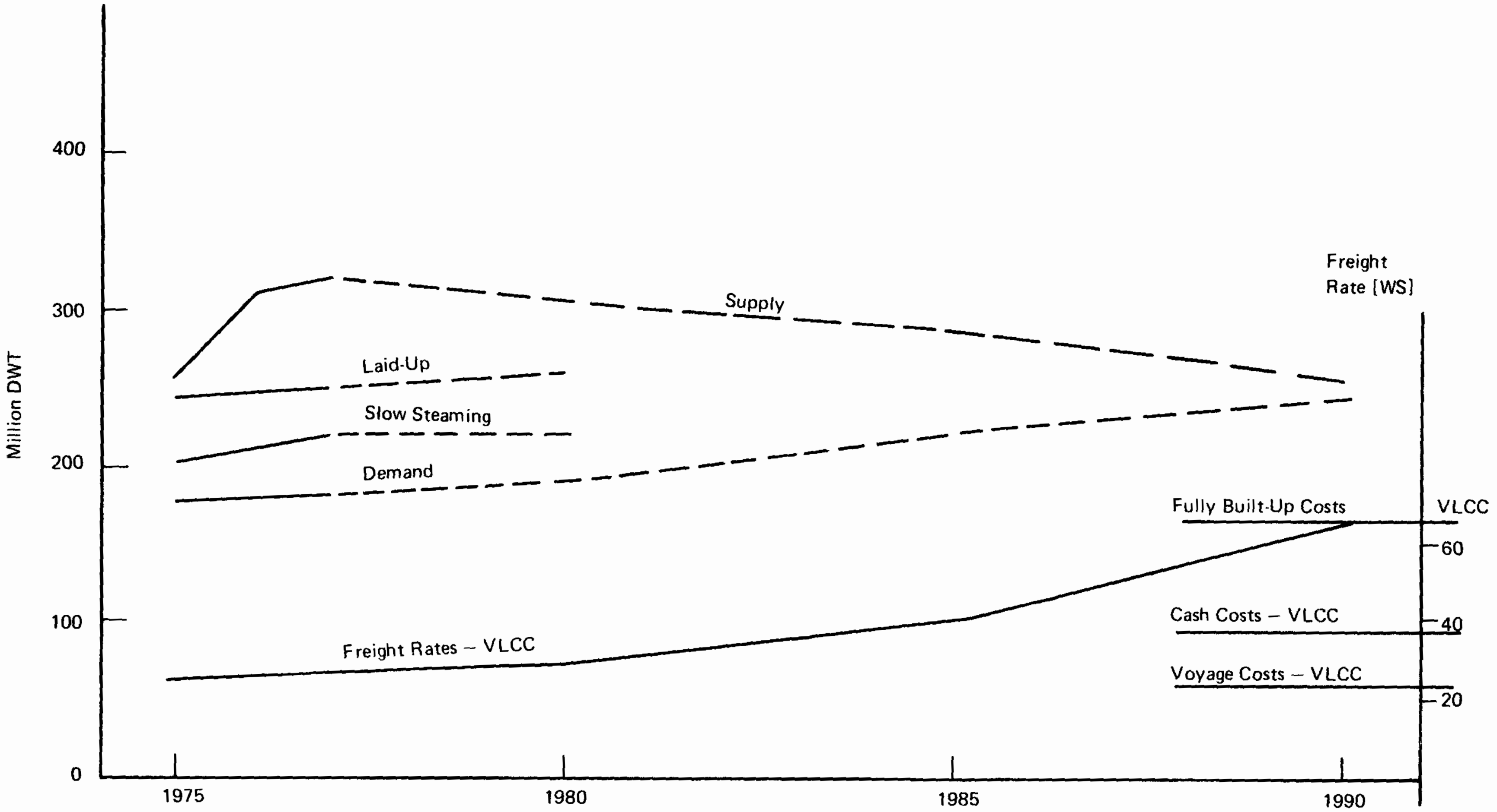

Figure A-1-5 WORLD TANKER FLEET DEMAND/SUPPLY (Non-U.S. Flag Tankers Above 30,000 DWT)

TABLE A-1-15

SUMMARY OF OCEAN OIL TRANSPORTATION COSTS

(1975 Dollars)

Hydrocarbon Form	Voyage Characteristics		Vessel Size 1000 DWT	Worldscale Index $/LT	1980			1985			1990		
	From	To			WS	$/LT	$/Bbl	WS	$/LT	$/Bbl	WS	$/LT	$/Bbl
1. Non-U.S. Flag													
Arabian Light Crude	Ras Tanura	Houston	260	16.08	30	4.82	0.64	40	6.43	0.87	67	10.77	1.43
Arabian Light Crude	Ras Tanura	Los Angeles	165	14.64	40	5.86	0.78	50	7.32	0.97	75	10.98	1.46
Arabian Light Crude	Ras Tanura	Yokohama	260	8.81	30	2.64	0.35	40	3.52	0.47	67	5.90	0.78
North Slope Crude	Valdez	Yokohama	260	5.20	30	1.56	0.22	40	2.08	0.29	67	3.48	0.48
2. U.S. Flag													
North Slope Crude	Valdez	Los Angeles	165	full cost	n/a	3.67	0.51	n/a	3.67	0.51	n/a	3.67	0.51
North Slope Crude	Valdex	Houston (via Panama)	60	full cost	n/a	20.30	2.82	n/a	20.30	2.82	n/a	20.30	2.82
Clean Products	Houston	New Haven	35	full cost	n/a	7.22	0.86	n/a	7.22	0.86	n/a	7.22	0.86
Dirty Products	Houston	New Haven	35	full cost	n/a	6.83	1.02	n/a	6.83	1.02	n/a	6.83	1.02
Asphalt	Houston	New Haven	35	full cost	n/a	8.04	1.34	n/a	8.04	1.34	n/a	8.04	1.34
Methanol	Houston	New Haven	35	full cost	n/a	8.29	1.03	n/a	8.29	1.03	n/a	8.29	1.03

Source: A.D. Little Estimates

oil terminal (such as LOOP or SEADOCK) was assumed. Terminal charges were assessed at 24¢/bbl.

1.2.3 Refinery Gate Crude Oil Costs

In Section 1.2.1 estimates were developed (in 1975 dollars) of the FOB price Ras Tanura of 34° Arabian light crude oil for 1980, 1985, and 1990. The purpose of the transportation cost analysis outlined in Section 1.2.2.2 was to highlight the environment for future tanker freight rate and cost development in order to assess the costs of moving crude oil from the loading terminal to a coastal refinery and, subsequently, refined products from the U.S. Gulf Coast to New Haven.

The assessment of the evolution of refining patterns and of environmental constraints indicates that most new domestic refining capacity serving PADD's I-IV will be constructed on the U.S. Gulf Coast. Arabian light crude oil is assumed to move in VLCC's directly to an offshore oil superport (such as LOOP or SEADOCK) on the U.S. Gulf, and then to the refinery gate of a new refinery located in the Houston area, since imported crude oil will be the marginal supply source in this refining area. Costs for this activity are illustrated in Table A-1-16.

TABLE A-1-16

PADD I-IV REFINERY GATE CRUDE COSTS

(1975 $/Bbl)

	1980	1985	1990
FOB Price of Arabian Light Crude Oil	11.42	16.93	17.83
Crude Oil Transportation	0.64	0.87	1.43
Superport Fee	0.24	0.24	0.24
Refinery Gate Crude Price*	12.30(2.10)	18.04(3.08)	19.50(3.33)

---

*Values in parenthesis are $/$\overline{\text{M}}$Btu.

Source: A.D. Little Estimates

In the case of PADD V, it was assumed that the marginal crude oil available for incremental refineries will be Alaskan North Slope oil. There is some uncertainty as to whether or not environmental opposition will prevent the construction of any new refineries in PADD V. However, the refinery supply/demand analysis indicates that additional refining capacity required in PADD V prior to 1990 could probably be obtained by debottlenecking and expansions at existing refinery locations. A further constraint in PADD V is created by the problem of disposal of surplus North Slope crude. Table A-1-17 clearly shows that the highest value for North Slope oil is obtained if all the crude is refined in PADD V. The lowest marginal value of all (some $2/Bbl lower than the PADD V absorption case) is for North Slope crude to be refined in the U.S. Gulf. Intermediate values would be obtained for other logistical options currently under investigation for the disposal of surplus Alaskan crude (e.g., Sohio, Kitimat, and Guadelupe Dunes pipeline proposals), or if Alaskan crude were exported to Japan. The Japanese export value scenario was used as a guideline in order to establish a reasonable value at which North Slope oil may be valued at the margin, and penalties of 20¢/bbl in 1980, 30¢/bbl in 1985 and 59¢/bbl in 1990 were deducted from the landed cost of North Slope oil in Los Angeles.* This produces refinery gate values for North Slope oil of (assuming the crude oil has the quality characteristics of Arabian light):

1980 - $12.00/bbl
1985 - $17.60/bbl
1990 - $18.70/bbl

Since North Slope oil has slightly inferior quality characteristics when compared with Arabian light oil, a quality differential to adjust the above crude prices to true North Slope refinery gate prices was determined. These prices are given in Table A-1-18.

Also considered, for the purposes of the refining cost allocation, was a price control case in which domestic crude oil prices would remain under FEA jurisdiction. On the advice of EPRI, the domestic crude oil

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*Scenario 1, Table A-1-17.

TABLE A-1-17

PROJECTED INTERNATIONAL PARITY PRICES FOR NORTH SLOPE* CRUDE AT LONG BEACH

($/Bbl - 1975 Dollars)

	1. All North Slope Refined in PADD V			2. Surplus North Slope Exported to Japan			3. Surplus North Slope moved via Panama Canal to PADD III		
	1980	1985	1990	1980	1985	1990	1980	1985	1990
FOB Value of Arabian Light	11.42	16.93	17.83	11.42	16.93	17.83	11.42	16.93	17.83
+ Ras Tanura - Houston†	-	-	-	-	-	-	0.88	1.11	1.67
- Yokohama	-	-	-	0.35	0.47	0.78	-	-	-
- Los Angeles	0.78	0.97	1.46	-	-	-	-	-	-
- Valdez - Houston	-	-	-	-	-	-	(2.82)	(2.82)	(2.82)
- Yokohama	-	-	-	(0.22)	(0.29)	(0.48)	-	-	-
+ Valdez - Los Angeles	-	-	-	0.51	0.51	0.51	0.51	0.51	0.51
Landed Value of North Slope Crude* at Los Angeles	12.20	17.90	19.29	12.06	17.62	18.64	9.99	15.73	17.19

---

*Excludes North Slope quality differential vs. Arabian light

†Includes SEADOCK terminal fee.

Source: A.D. Little Estimates

TABLE A-1-18

PADD V REFINERY GATE CRUDE COSTS

(1975 $/Bbl)

	1980	1985	1990
Landed Value of "North Slope"* oil at Los Angeles	12.20	17.90	19.29
Surplus Oil Disposition Penalty	(0.20)	(0.30)	(0.60)
North Slope Quality Adjustment	(0.18)	(0.12)	(0.05)
Refinery Gate Crude Price†	11.82 [2.02]	17.48 [2.99]	18.64 [3.18]

---

* "North Slope" crude oil of Arabian Light quality.

† Values in brackets are $/$\bar{M}$ Btu.

Source: A. D. Little Estimates

price projections and domestic/imported crude proportions from the RP-759-2 study prepared in March 1977 by Foster Associates was used. The weighted average refinery gate crude prices under a domestic price controls environment are shown in Table A-1-19.

1.2.4 Refining Costs and Ex-Refinery Products Prices

1.2.4.1 The Theory of Refinery Cost Allocation

The United States petroleum industry, including certain offshore Caribbean refineries serving East Coast import requirements, is comprised of four basic refinery types. The Topping Refinery of Figure A-1-6 consists principally of a distillation column and associated offsites to separate crude oil directly into the primary products of naphtha, gas oil, and fuel oil. If the crude oil is of high sulfur content, these products will be similarly of high sulfur content. Such refineries are used for production of specialty products, such as naphtha (for petrochemical feedstocks) or asphalt.

If gasoline production is required, the addition of a catalytic naphtha reformer is common, thereby creating a Hydroskimming Refinery (Figure A-1-7). If a low sulfur crude is used, such a refinery would produce gasoline and low sulfur fuel oil products. With a high sulfur crude, either high sulfur fuel oils or specialty products such as asphalt are produced.

With the current environmental restrictions on the use of high sulfur fuel oils, it is becoming increasingly common in the United States and Caribbean refineries to desulfurize the fuel oils produced from high sulfur crude oils. Since high sulfur crude oil (such as Arabian light) will be increasingly used in U.S. refineries as crude oil imports grow, it is not surprising that two current major U.S. refinery expansions will employ fuel oil desulfurization, converting them to Fuels Refineries similar to that in Figure A-1-8.

In the United States the demand for gasoline has substantially exceeded the supply available naturally from crude oil through refinery processing types typified by Figures A-1-7 and 8. Conversion refineries,

TABLE A-1-19

REFINERY GATE CRUDE OIL COSTS WITH CONTROLS

(1975 $/Bbl)

	GULF COAST			WEST COAST		
Refinery Location:	1980*	1985[†]	1990[†]	1980*	1985	1990
Landed Imported Crude*	12.30	18.04	19.50	12.20[§]	17.90	19.29
Percent	47	39	32	47	39	32
Weighted Domestic Crude	10.03	10.47	12.20	10.03	10.47	12.20
Transportation	0.14	0.14	0.14	0.09	0.09	0.08
Delivered Price	10.17	10.61	12.34	10.12	10.56	12.28
Percent	53	61	68	53	61	68
Weighted Average Crude Price	11.17	13.51	14.63	11.10	13.42	14.52

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* A. D. Little, Inc. Estimates

[†] Foster Associates Forecast

[§] Represents the direct landed cost of imported crude oil (Arabian Light) on the U.S. West Coast.

Source: Foster Associates & A. D. Little Estimates

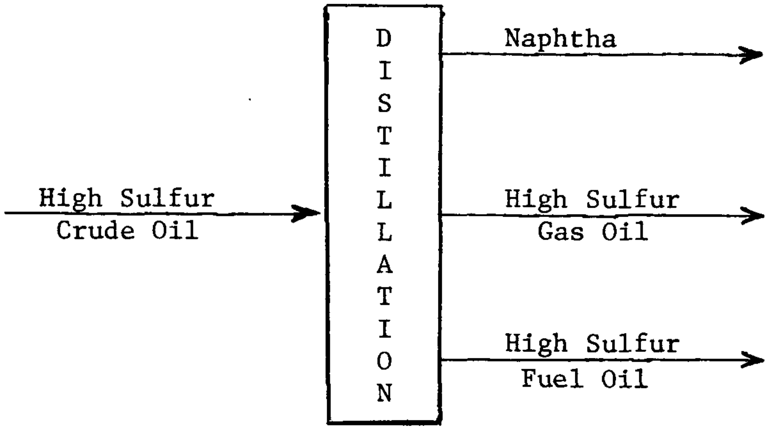

FIGURE A-1-6

TOPPING REFINERY

* * * * *

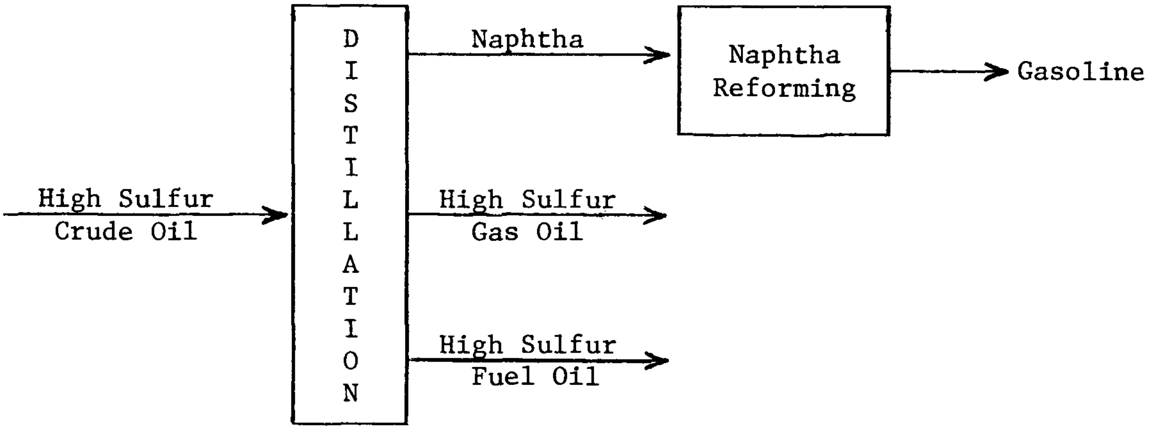

FIGURE A-1-7

HYDROSKIMMING REFINERY

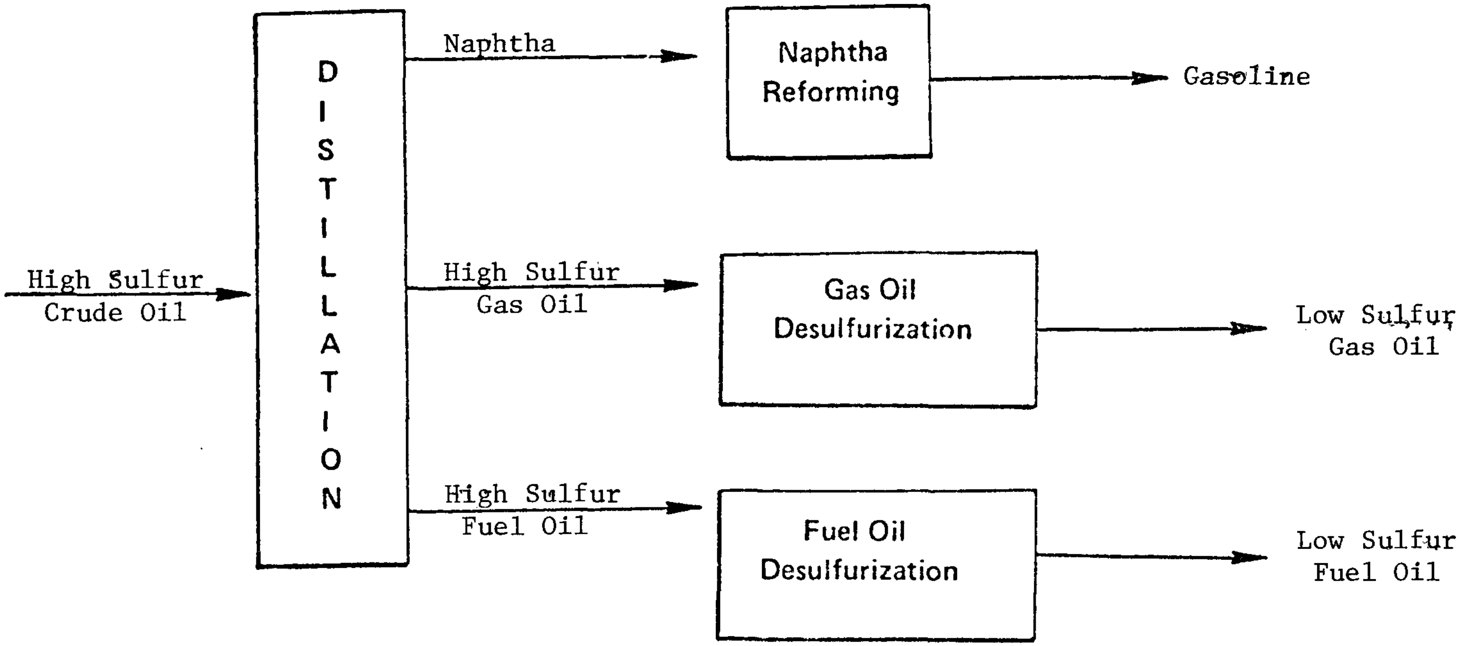

FIGURE A-1-8

FUELS REFINERY

shown in Figure A-1-9, have thus been constructed to convert excess fuel oil into gasoline by catalytic cracking or hydrocracking, providing an historic average yield of gasoline of 50% of crude processed.

In order to make projections of product prices, econometric models of the composite of these refinery types representing U.S. industry must be developed, reflecting both the anticipated changes in crude oil distillation and other processing costs as well as the future profitability of refining operations. For example, Table A-1-20 illustrates the effect of refinery profitability on required product prices for one of these refinery types--a conversion refinery processing Arabian light crude oil. Refinery profitability is measured by a capital charge factor, reflecting annual cash flow necessary to:

- repay debt financing, including interest;
- pay taxes, after allowance for depreciation and interest;
- earn a return on equity in the project.

Under conventional fiscal regimes with about 50% corporate tax rates, a capital charge of around 25% is necessary to yield a 12-15% discounted cash flow rate of return over a 20-year project life.

The conversion refinery will produce about 93,000 bbl/d of products from 100,000 bbl/d of crude (the remaining 7,000 bbl/d being consumed in refinery fuel and losses) consuming no natural gas, purchased natural gasoline or butanes. A zero percent capital charge is reflected in the set of product prices which average $13.72/bbl, yielding $1.276 million/day of total revenues. At a crude oil cost of $12/bbl, fixed operating costs of 51¢/bbl and variable operating costs of 25¢/bbl, these revenues are only adequate to cover processing costs, yielding a zero capital charge.

By contrast, if product prices average $16.62/bbl, product revenues of $1.546 million/day will be adequate to meet all processing costs and, in addition, return $270,000/day to capital recovery--equivalent to a 20% capital charge on the investment shown in Table A-1-20.

Over an extended time interval and in a free market environment, the prices of individual petroleum products should reflect the full investment

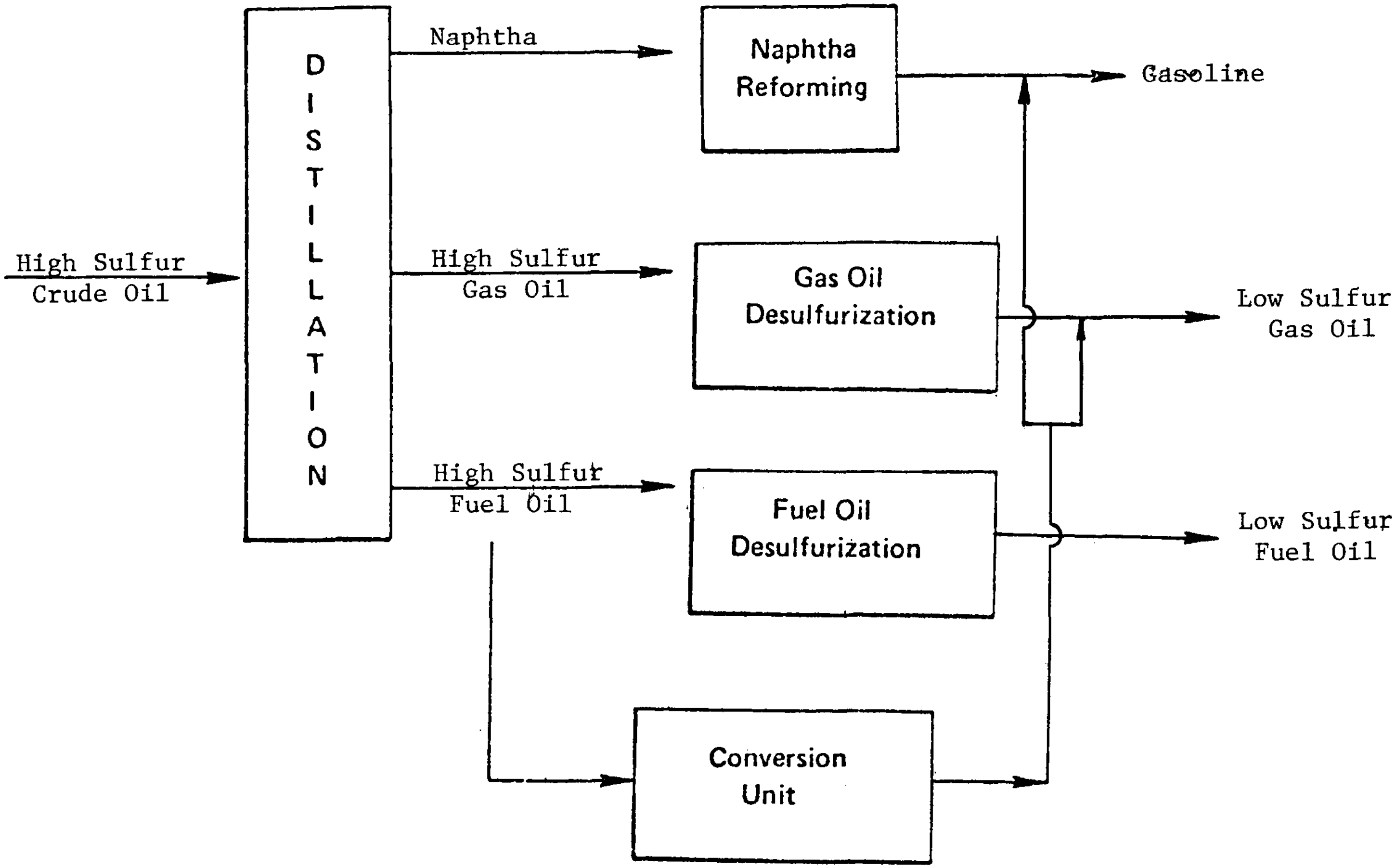

FIGURE A-1-9

CONVERSION REFINERY

TABLE A-1-20

ILLUSTRATIVE CASH FLOW SUMMARY FOR CONVERSION REFINERY

100,000 Bbl/d Arabian Light Crude Oil
($446 Million Investment)

CAPITAL CHARGE ON TOTAL REFINERY		0%		8%		20%	
Products	000 Bbl/d	$/Bbl	$/KBbl/d	$/Bbl	$/KBbl/d	$/Bbl	$/KBbl/d
Gasoline	46.7	14.27	666	15.40	719	17.10	799
Low Sulfur Gas Oil	29.8	13.52	403	14.71	438	16.50	492
0.5% Sulfur Fuel Oil	16.5	12.52	207	13.71	226	15.50	256
Total	93.0	13.72	1276	14.88	1384	16.62	1546
Costs							
Crude Oil Cost	100.0	12.00	1200	12.00	1200	12.00	1200
Fixed Operating Costs	-	0.51	51	0.51	51	0.51	51
Variable Costs	-	0.25	25	0.25	25	0.25	25
Capital Charges	-	-	-	1.08	108	2.70	270
Total		12.76	1276	13.84	1384	15.46	1546

Source: A.D. Little Estimates

required to construct new refineries, the full operating costs for these refineries, and the full cost of conversion of one product into another. The achievement of these price levels should be required to justify the new investments needed to meet future demand growth. For example, the price of gasoline should reflect some proportion of the total operating and capital costs of a new refinery, with that proportion being dictated primarily by the costs of conversion of fuel oil into gasoline. In the shorter term, however, supply/demand imbalances may create market forces preventing the full recovery of refining costs. For example, if refinery capacity is generally in excess, it would not be expected that a new refinery would recover the full capital costs of its construction. Furthermore, governmental actions can interfere with these free market forces, preventing equilibrium prices of the products from being achieved. For example, price controls on gasoline may require reallocation of the refining costs into other refinery products, even when the full capital charge on the refinery is realized.

Clearly these models must be capable of reflecting not only future real changes in crude oil prices, but also future changes in overall refinery profitability as well as individual processing unit profitability. This is achieved by constructing a series of equations for the four basic refinery types, described in Table A-1-21, relating processing costs to the composite prices of products produced. These equations, which are constructed to simulate the U.S. and Caribbean refining industry, are then solved simultaneously to provide the product price projections.

In the first equation of Table A-1-21, for example, the average price of petroleum products from any of the refinery types must be related to the total cost of processing, including crude costs, cash operating costs, and capital charges. If the available supply of conversion refineries greatly exceeds the demand for products from these refineries or if governmental policies regarding product or crude pricing are unduly restrictive, little or no return on capital may be achieved. In this case, the zero percent capital charge entry of Table A-1-20 may be applicable or, in cases of even more extreme excess supply, not all of the operating costs may be recovered. Under these conditions less efficient refineries would be eventually shut down, bringing supply more in balance with demand.

TABLE A-1-21

EQUILIBRIUM PRICE RELATIONSHIPS

EQUATION	BASIS
1. Crude + Refining Costs = Value of all Products	All products jointly bear refining costs
2. Naphtha + Reforming Costs = Gasoline	(Gasoline-Naphtha) margin justifies reforming unit investment
3. High Sulfur Gas Oil + Desulfurization Costs = Low Sulfur Gas Oil	Margin justifies desulfurization unit investment
4. High Sulfur Fuel Oil + Desulfurization Costs = 0.5% S Fuel Oil	Margin justifies desulfurization unit investment
5. High Sulfur Fuel Oil + Desulfurization Costs = 0.3% S Fuel Oil	Margin justifies desulfurization unit investment
6. High Sulfur Fuel Oil + Conversion Costs = Gas Oil & Gasoline	Margin justifies conversion unit investment
7. Low Sulfur Gas Oil + 0.5% S Fuel Oil = 0.3% S Fuel Oil	Blending values relate gas oil to fuel oil
8. Kerosene = Linear Interpolation of Naphtha and Gas Oil Values	Changing cut points will adjust kerosene production
9. Vacuum Bottoms + Gas Oil = High Sulfur Fuel Oil	Blending values relate vacuum bottoms to fuel oil

Source: A.D. Little Estimates

By contrast if, in a free market environment, a strong demand for this conversion refinery type is foreseen, margins will rise allowing sufficiently profitable operation to justify the construction of new refineries of this type, perhaps to the 25% capital charge level of Table A-1-20. In view of the recent rapid escalation of refinery construction costs, shown in Figure A-1-10, margins would have to increase substantially in the future to provide the incentive for new refinery construction.

1.2.4.2 Refinery Supply and Demand

As shown in Table A-1-22, the eastern United States (PADD's I-IV) net demand on refineries for petroleum products will rise to over 17 million bbl/d by 1985, from a current level of about 13 million bbl/d. This projection, shown in Figure A-1-11, takes into account expected levels of gasoline conservation, coal exploitation, nuclear utilization, and natural gas availability. Historic refining capacity is also shown in Figure A-1-11, along with those announced additions to refining capacity considered to be firm. Historically, demand has exceeded eastern U.S. refining capacity, with the balance supplied by the Caribbean and, to a lesser degree, eastern Canada.

It is apparent that the combined refinery capacity will meet projected demand by about 1983. The eastern U.S. refinery capacity utilization has historically been high, with the Caribbean refineries underutilized. By 1983, new capacity must be constructed or products imported from outside this region must be attained, both of which would result in an upward pressure on refining margins.

In Figure A-1-12 the changing character of the PADD I-IV product demand barrel during the coming decade is shown. For reference purposes, the 1975 demand is shown to consist of 54% gasoline and light distillates, and 21% heavy products. This high demand for gasoline has motivated the installation of a high percentage of conversion refineries in the U.S., since refineries provide yield patterns typical of the demand barrel.

The incremental demand projected beyond 1975 will become progressively heavier through 1990, reflecting successful gasoline conservation and

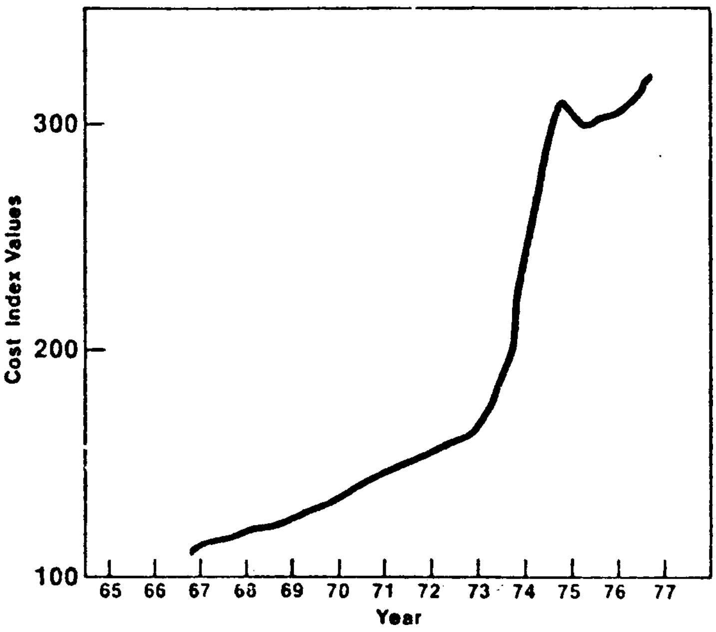

FIGURE A-1-10

REFINING & PETROCHEMICAL PROCESS FACILITIES
TOTAL ERECTED COST INDEX

Source: C.R. Sitter, Senior Vice President, Exxon, "The Automotive News World Congress", Detroit, Michigan, July 11 (1977).

TABLE A-1-22

REFINERY SUPPLY AND DEMAND
PADD's I-IV

(Million Bbl/d)

	REFINERY CAPACITY				REFINERY DEMAND				
	PADD's I-IV		Caribbean & E. Canada	Total Available	Product Demand	Processing Gain	NGL Production	Non-Caribbean Imports	Net Demand on Refineries
	Capacity	Available at 90% Stream Factor	Net Surplus at 90% Stream Factor						
1970	10.30	9.27	2.55	11.82	12.74	(0.31)	(1.61)	(0.18)	10.64
1971	10.77	9.69	2.71	12.40	13.18	(0.32)	(1.64)	(0.20)	11.02
1972	11.07	9.96	2.89	12.85	14.20	(0.33)	(1.70)	(0.23)	11.94
1973	11.63	10.47	3.07	13.54	14.97	(0.37)	(1.70)	(0.24)	12.66
1974	12.20	10.98	3.24	14.22	14.46	(0.40)	(1.66)	(0.25)	12.15
1975	12.47	11.22	3.40	14.62	14.11	(0.38)	(1.60)	(0.19)	11.94
1976	12.97	11.67	3.30	14.97	15.09	(0.37)	(1.58)	(0.17)	12.97
1977	13.48	12.13	3.26	15.39	-	-	-	-	-
1978	13.97	12.57	3.24	15.81	-	-	-	-	-
1979	14.37	12.93	3.25	16.18	-	-	-	-	-
1980	14.55	13.10	3.20	16.30	16.58	(0.20)	(1.40)	(0.15)	14.83
1985	14.55	13.10	2.95	16.05	18.78	-	(1.30)	(0.15)	17.33
1990	14.55	13.10	2.70	15.80	18.65	0.25	(1.30)	(0.15)	17.45

Source: A.D. Little Estimates

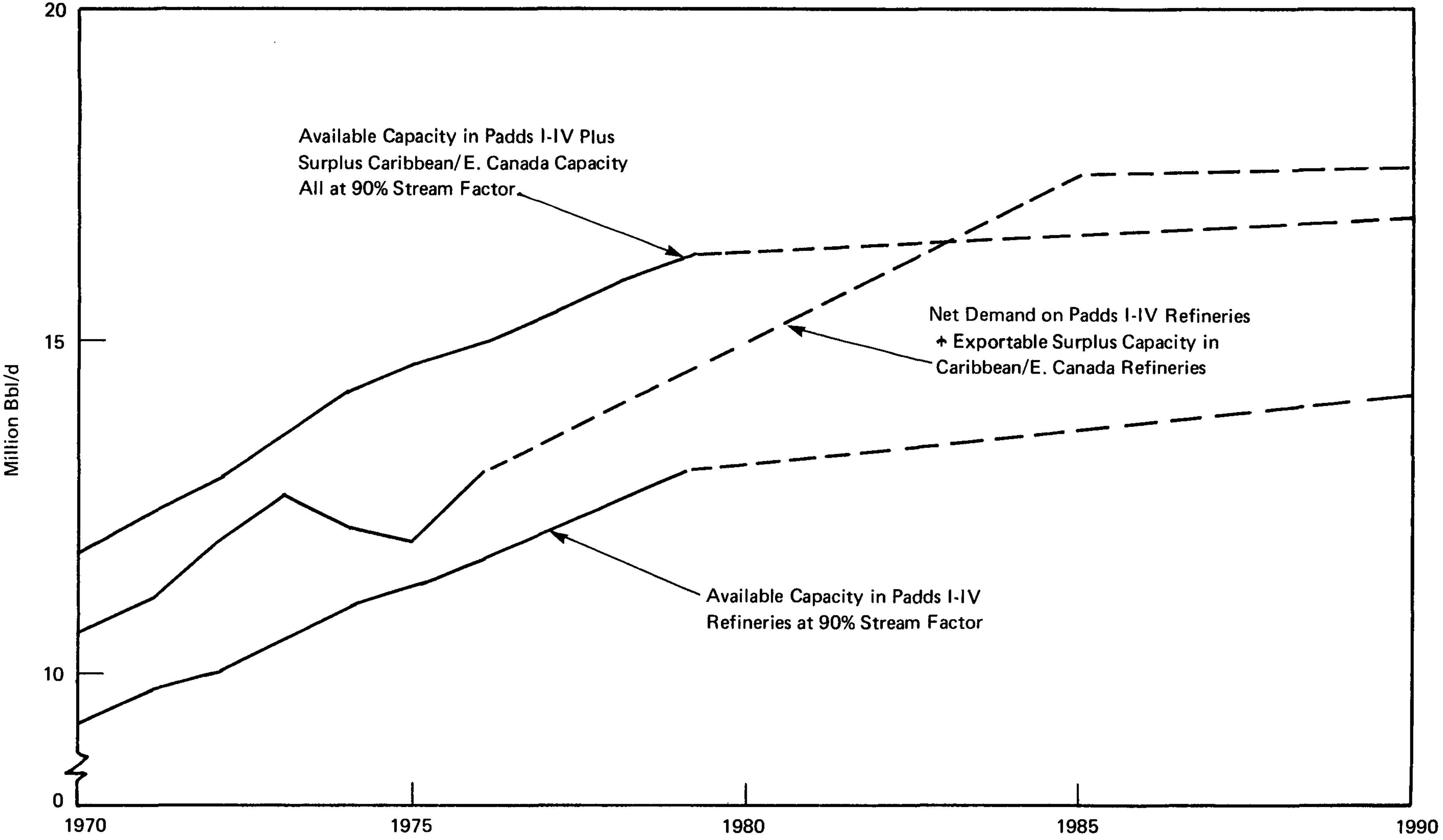

Figure A-1-11 U.S. PADDS I-IV REFINERY SUPPLY/DEMAND BALANCE

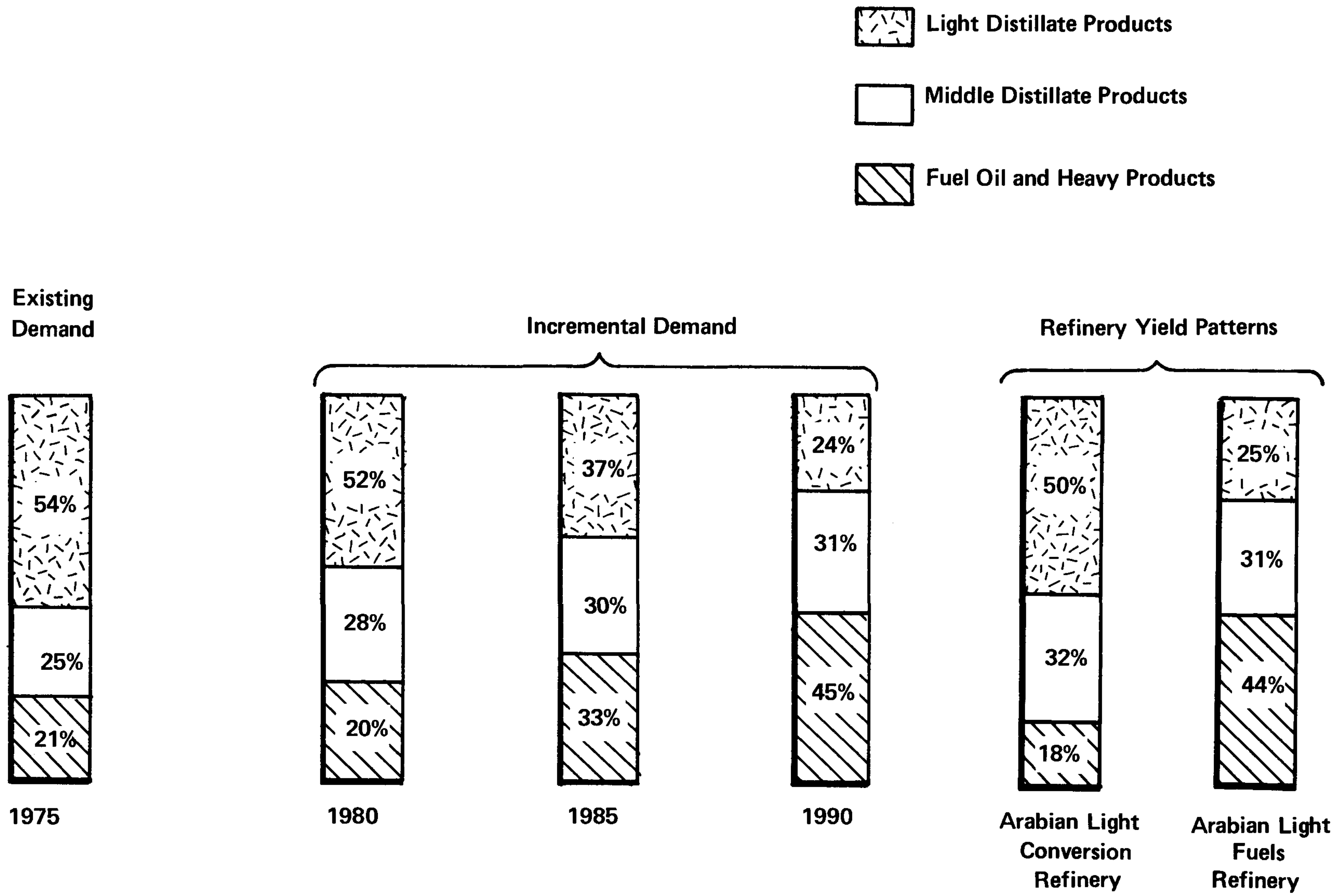

Figure A-1-12 REFINED PRODUCTS DEMAND DISTRIBUTION AND REFINERY YIELD PATTERNS PADDSI–IV insufficient availability of usable coal, nuclear energy and natural gas. Compared to the 1990 incremental demand barrel, the product slate derived from the fuels refinery is seen to be more closely aligned than that from a conversion refinery. This move toward a heavier demand barrel represents a profound shift in the historic trend which had previously emphasized gasoline at the expense of fuel oil. This changing demand barrel will cause a realignment in the relative product pricing relationships.

Therefore, it was concluded that overall refining margins in the eastern U.S. will rise over the coming decade to support new additions to refining capacity indicated in Figure A-1-11 to be necessary by the early 1980's. However, as indicated by Figure A-1-12, this margin improvement will probably be borne more heavily by fuel oil products.

The effect of increasing demand for heavy fuel oil products at the expense of gasoline is illustrated in Table A-1-23, for a fixed price of Arabian light crude oil. Because of the requirement for new refinery construction, the capital charge on the overall refinery was increased uniformly over time. However, the capital charge for the conversion units was maintained at a high level for the first two columns and then reduced for the third column of Table A-1-24 to illustrate the price implications of the heavier product barrel of Figure A-1-12. It is seen that the light product prices (gasoline and naphtha) would go through a maximum, whereas the fuel oil products (gas oil and fuel oil) progressively increase. Since the light products prices are reduced, the heavy products prices must be increased to bear the additional overall refining margin. The specific price entries of Tables A-1-23 and 24 are provided for illustrative purposes only, and do not represent the projected absolute magnitudes of these effects, because future demand projections are not totally independent of price.

Superimposed on the demand shift is the effect of varying crude prices shown in Table A-1-24 which presents the projections of the A.D. Little price model for a fixed, arbitrary capital charge for all refinery processes as the crude oil price is increased. The products from the refinery processing Arabian light crude oil generally increase in price uniformly,

TABLE A-1-23

ILLUSTRATIVE EFFECT OF CAPITAL CHARGES ON EX-REFINERY PRODUCT VALUES

(%/Bbl)

Based on $12/Bbl Arabian Light

Capital Charges Refinery/Process Unit	Near Term Low/High	Medium Term Intermediate/High	Long Term High/Low
Gasoline	16.45	17.30	16.54
Naphtha	13.68	14.47	14.20
Gas Oil	14.52	15.42	15.61
0.5% S Fuel Oil	13.82	14.72	14.91

Source: A.D. Little Estimates

TABLE A-1-24

ILLUSTRATIVE EFFECT OF CRUDE OIL PRICES ON EX-REFINERY PRODUCT VALUES

($/Bbl)

Intermediate Capital Charges

CRUDE PRICE:	10.00	15.00	20.00	25.00
Gasoline	13.18	19.43	24.68	29.92
Naphtha	11.77	16.69	21.60	26.51
Gas Oil	12.64	18.15	23.65	29.16
0.5% S Fuel Oil	11.94	17.45	22.95	28.46
3.0% S Fuel Oil	9.13	14.26	19.38	24.51

Source: A.D. Little Estimates

with the spread between the product prices being influenced by processing unit yields and fuel consumption (whose costs are a function of the crude oil price). For example, the margin between gasoline and high sulfur fuel oil, dictated by conversion costs, increases from $4.05/bbl at a $10/bbl crude price to $5.41/bbl at $25/bbl crude oil. The margin between 0.5% sulfur fuel oil and high sulfur fuel oil increases from $2.81/bbl to $3.95/bbl over the same range of crude oil prices.

As shown in Table A-1-25 and Figures A-1-13 and 14, similar trends are projected to occur in PADD V. The need for new capacity in the early 1980's will require increased overall refining margins, and the greatest incremental demand is represented by fuel oil. Therefore, fuel oil prices must increase relatively faster than light products in this refining region also.

1.2.4.3 Basis for Price Projections

The incremental refining capacity required to serve PADD's I-IV was assumed to be located in Houston, due to favorable port facilities, crude oil and products logistics systems, land availability, positive administrative attitudes (taxes, environmental regulations, etc.), petrochemical interface, and being at the center of existing industry. Product price projections in Hartford, Dallas, and Columbus were then obtained by adding transportation costs (Section 1.2.5). Because the marginal barrel of crude oil will be imported, however, differential crude and product price transportation costs are not in general so large that the selection of refinery location will greatly influence projected product prices.

The incremental refining capacity to be installed to serve PADD V was assumed to be installed in Los Angeles due to its proximity to the chosen demand center in PADD V, and proximity to North Slope crude off-loading facilities expected at Long Beach. Particularly for PADD V, the incremental transportation costs for other possible refining locations are sufficiently small that the Los Angeles location is not critical to the projected product prices.

TABLE A-1-25

REFINERY SUPPLY AND DEMAND PADD V

(Million Bbl/d)

	REFINERY CAPACITY		REFINERY DEMAND				
	Capacity	Available at 90% Stream Factor	Product Demand	Processing Gain	NGL Production	Non-Caribbean Imports	Net Demand on Refineries
1970	1.97	1.77	1.95	(0.05)	(0.05)	(0.08)	1.77
1971	2.08	1.87	2.04	(0.06)	(0.05)	(0.12)	1.81
1972	2.19	1.97	2.16	(0.06)	(0.04)	(0.12)	1.94
1973	2.25	2.03	2.34	(0.08)	(0.04)	(0.16)	2.06
1974	2.30	2.07	2.19	(0.09)	(0.03)	(0.14)	1.93
1975	2.36	2.12	2.22	(0.08)	(0.03)	(0.11)	2.00
1976	2.59	2.33	2.35	(0.09)	(0.03)	(0.10)	2.13
1977	2.98	2.68	-	-	-	-	-
1978	3.02	2.72	-	-	-	-	-
1979	3.05	2.75	-	-	-	-	-
1980	3.08	2.77	2.92	(0.08)	(0.02)	(0.10)	2.72
1985	3.08	2.77	3.32	(0.06)	(0.05)	(0.10)	3.11
1990	3.08	2.77	3.55	(0.04)	(0.10)	(0.10)	3.31

Source: A.D. Little Estimates

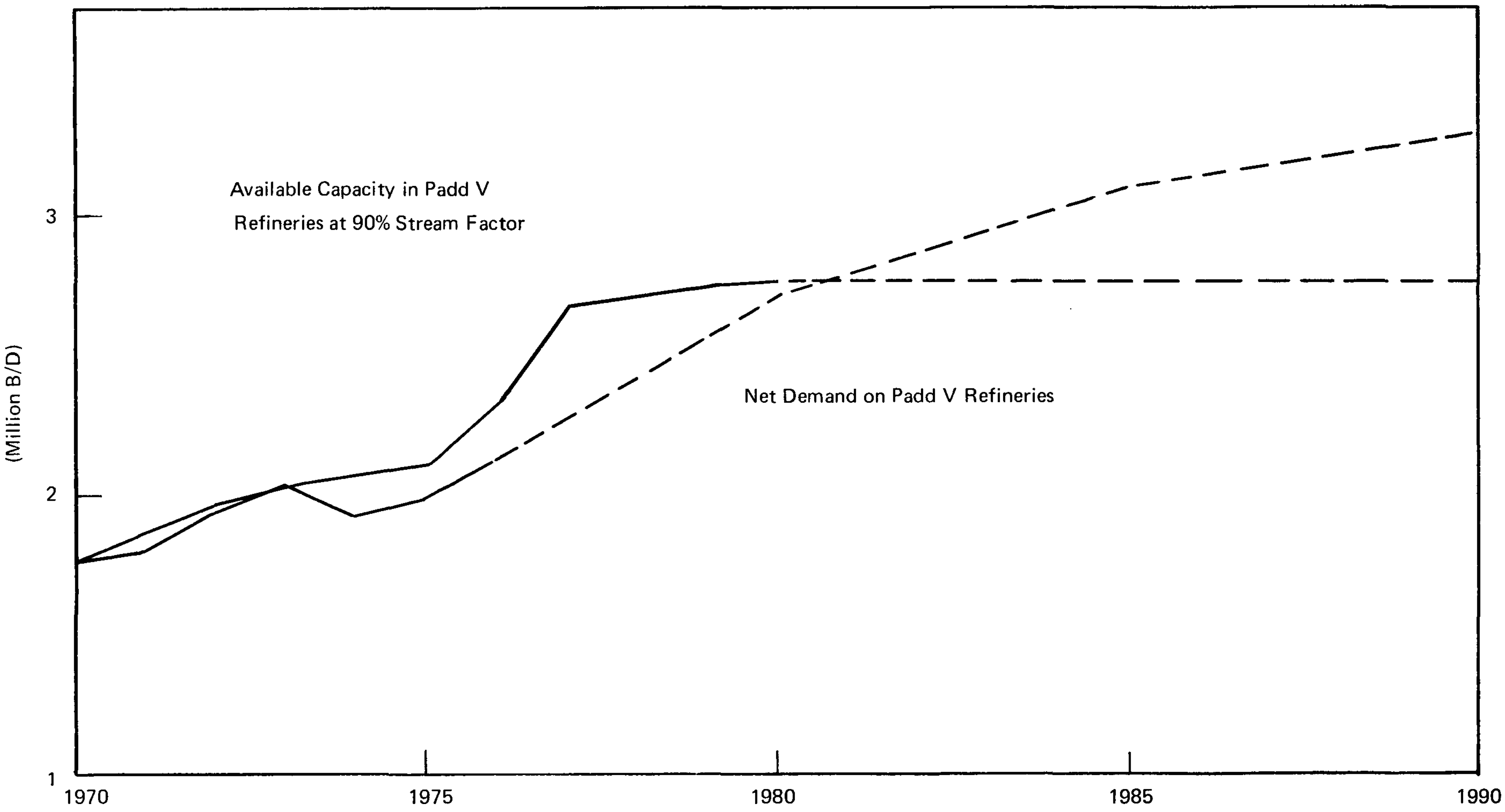

Figure A-1-13 U.S. PADD V REFINERY SUPPLY/DEMAND BALANCE

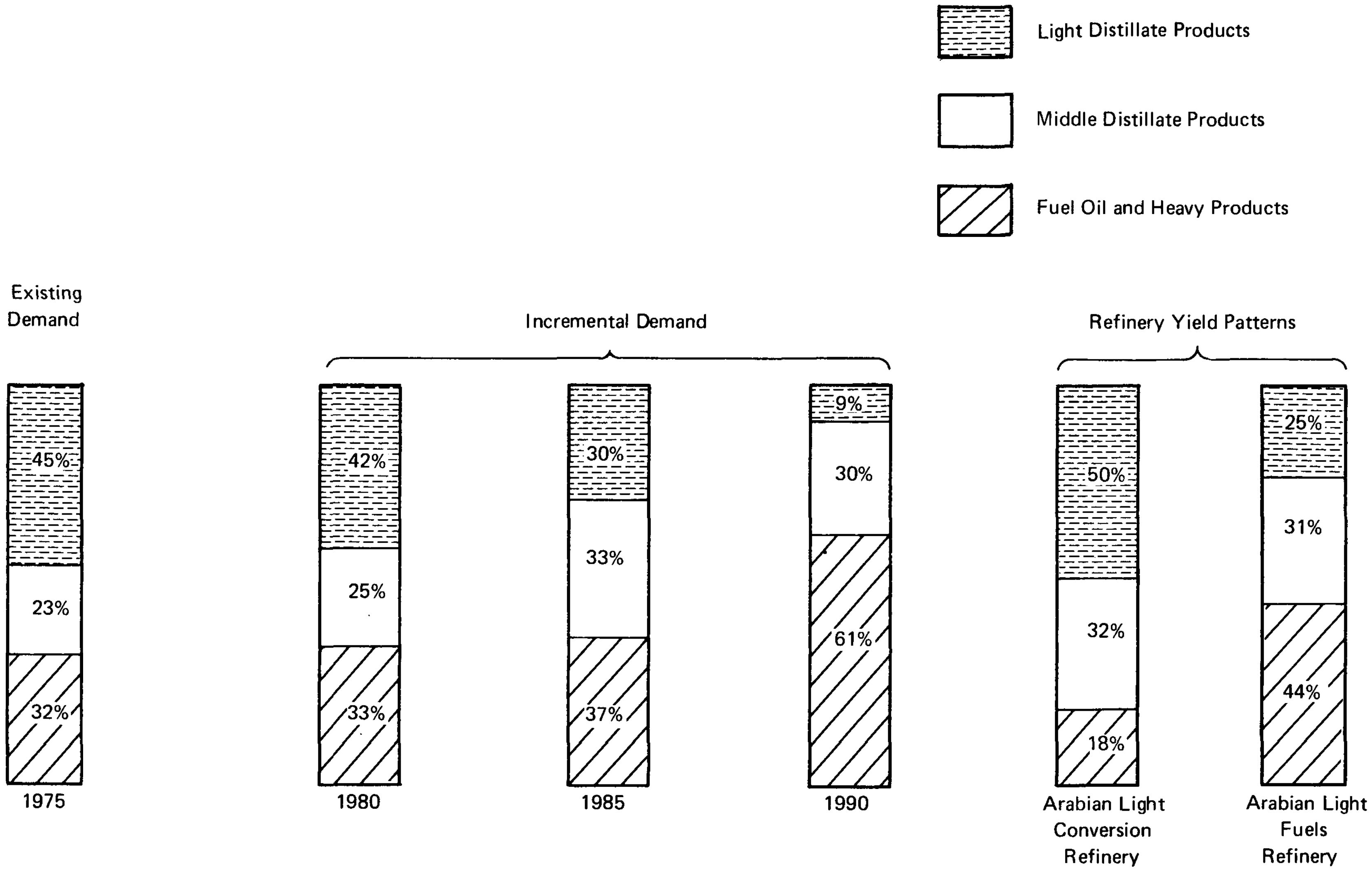

Figure A-1-14 **REFINED PRODUCTS, PRODUCTS DEMAND DISTRIBUTION AND REFINERY YIELD PATTERNS, PADD V**

For PADD's I-IV and for PADD V product price projections, sets of values for certain primary products obtained from the chosen refinery configurations were developed. These primary products were:

- unleaded gasoline
- virgin naphtha
- low sulfur gas oil
- virgin gas oil
- 0.5% S No. 6 fuel oil
- high sulfur No. 6 fuel oil

In the original development of the cost allocation model, substantial effort was directed at calibration and validation of the model by comparing its price projections to historical data. For certain products, such as naphtha, there is no established spot or contract market for which reasonable prices are available. Unleaded gasoline production economics are still in a state of flux due to the continuing lead phase-down schedule still in effect. High sulfur gas oil has no appreciable market in PADD V, and hence historic price data are questionable. High sulfur No. 6 fuel oil is primarily used for bunker fuel, which exhibits volatile price-demand fluctuations. Low sulfur fuel oils show substantial seasonal fluctuations. Therefore, comparisons of our price projections with current prices must be made with great caution. Similarly, long-term projections of product prices will not reflect such short-term variations about the major price trends.

From these basic refinery products, prices for the other products of interest (listed in Table A-1-28) were determined. Price projections were developed independently for LPG (propane). Since this is not a mainstream oil product, its price often varies independently of the conventional refinery output streams. In addition, in the U.S., the majority of propane produced is not from refineries, but from gas processing plants, where refinery economics are not relevant.

Since relatively large additional quantities of virgin kerosine can be produced by adjustment of the distillation cut point, the price of virgin kerosine was assumed to be equal to a linear interpolation of naphtha and No. 2 fuel oil. Furthermore, kerosine is a minor refinery product; future prices will be heavily dependent on local supply and demand and will have little bearing on overall refinery economics.

Prices for low sulfur distillate products were calculated to reflect the costs associated with reduction of the sulfur content of the naphtha and distillate (kerosine and gas oil) cuts from the naturally occurring levels down to 0.5 ppm (as well as 200 ppm for the distillate materials), in order that they might be acceptable as dispersed fuel cell feedstocks. Since there is little market projected for these fuels other than for fuel cells, it is likely that this price differential must be realized in order for refiners to be encouraged to produce these products.

Naphtha desulfurization was based on a single-stage naphtha treater which would reduce the naphtha sulfur content from 320 ppm to less than 1 ppm sulfur. The economics were based on a 10,000 bbl/d unit. Capital costs included offsites at 45% of process unit cost, while a 25% per year capital recovery factor was employed. The processing costs included operating costs, capital cost and hydrogen feedstock cost. The hydrogen feedstock cost in $/MCF was based on the crude cost in $/bbl divided by 6. A sample calculation for a crude cost of $12/bbl (Table A-1-26) and a hydrogen feedstock cost of $2/MCF is attached.

Distillate desulfurization was based on a two-stage distillate treater. The first stage would reduce the distillate sulfur content from 4,000-6,000 ppm to about 200 ppm. The second stage would reduce the distillate sulfur content to less than 1 ppm. The economics were developed for a 10,000 bbl/d two-stage distillate hydrotreater and were on the same basis as for naphtha desulfurization. A sample calculation is shown in Table A-1-27 for a crude cost of $12/bbl and a hydrogen feedstock cost of $2/bbl. Both operating costs and hydrogen feedstock costs are considerably higher for distillate desulfurization than naphtha desulfurization, since the degree of sulfur removal is, of course, far greater.

In the case of compliance, No. 6 fuels with sulfur levels differing from the 0.5% S level developed by the refinery desulfurization cost

TABLE A-1-26

NAPHTHA DESULFURIZATION COSTS

10,000 Bbl/d

		$ Million
Investment:		
Naphtha pre-treater, product stripper, 20% prefractionator overhead		3.8
Utilities, offsites and support facilities at 45%		1.7
Total Investment		5.5
Feed:		
°API	56	
Sulfur, ppm	320	
Distillation, °F	130-360	
Product (de-Isohexanized):		
°API	53	
Sulfur, ppm	<1	
Distillation, °F	200-360	
	FOE B/Bbl Charge	$/Bbl
Operating Costs		
Direct Fuel	0.010	0.120*
Utilities Fuel	0.008	0.096*
Other Variable Costs	-	0.044
Starting Costs	-	0.030
Capital Costs (25% capital charge = $1.375 million/yr)	-	0.420[†]
Total Operating Cost		0.710
H_2 Feedstock Cost	100 Scf/bbl	0.200[§]
TOTAL PROCESS COST (Excluding feedstock)		0.910

---

*Fuel at $12.00/FOE bbl.

[†] Based on 4 year payout and 0.90 stream factor

[§] H_2 at \$2.00/MScf

TABLE A-1-27

DISTILLATE DESULFURIZATION COSTS

10,000 Bbl/d

	Arabian Light	North Slope
Feedstock		
°API	33	33
Distillation, °F	350-650	350-650
Suflur, ppm	6,000	4,000
Product Quality		
°API	34	34
Distillation, °F	375-650	375-650
Sulfur, ppm	0.5	0.5

	$/Million
Investment	
Process	11.5
Utilities[*], 45% process	5.2
Total Investment	16.7

	FOE B/Bbl Charge	$/Bbl
Operating Costs		
Direct Fuel	0.030	0.36[†]
Utilities Fuel	0.012	0.14[†]
Other Variable Costs	-	0.06
Starting Costs	-	0.06
Capital Costs (25% capital charge= $4.175 million/yr)	-	1.27[*]
Total Operating Costs		1.89
H_2 Feedstock Costs, 300 Scf/bbl		0.60[§]
TOTAL PROCESSING COSTS		2.49

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[*] 4 year payout with 0.90 streamtreater.

[†] Fuel at $12/FOE bbl.

[§] $2.00/MCF.

equation relating high sulfur fuel oil to low sulfur fuel oil, it was assumed that all fuels of 1% S and below could be obtained by blending. Thus, the price gradient between low sulfur gas oil and 0.5% established in the product costs can be extrapolated backwards to 1% S No. 6 fuel. Compliance No. 6 fuel sulfur levels assumed in this analysis were:

Los Angeles/Hartford	0.5% S
Dallas	0.7% S
Columbus	1.0% S

High sulfur vacuum bottoms (generally similar to asphalt) has no established market at present. Indeed, even the asphalt market is characterized by very few suppliers and unusual price relationships. Nevertheless, vacuum bottoms, on the margin, is typically marketed by blending with distillates for sale as fuel oil. For the high sulfur vacuum bottoms analyzed in this study, this outlet would typically be high sulfur bunker fuel oil. Hence, the vacuum bottoms prices developed in the present study were based upon blending with high sulfur gas oil to produce bunkers. Since these latter prices had already been projected, it was possible to back-calculate a consistent vacuum bottoms price at which the refiner would be willing to market the bottoms as a fuel cell feedstock rather than blending to bunker fuel oil.

Capital charges relevant to overall refinery margins were based upon the supply/demand curves of Figures A-1-11 and A-1-13. These capital charges specifically were drawn from evaluations of historic refinery margins under conditions of excess and tight supply capacity; evaluations of the relative investments required for debottlenecking, major refinery expansions, and new grassroots refineries; and testimony by refining company officers in the Senate hearings on multinational companies on target rates of return on assets.

In addition, supply/demand evaluations for individual petroleum products were investigated to allow specification of appropriate capital charges on downstream processing equipment. Figures A-1-15 through A-1-18 typify the results of such analyses for motor gasoline production. The major processing units used to produce the gasoline in the United States are

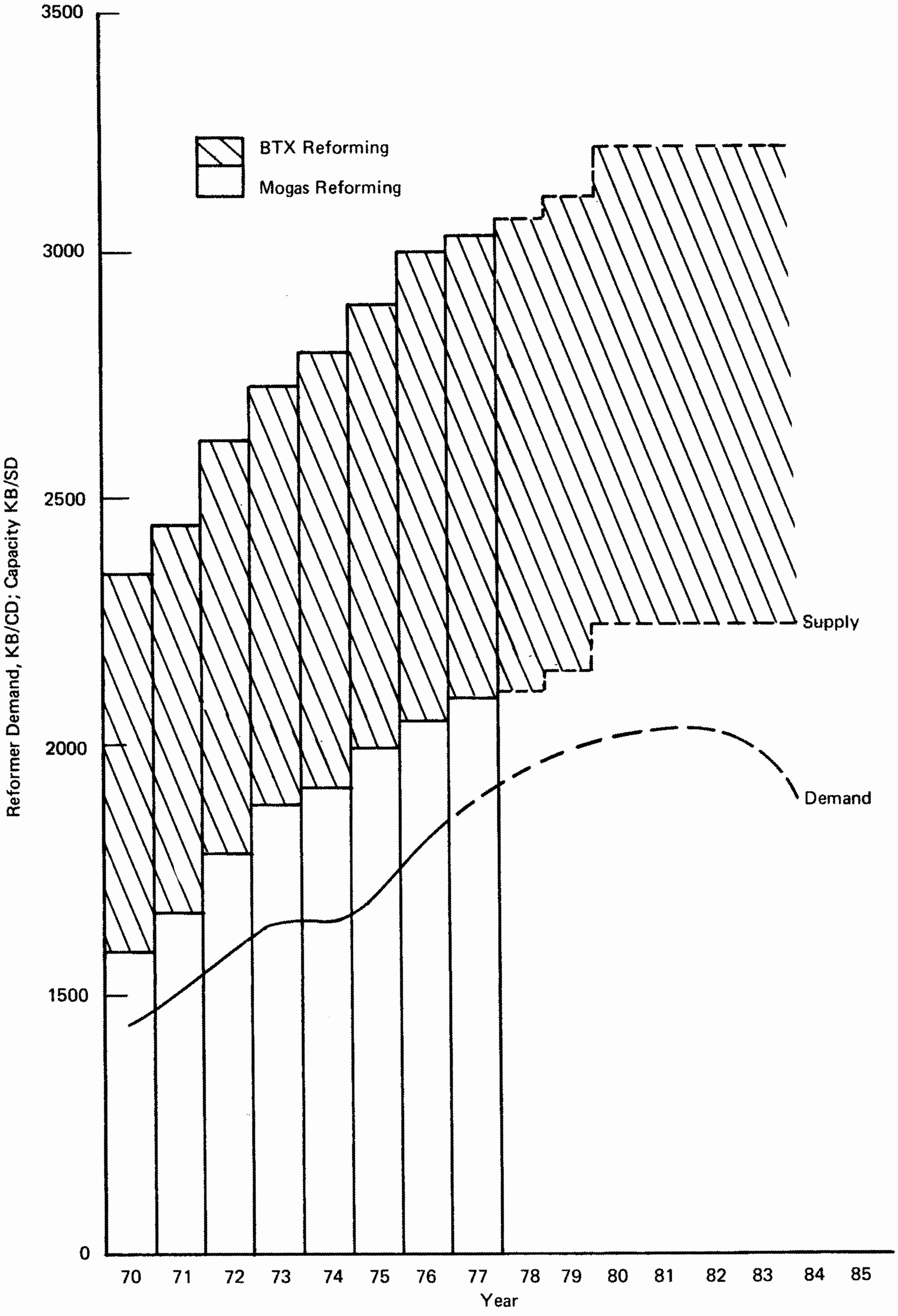

Figure A-1-15 PADDS I-IV REFORMER SUPPLY/DEMAND

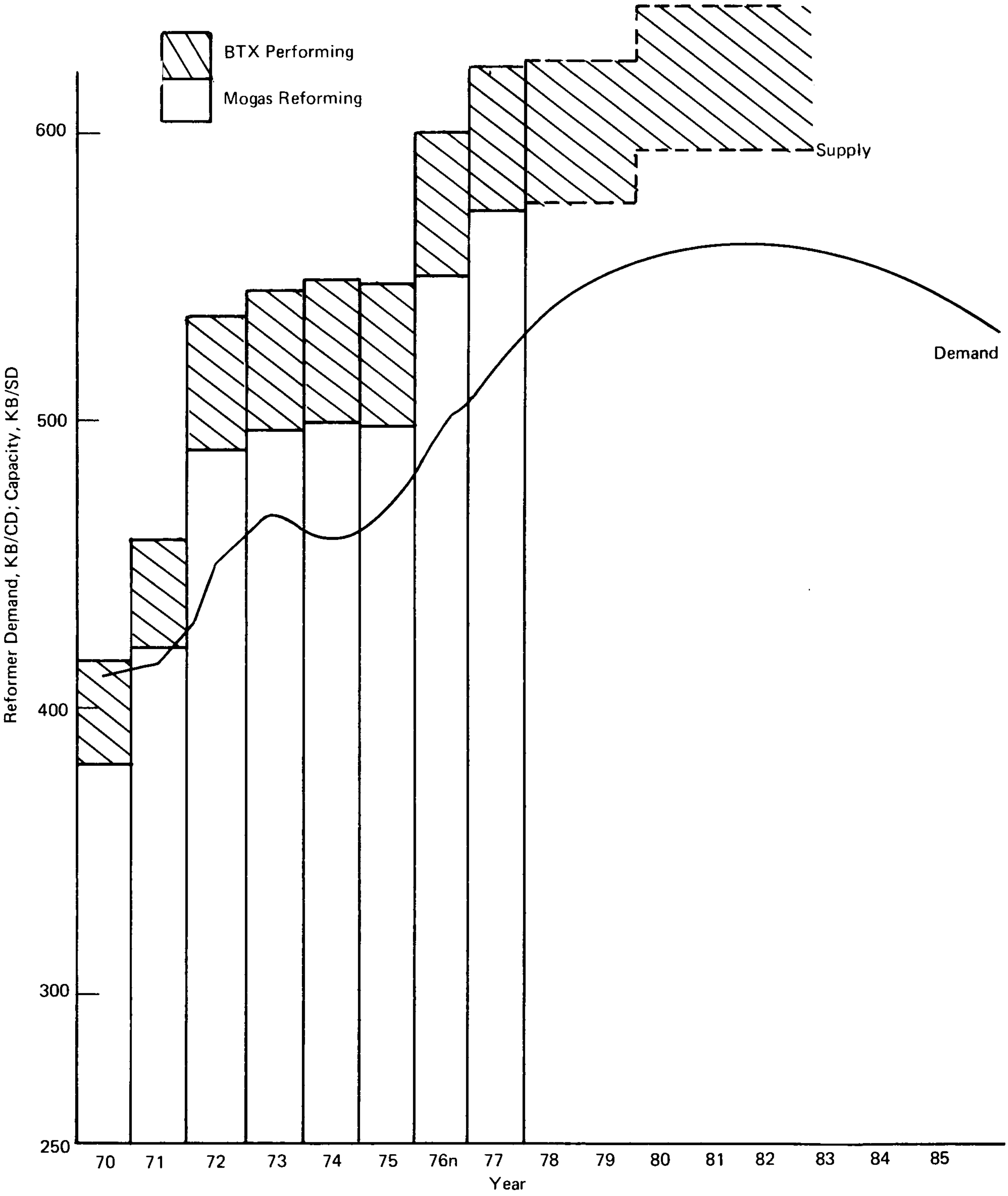

Figure A-1-16 PADD V REFORMER SUPPLY/DEMAND

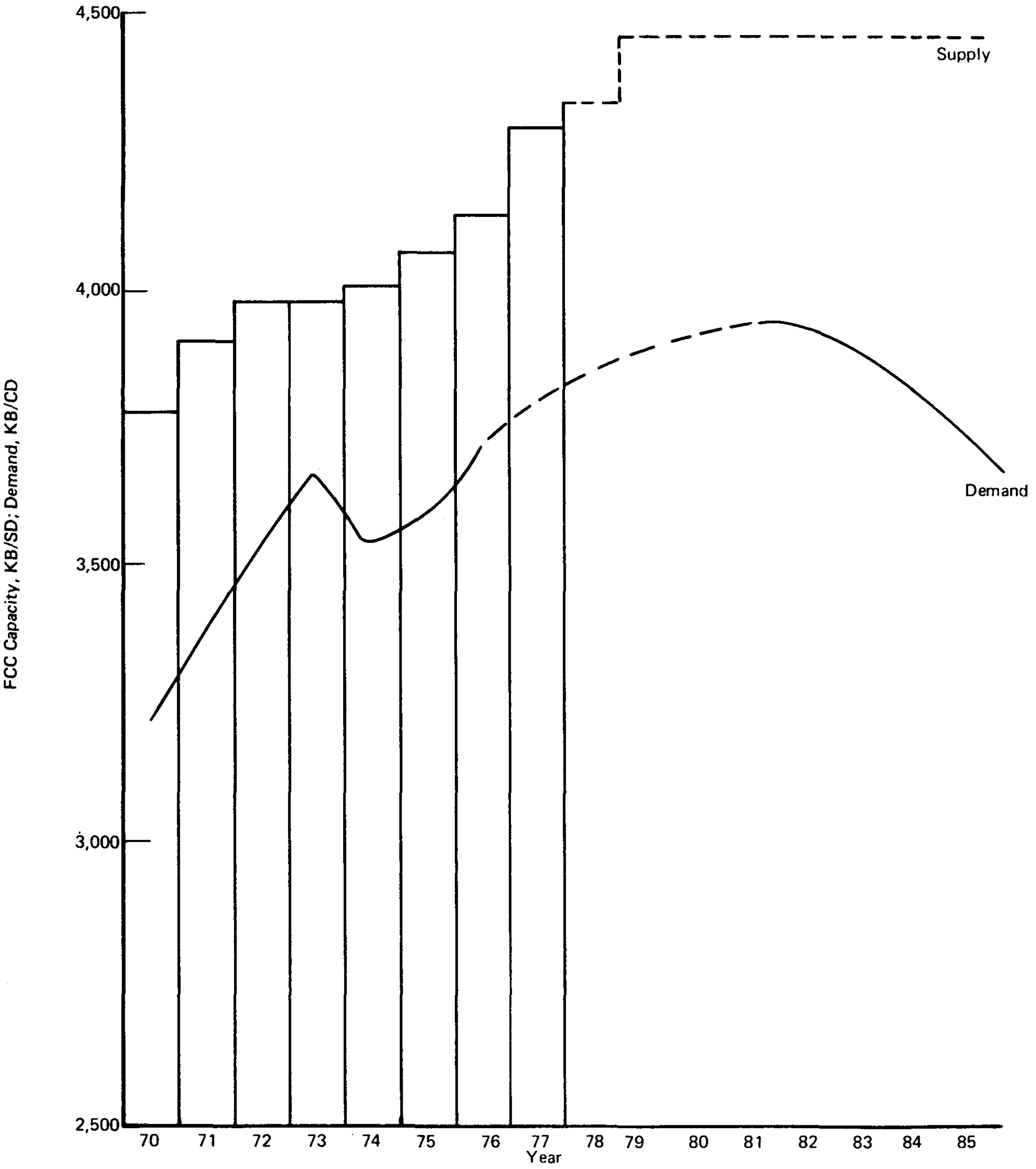

Figure A-1-17 PADDS I—IV FCC UNIT SUPPLY/DEMAND

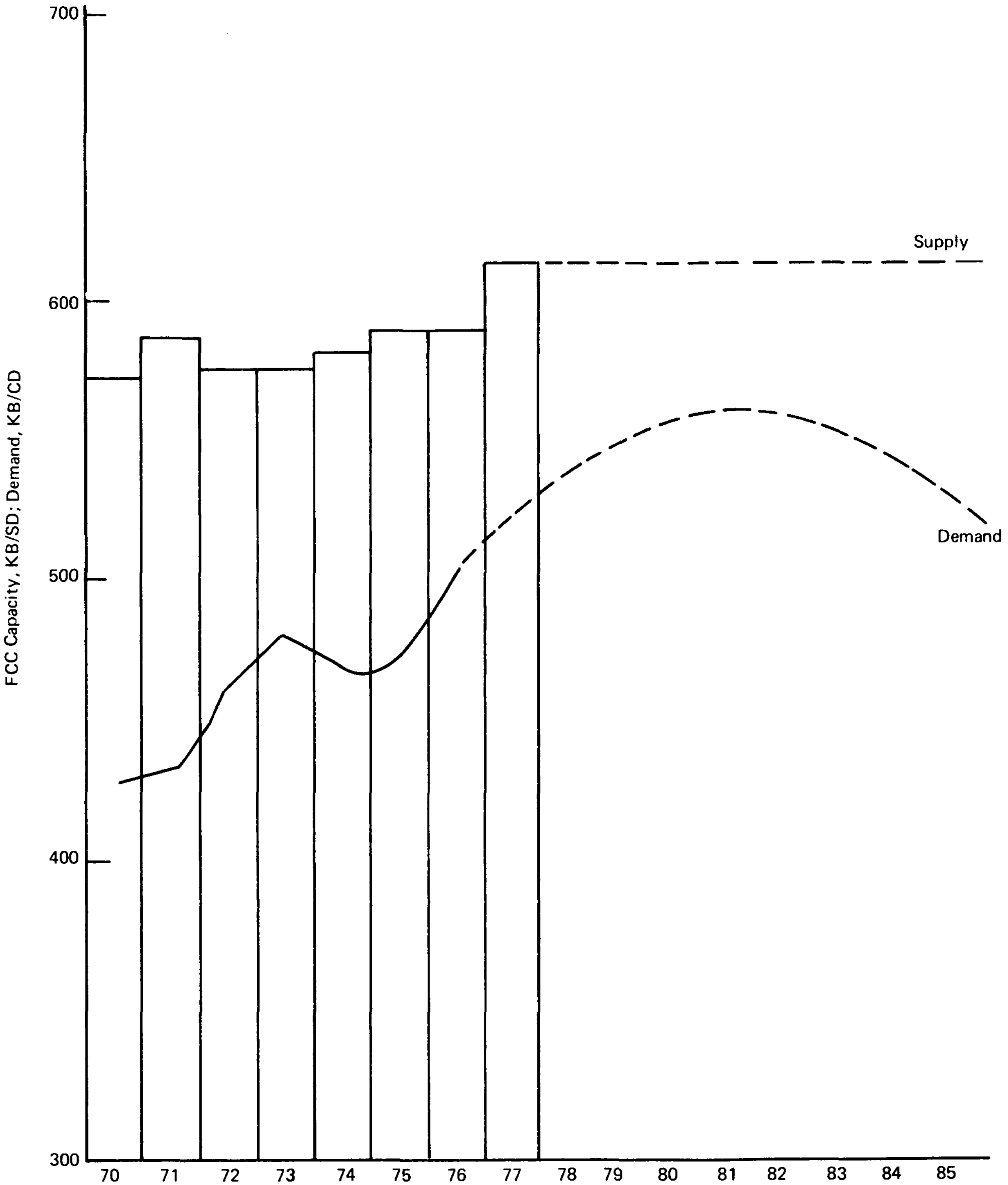

Figure A-1-18 PADD V FCC UNIT SUPPLY/DEMAND catalytic reforming units and fluid catalytic cracking (FCC) units. Refinery catalytic reforming units, in addition, are used to produce aromatics (BTX) for sale in the petrochemical market. An analysis of the capacity presently used and projected to be required for BTX production is shown in the cross-hatched portion of Figures A-1-15 and A-1-16, and has a secondary influence on gasoline margins. The historic and projected demands on the motor gasoline reformers are compared to rated capacity in the lower portion of these figures (available or useful capacity will be about 90% of the indicated capacity). The conclusion was drawn from these figures that reforming margins will increase between 1977 and 1981, and then ease somewhat. Reformer severity is expected to increase to around 100 by the early 1980's due to the impact of the lead phase-down program. Examinations of the supply and demand for the other major source of motor gasoline, the FCC unit, is portrayed in Figures A-1-17 and A-1-18. A similar conclusion may be drawn; an upward pressure on gasoline margins should exist until 1981, followed by an easing of margins thereafter. These generalizations are quantified by selection of appropriate capital charges on reformers and FCC conversion units in a fashion analogous to that described for the overall refinery. As discussed in the following section, these trends will have long-term implications for naphtha and gasoline prices relative to other distillate prices.

Similar analyses on other refinery processing units were conducted for PADD's I-IV and PADD V individually. For example, the price of low sulfur No. 6 fuel oil is dependent not only on desulfurization costs but also upon the price of low sulfur crude oil. On the other hand, most exporting nations price their crude oil to account for differential refinery processing costs (including desulfurization) between crude as well as differential transportation costs. In theory, then, it can be argued that over the long run, the price of low sulfur fuel oil in Los Angeles derived from low sulfur Indonesian crude would be the same as that derived from desulfurization of fuel oil from Arabian light crude oil. Historically, medium and short term distortions from this equilibrium trend have taken place. For example, Japan now has a substantial excess of desulfurization capacity, probably as a result of national policy decisions. Pacific basin sulfur balances suggest this trend will persist throughout the study period, resulting in lower high sulfur/low sulfur fuel oil margins than would be anticipated based upon full recovery of desulfurization process costs. In contrast, Atlantic basin sulfur balances do not exhibit such long-term dislocations. Therefore, low sulfur/high sulfur fuel oil price differentials for the West Coast will be lower than expected from the differential desulfurization economics between the West Coast crude (Alaskan North Slope) aNd the Gulf Coast crude (Arabian Light). In addition, as discussed in Section 1.2.3, the refinery gate crude oil costs differ between these two locations, further influencing these differential prices. The price projections of Section 1.2.4.4 take into account all of these factors--crude oil price, relative processing costs, and anticipated capital charges. Indeed, a primary advantage of price projections from such models is the ability to separately include quantitative expectations of each of these factors in projecting their composite effect on product prices.

1.2.4.4 Projected Ex-Refinery Products Prices

Using the methodology outlined in Sections 1.2.4.1 and 1.2.4.2, including:

- timing of the need for new refining capacity, hence the point at which overall refining margins become fully incremental;
- shape of the incremental demand barrel, influencing the type of incremental refinery construction;
- impact of the changing pattern of demand on quality improvement processes, particularly in terms of gasoline lead levels (affecting reformer utilization and, hence, the gasoline/naphtha price differential) and heavy fuel oil sulfur contents, affecting desulfurizer employment and the low sulfur/high sulfur fuel oil margin.

The ex-refinery prices shown in Tables A-1-28 and A-1-29 were determined for the Gulf Coast and West Coast refining locations.

In order to relate the price projections to current refining trends and products price quotations, the model price projections in 1980 for the

TABLE A-1-28

PROJECTED EX-REFINERY PRODUCTS PRICES - GULF COAST

(1975 $/Bbl)

	1980		1985		1990	
CRUDE PRICE SCENARIO	CONTROLS	DECONTROL	CONTROLS	DECONTROL	CONTROLS	DECONTROL
Refinery Gate Crude Price:	11.17	12.30	13.51	18.04	14.63	19.50
Products						
Unleaded Gasoline	16.90	18.08	19.98	24.74	20.69	25.80
0.5 ppm Naphtha	14.90	16.13	18.26	22.88	19.03	23.98
Virgin Naphtha	14.03	15.21	17.30	21.76	18.03	22.81
0.5 ppm Kerosine	15.75	17.10	19.62	24.63	20.91	26.29
200 ppm Kerosine	14.39	15.84	18.64	23.63	20.35	25.72
Virgin Kerosine	14.13	15.38	17.82	22.55	19.04	24.12
0.5 ppm No. 2 Fuel	16.60	18.06	20.97	26.38	22.79	28.60
200 ppm No. 2 Fuel	14.52	15.84	18.64	23.63	20.35	25.72
Virgin Gas Oil	14.22	15.54	18.34	23.33	20.05	25.42
0.5% S No. 6 Fuel	12.52	13.84	16.64	21.63	18.35	23.72
0.7% S No. 6 Fuel	12.23	13.54	16.29	21.25	17.96	23.30
1.0% S No. 6 Fuel	11.80	13.09	15.77	20.68	17.39	22.66
High Sulfur No. 6 Fuel	9.62	10.65	13.35	17.82	14.76	19.60
High Sulfur Vacuum Bottoms	8.47	9.68	11.88	16.44	13.10	18.01

Source: A.D. Little Estimates

TABLE A-1-29

PROJECTED EX-REFINERY PRODUCTS PRICES - WEST COAST

(1975 $/Bbl)

	1980		1985		1990	
CRUDE PRICE SCENARIO	CONTROLS	DECONTROL	CONTROLS	DECONTROL	CONTROLS	DECONTROL
Refinery Gate Crude Price:	11.10	11.82	13.42	17.42	14.52	18.64
Products						
Unleaded Gasoline	17.30	18.07	19.59	23.91	20.13	24.52
0.5 ppm Naphtha	15.75	16.51	18.30	22.55	18.88	23.18
Virgin Naphtha	14.87	15.60	17.34	21.45	17.88	22.04
0.5 ppm Kerosine	16.91	17.71	19.90	24.38	21.10	25.63
200 ppm Kerosine	15.53	16.28	18.37	22.59	19.49	23.77
Virgin Kerosine	15.27	16.02	18.11	22.33	19.23	23.51
0.5 ppm No. 2 Fuel	18.07	18.91	21.50	26.20	23.31	28.07
200 ppm No. 2 Fuel	15.96	16.73	19.18	23.50	20.88	25.27
Virgin Gas Oil	15.66	16.43	18.88	23.20	20.58	24.97
0.5% S No. 6 Fuel	12.85	13.61	16.39	20.68	18.88	23.27
1.5% S No. 6 Fuel	10.60	11.34	13.23	17.39	14.42	18.62
Medium Sulfur Vacuum Bottoms	9.34	10.07	11.82	15.94	12.88	17.03

Source: A.D. Little Estimates

U.S. Gulf Coast were compared with actual prices quoted in the trade press. In order to "normalize" the calculations, all prices were converted to 1975 dollars. This comparison is shown in Table A-1-30.

A number of important issues emerge from an analysis of Table A-1-30:

- Since overall refinery margins in 1976 and 1977 were depressed, the weighted average product realization ex-refinery is higher in 1980 than today.

- Naphtha prices are expected to continue to remain below distillate prices. Currently, there is a significant naphtha surplus in the Caribbean and the margin between naphtha and gas oil has risen to over $2/bbl (in 1975 terms). Longer term it is expected that this margin will be reduced, but the outlook for naphtha/gasoline demand does not indicate long-term strength in light distillate prices, resulting always in an expectation of a naphtha price which is below that of gas oil.

- Due to the current surplus of direct and indirect fuel oil desulfurization capacity in offshore refineries oriented towards the U.S. low sulfur No. 6 fuel market, margins between 0.7% S and high sulfur fuel oil are now well below full incremental cost. It is expected this situation will reverse as growing fuel oil demand, particularly for low sulfur fuels in the wake of declining domestic natural gas production, creates the need for further desulfurization capacity. This trend will be accentuated by the declining production of sweet U.S. crude oil which must be replaced by either sour North Slope or Middle East crudes. The U.S., if it wishes to avoid converting current onshore refineries which cannot handle sour crudes, may be forced into bidding up the price for the pool of low sulfur Algerian, Libyan, Nigerian and Indonesian crudes in order to avoid such expensive hardware conversion costs. In either of these cases, the cost of low sulfur fuel oil vis-a-vis high sulfur fuel oil would increase.

- Fuel oil prices in general will rise as the heavy incremental demand barrel permits a greater allocation of refining costs onto the fuel oil fraction. Indeed by 1990 projections show that high sulfur fuel oil prices will be very close to the input costs of crude oil into the refinery. Historically high sulfur fuel oil prices have been well below crude cost. A major influence on the future level of fuel oil prices will be U.S. coal and nuclear electricity production. If production of these alternative energy forms (which primarily substitute for oil products at the

TABLE A-1-30

COMPARISON OF HISTORICAL PRICE QUOTATIONS WITH 1980 VALUES OBTAINED FROM PRICE MODEL

U.S.Gulf Coast Basis

($/Bbl)

	U.S. Gulf Coast[†] Quotations				Caribbean[§] Quotations				1980 Prices
	1976	1977	1976	1977	1976	1977	1976	1977	With Controls
	Actual	Actual	1975 Dollars*		Actual	Actual	1975 Dollars*		1975 Dollars*
Naphtha	-	-	-	-	14.11	12.86	13.31	11.48	14.03
Jet Kerosine	13.23	14.65	12.48	1308	-	-	-	-	14.13[¶]
Gas Oil	12.65	13.99	11.93	12.49	14.71	15.55	13.88	13.88	14.22
No. 6 Fuel									
0.7%	-	11.82	-	10.55	-	-	-	-	12.23
1.0%	10.37	11.47	9.78	10.24	11.66	12.15	11.00	10.85	11.80
3.0%	9.05	10.87	8.54	9.71	10.89	11.82	10.27	10.55	9.62

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*1976 values deflated by 1.06 to 1975 equivalent; 1977 values deflated by 1.12 to 1975 equivalent.

[†] Gulf Coast prices are somewhat lower than Caribbean prices due to the impact of price-controlled domestic crude on average refinery gate crude prices.

[§] Caribbean prices reflect full international crude prices, and include product freight to the U.S. Gulf.

[¶] Virgin kerosine (jet kerosine value would be somewhat higher).

Sources: 1976 Prices - Platt's Oilgram Price Service - September 20, 1976.
1977 Prices - Platt's Oilgram Price Service - September 20, 1977.

heavy end of the barrel) is higher than that projected in the supply/demand analysis, it is expected that the overall level of fuel oil prices would be somewhat lower than projected here (particularly for 1985 and 1990).

1.2.5 Distribution and Marketing Costs

In this section the cost of various petroleum logistical links from the refinery gate to the fill pipe of a utility company storage tank are determined. These transportation and marketing costs are added to the ex-refinery product cost in order to derive the price of energy delivered to the utility. For illustrative purposes, the following four sample utilities were chosen to reflect both regional and logistical differences:

Sample Utility Location	Refinery Source
Hartford, Connecticut	Gulf Coast
Columbus, Ohio	Gulf Coast
Dallas, Texas	Gulf Coast
Orange County, California	Los Angeles

These sample utilities are the same locations chosen in the previous A. D. Little study for EPRI (RP-318) in October 1975. This analysis not only updates the logistical costs but reflects changes from the prior work in refinery sources and modes of transportation as well as a wider variety of fuel alternatives. From a logistic viewpoint, the following distinct classes of fuel cell fuels were considered:

Product Type	Product Group	Products Included
Clean	Commodity distillates	No. 2 oil, kerosine, naphtha, gasoline
Clean	Specialty distillates	0.5 ppm No. 2 oil, kerosine, and naphtha
Clean	Light gases	LPG
Clean	Specialty chemicals	Methanol
Dirty	Residual oil	High sulfur No. 6 oil, low sulfur No. 6 oil
Dirty	Vacuum bottoms	Asphalt

Each of the above product groups would have distinctive logistical networks and associated costs, although the logistical cost for individual products within each group are assumed to be equal. All costs are expressed in the 1975 dollars in order to be consistent with A. D. Little's previous work for EPRI, and are summarized in Table A-1-31.

The annual volume of clean fuel products transported and stored is assumed to be sufficient for 10 dispersed 26 mw fuel cells (3,000-5,000 bbl/d depending upon the calorific value of the fuel). Dirty products would be used in a central conversion fuel cell facility with a generating capacity of approximately 1200 mw, with a dirty products fuel demand exceeding 36,000 bbl/d.

In order to focus upon logistical and regional cost differentials, a common overhead allocation of $1.01/bbl was applied to all products in all locations. This cost will capture sales and general overhead expenses as well as a return on investment for the marketing profit center of a major oil company or an independent marketer. A breakdown of these other marketing expenses (OME) is shown in Table A-1-32.

1.2.5.1 Hartford Area Sample

Logistical linkages for fuel cell fuels in Hartford are shown in Figures A-1-19 and A-1-20 for clean and dirty petroleum products. Except for LPG, all products would be shipped by tanker from the U.S. Gulf Coast to a marine terminal in New Haven, Connecticut. Clean products would require the equivalent of approximately five dedicated 35,000 dwt tankers per year. Commodity clean products movements would not require special handling and could be accommodated in existing clean products tanker logistics. However, specialty products would require higher cost segregated transport and storage to avoid contamination. A central conversion facility using residual fuel oil will require over 10 times as much volume as the dispersed option which would necessitate the receipt of a dirty tanker cargo every week.

Primary terminals provide the initial direct storage link between refinery sources and the consumer. In both the dispersed and the

TABLE A-1-31

SUMMARY OF PRODUCTS DISTRIBUTION AND MARKETING COSTS

(1975 $/Bbl)

TO:		Hartford	Columbus	Dallas	Los Angeles Area
FROM:		U.S. Gulf	U.S. Gulf	U.S. Gulf	Los Angeles
Fuel Cell Option	Product Group				
Dispersed	Commodity-Clean	2.65	2.65	1.63	1.37
Dispersed	Specialty-Clean	2.88	3.38	2.52	1.49
Dispersed	Methanol	3.45	3.70	2.91	1.66
Dispersed	LPG	4.60	3.45	2.49	1.48
Central	Residual Fuels	3.22	3.75	2.69	1.58
Central	Asphalt	3.90	4.47	3.14	1.47

Source: A.D. Little Estimates

TABLE A-1-32

OTHER MARKETING EXPENSES (OME)*

Functional Expense	Cost Element	Allocated Costs (1975 $/Bbl)
Labor*	Corporate and sales salaries, benefits and employee expenses	0.120
Services/Products Used*	Corporate, legal expenses, fuels, utilities, etc.	0.039
Miscellaneous Expenses*	Sales promotion, insurance, etc.	0.003
Capital Charge	Return on working capital and fixed assets	0.848
TOTAL OME		1.010

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*For sales and corporate overhead accounts only. Other costs are either directly purchased services or charged to other functional areas (i.e., terminals, delivery fleets, etc.).

Region - Northeast
Sample City - Hartford, Conn.

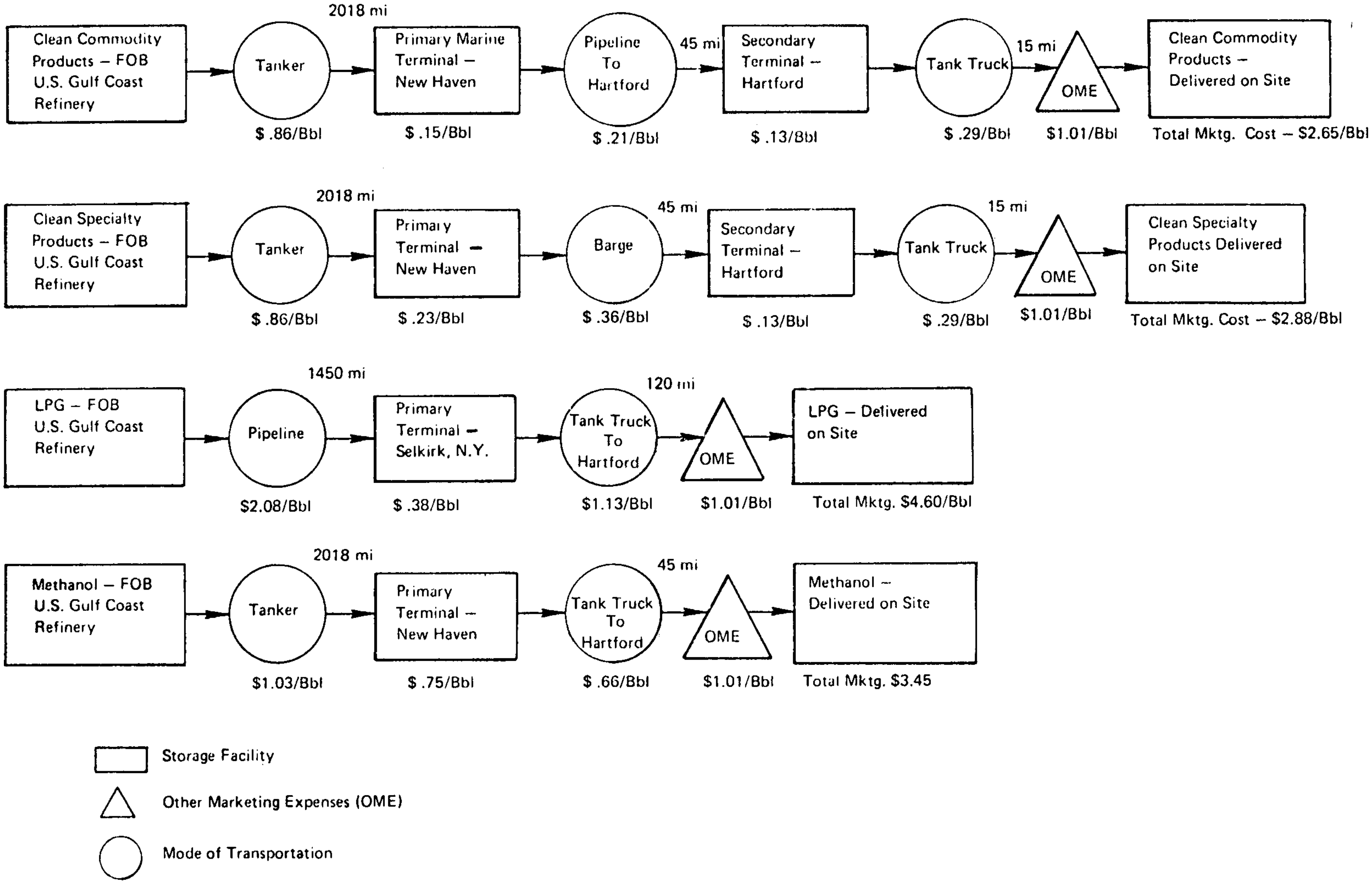

FIGURE A-1-19 CLEAN PRODUCTS - 10 DISPERSED FUEL CELLS LOGISTICAL LINKAGES AND COSTS

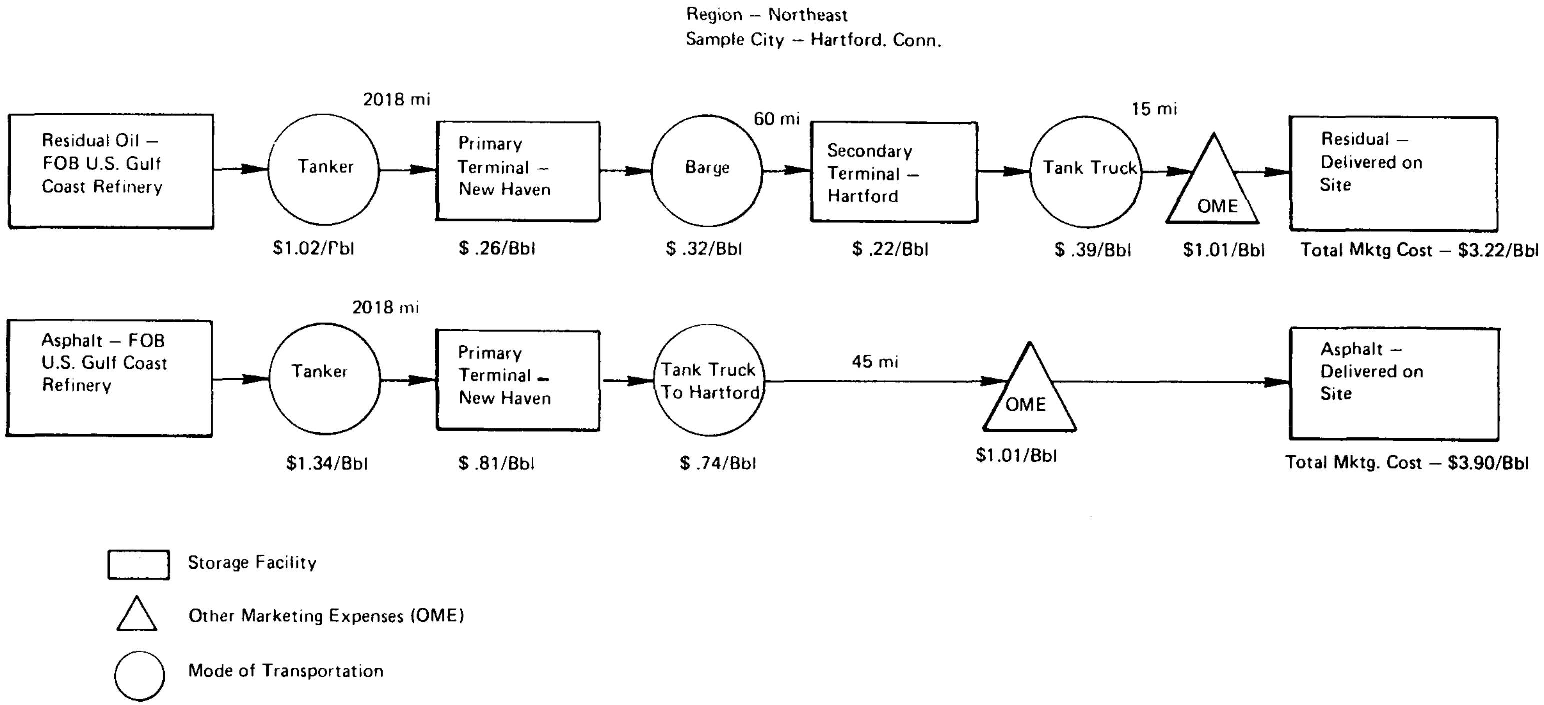

FIGURE A-1-20 DIRTY PRODUCTS - CENTRAL CONVERSION OPTION LOGISTICAL LINKAGES AND COSTS

centralized fuel alternatives, the consumption of fuel cell fuels is sufficiently large to necessitate additional storage in New Haven. In the case of clean products, an estimated 200,000 Bbl of additional storage would be required (two 100,000 barrel tanks). A small premium would be charged for specialty clean products since dedicated tanks and lines would limit the terminal operator's handling flexibility (specialty fuels cannot be mixed with products going to other customers). If dirty products were used at a central conversion facility approximately 2 million Bbl of new storage would be required, which is roughly equal to the current level of residual oil storage in New Haven. Storage tanks for dirty products must be insulated and heated which results in higher terminal throughput costs than for clean products. Methanol storage is also more expensive than for clean products since stringent quality control handling is necessary and methanol is considered to be a chemical which commands a higher rate schedule than petroleum products.

The clean products terminal throughput fee of $0.15/Bbl is equal to the current market price and does not reflect the fully allocated cost of new storage. However, the throughput fee for all other product groups approaches a full allocation of the storage and handling cost at a primary terminal.

The distance from New Haven to Hartford, Connecticut is approximately 45 miles. The most economical mode of transportation for this link is the Jet Line pipeline which moves approximately 40,000 bbl/d from New Haven to secondary bulk plants in Hartford and Springfield, Massachusetts. The pipeline tariff for distillates from New Haven and Hartford is $0.14/Bbl plus a line loss factor of 0.5% of the throughput (equivalent to approximately $0.07/Bbl). Specialty clean products could be moved by pipeline to Hartford, but contamination would be a problem. Sulfur contents of these specialty fuels are 0.5 ppm while commodity distillate fuels contain from 200 to 5000 ppm of sulfur. Residual clingage of higher sulfur fuels previously shipped through the pipeline potentially could contaminate the specialty low sulfur fuels. From a conservative logistical perspective, low sulfur clean products are assumed to be moved to New Haven by dedicated barges, which is also the most economical transportation mode for residual oil. Methanol and asphalt sent to Hartford will continue to be direct trucking shipments from a New Haven primary terminal to the customer's fill pipe since secondary distribution facilities for these products are not available in Hartford.

In the Hartford area, there are no public terminalling facilities. However, there are a few terminals formerly owned by majors which are now run by local resellers, which have sufficient storage or expansion potential for fuel cell fuels. Secondary bulk plant costs in Hartford are slightly lower than the primary terminal costs in New Haven, despite the significantly lower throughput volumes, since real estate costs and taxes are much lower in Hartford than New Haven.

The dispersed fuel cell facilities are assumed to be located within a 15-mile radius of Hartford and receive full load deliveries (7500 gallon tank trucks). For 10 dispersed 26 mw fuel cells, approximately 18 truck deliveries per day would be required, or roughly 2 daily loads per fuel cell location. If each truck made 2 loads per day on a double-shifted basis, then a fleet of three or four tractor/trailers would be sufficient. Since this level of volume would justify dedicated rigs, there would be no truck delivery cost differential between commodity and specialty clean products (except for methanol which is based upon a chemical rate). A fleet of approximately 28 dedicated double-shifted trucks would be required to move residual oil from a Hartford secondary bulk plant to a centralized conversion facility in the local area. This centralized plant would receive over 180 truck deliveries per day of either residual oil of asphalt. Since asphalt would be coming directly by truck from New Haven, a fleet of 38 dedicated vehicles would be required.

LPG deliveries to Hartford have a unique logistical pattern. New Haven does not have any LPG marine facilities. The majority of propane delivered to Hartford is supplies by truck from the LPG caverns in Selkirk, New York. Gulf Coast LPG is shipped to this point by the Texas Eastern Pipeline for a total cost of \$2.09/Bbl (which includes a 3%

line loss). LPG would then be directly shipped to dispersed fuel cells in Hartford in 10,000 gallon truck transports.

1.2.5.2 Columbus Area Sample

A large proportion of the Ohio demand for petroleum products is currently supplied by local PADD II refineries. However, it has been assumed that Gulf Coast refineries would be the source of the incremental petroleum fuel consumed by utility fuel cells since environmental constraints will force most new refining capacity into PADD III. Commodity clean product fuels and LPG would be moved to southwestern Ohio by the 20" Texas Eastern pipeline (see Figure A-1-21). Commodity clean products would be taken off at the Texas Eastern terminal in Lebanon, Ohio and then transshipped into the Columbus area. The LPG consumed in the Columbus area would be shipped to the Texas Eastern storage caverns in Todhunter, Ohio which is 150 miles southwest of Columbus. At the present time, winter shipments on the Texas Eastern pipeline are quite tight and space is allocated between the various historical shippers. However, Texas Eastern plans to loop this line and double the current 200,000 Bbl/d capacity by 1980. This should permit uninterrupted shipment of clean products and LPG via the pipeline from Gulf Coast refineries.

The other potential fuel cell fuels would be shipped from the Gulf Coast by barge to terminals in the Cincinnati areas (Figure A-1-22). Typically, barge movements consists of 4 to 5 barges each with capacity of 40,000 Bbl. The Cincinnati River terminals would receive barges of specialty products for fuel cells (low sulfur oils and methanol) every seven weeks. This movement would require approximately 200,000 Bbl of new storage. With ten times the consumption of the dispersed fuel cell option, dirty products used in a central conversion facility would require approximately 1.5 million Bbl of storage. This capacity would provide approximately 20 days' supply which should compensate for a periodic disruption of the river traffic due to bad weather. Currently, dirty product storage in Cincinnati is relatively tight and could not

Region — East Central
Sample City — Columbus, Ohio

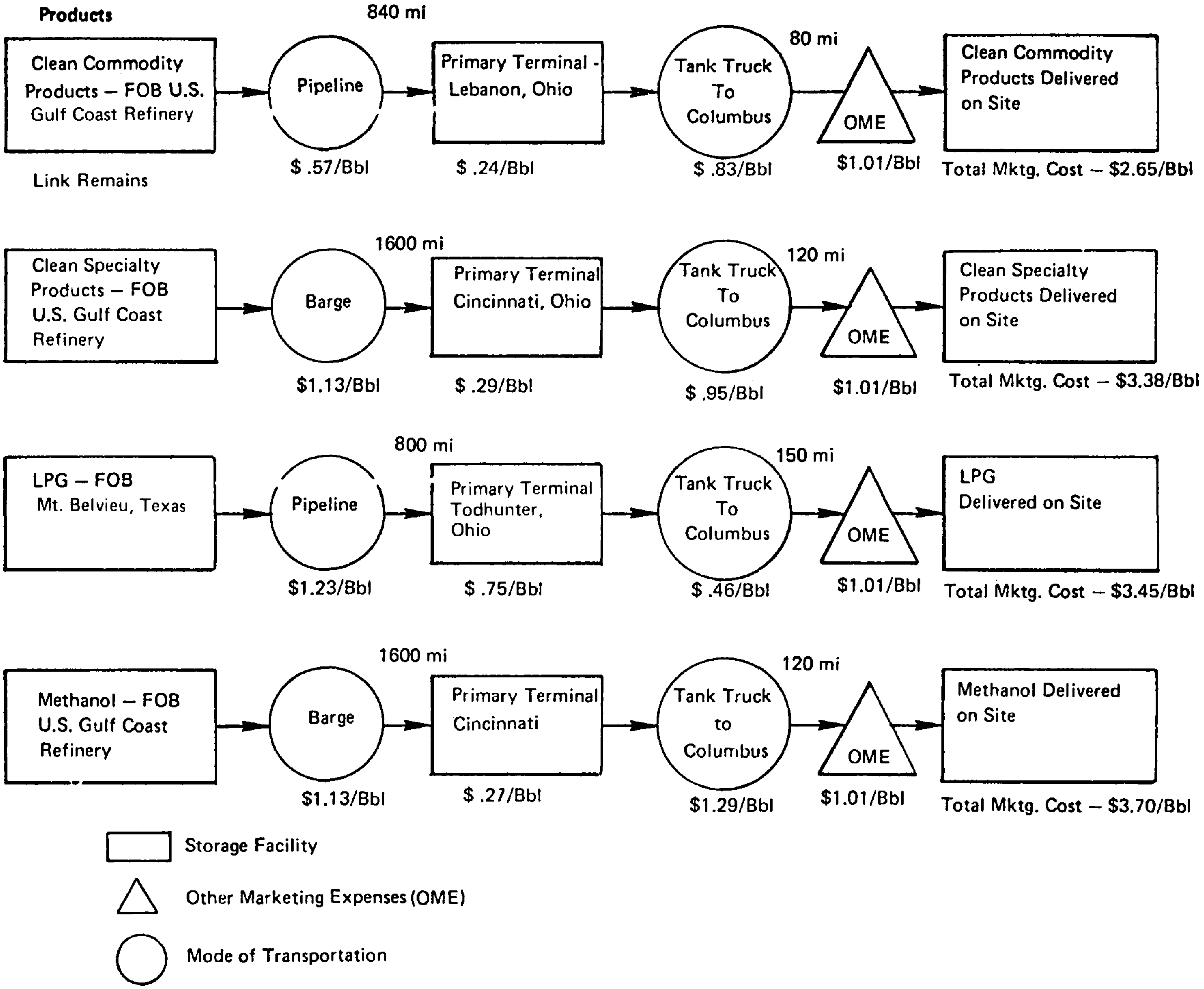

FIGURE A-1-21 CLEAN PRODUCTS — 10 DISPERSED FUEL CELLS LOGISTICAL LINKAGES AND COSTS

Region — East Central
Sample City — Columbus, Ohio

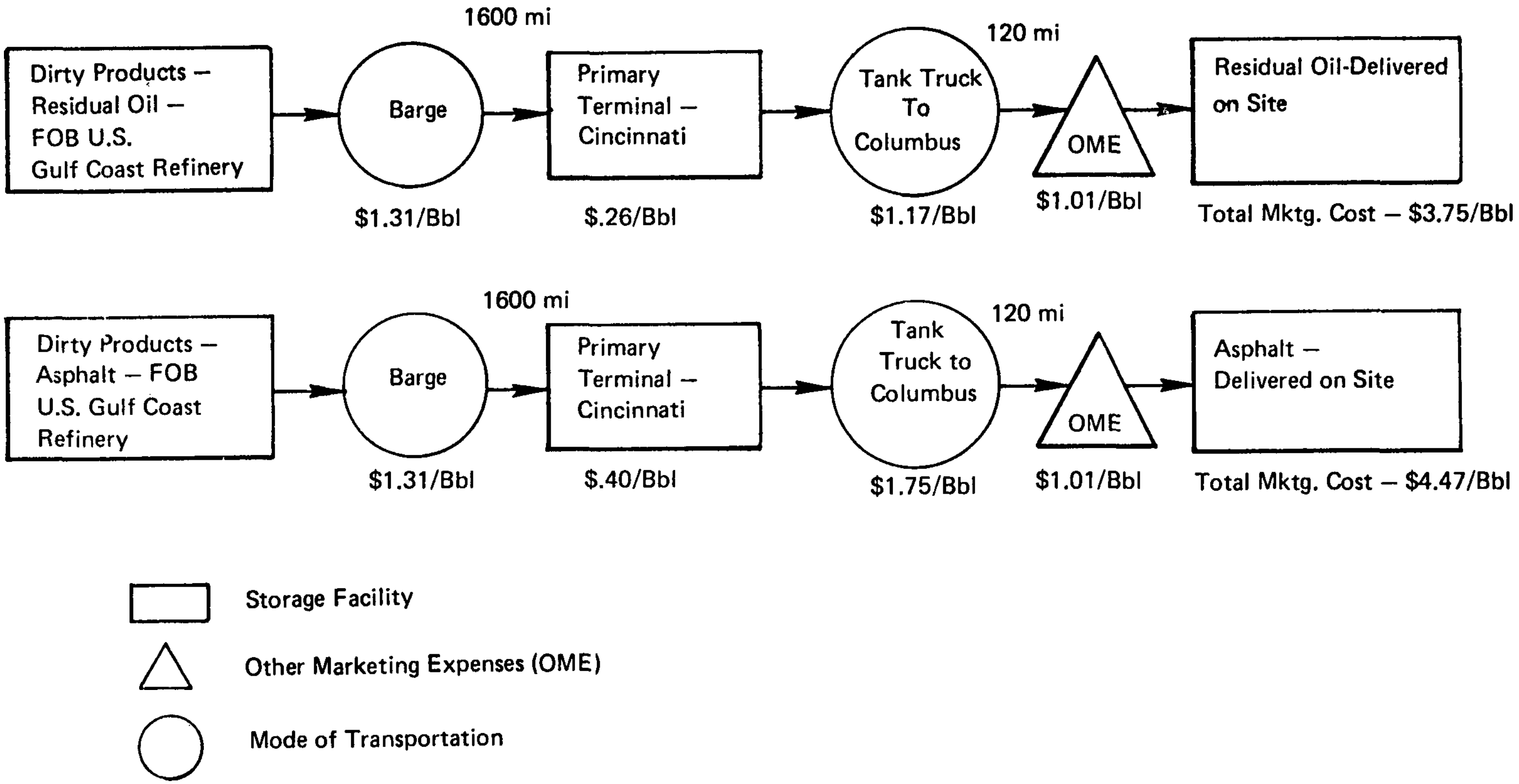

FIGURE A-1-22 DIRTY PRODUCTS - CENTRAL CONVERSION OPTION LOGISTICAL LINKAGES AND COSTS

handle the large incremental throughput of fuel cells.

As in New England, commodity clean products would have the lowest terminal throughput tariff with the following storage premiums allocated to other product groups:

Product Group	Reason for Throughput Premium
Specialty clean products	Segregated non-fungible handling costs
Methanol	Billed as a chemical product
Dirty products	Higher heating costs
LPG	Higher cost operation than liquid terminal facilities

Clean products in primary terminals in Lebanon would then be directly trucked 80 miles to the utility fuel cells. Dirty and specialty products would have to travel 120 miles from Cincinnati, while the LPG terminal is even further away. Truck delivery costs in Ohio are rigorously controlled by the state on a variable cost basis by product group. The LPG unit costs are lower since a greater volume can be carried with existing state truck regulations (10,000 gallons for LPG as against 6-8,000 gallons for other products).

Based on double-shift operations, clean product deliveries to Columbus from Lebanon would require a fleet of approximately 7 tractor trailers. Each of the 10 dispersed fuel cells would receive 2 loads per day. The added volume and distance for LPG shipments would increase the fleet requirement to 9 LPG rigs. Dirty products consumed at a central conversion facility would require a massive logistical effort. Over 250 dirty product deliveries per day would be necessary which would entail a truck fleet of approximately 90 vehicles.

1.2.5.3 Dallas Area Sample

As shown in Figure A-1-23, commodity clean products could be shipped from the Houston area to Dallas through several product pipelines. LPG markets in Dallas today are derived from local gas plants and refineries.

Region — Southwest
Sample City — Dallas, Texas

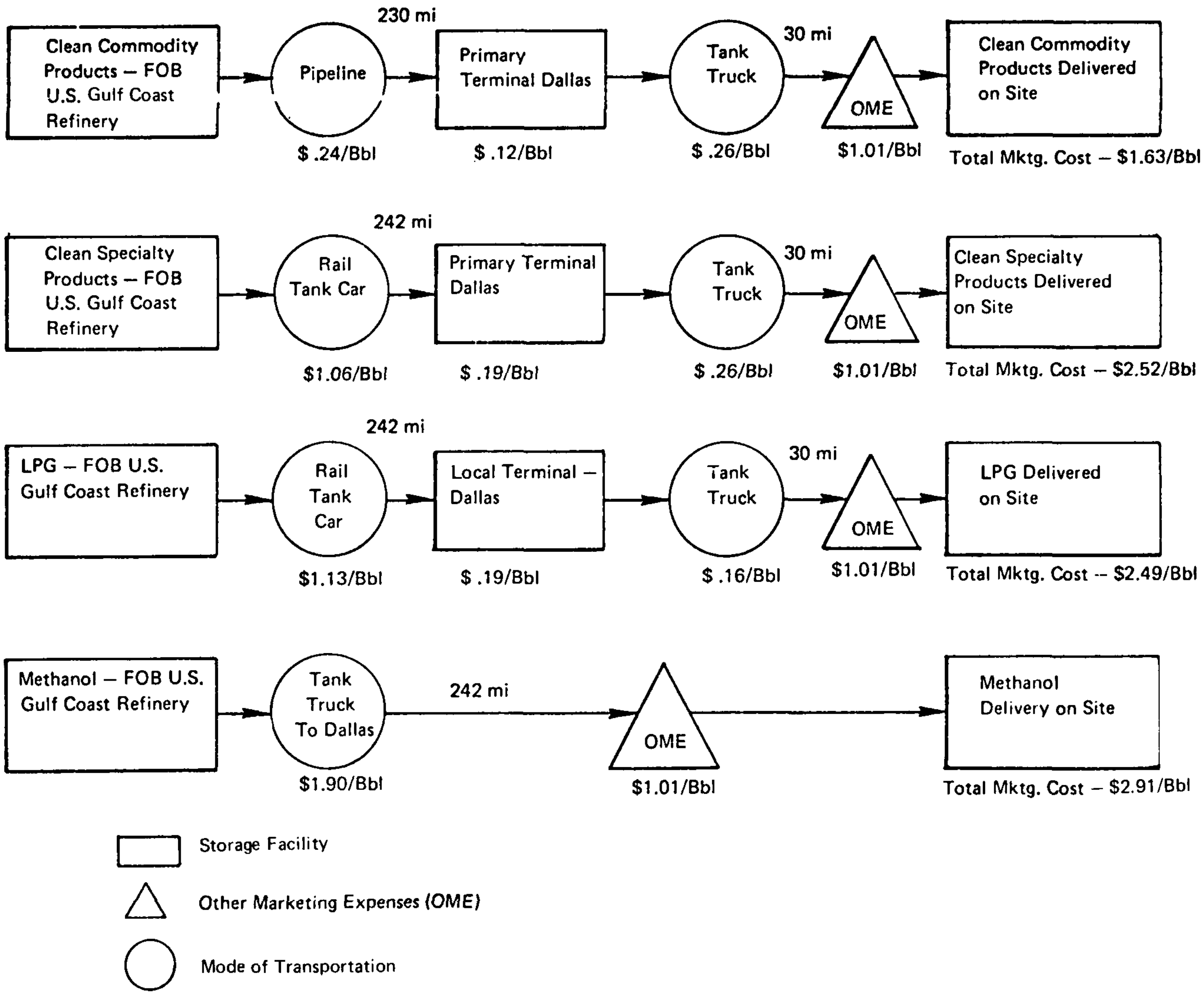

FIGURE A-1-23 CLEAN PRODUCTS – 10 DISPERSED FUEL CELLS LOGISTICAL LINKAGES AND COSTS

However, it is assumed that LPG for fuel cells would be derived from a terminal in the Houston area. Product pipelines flowing to the northwest from Houston seem to be quite full at this time and are generally not geared to move LPG. Furthermore, the quality control problem of shipping very low sulfur specialty clean products over a non-dedicated pipeline is still at issue. Thus, these two product groups could be moved to terminals in Dallas most effectively by jumbo tank cars. From a local Dallas terminal and/or bulk plants, full tank truck deliveries of clean products and LPG would be made to the dispersed fuel cells which are assumed to be located within a thirty mile radius of Dallas. Methanol and the two dirty products have to be directly shipped from Houston by truck to fuel cells located in Dallas since there are no local terminal facilities for these products (see Figure A-1-24). The truck transportation rates from Houston as well as local delivery tariffs would be subject to the regulatory authority of the Texas Railroad Commission.

As in other regions, truck deliveries of clean products from local terminals would only require a dedicated truck fleet of three or four vehicles. However, direct deliveries of methanol from Houston would require a fleet of over twenty-two truck transports. Based upon dirty products movements from the Houston area, the logistics of servicing a centralized conversion facility in Dallas seem quite infeasible. Over 230 to 250 truck deliveries per day would be required which would necessitate a dedicated truck fleet of 280 to 300 vehicles. This level of activity entails the receipt of 10 trucks per hour with the need for over a dozen offloading points. This logistical problem could be mitigated if a central conversion facility included a railroad right-of-way which could permit direct shipments of jumbo tank cars to on-site storage. However, this step would further require the utilization of dirty product tankcars which currently are generally not available in this area.

1.2.5.4 Greater Los Angeles Area Sample

As in the Dallas example, petroleum fuels for fuel cells in the Los Angeles area are assumed to be available from local refineries. This alleviates the need for both long distance bridging transportation and

Region – Southwest
Sample City – Dallas Texas

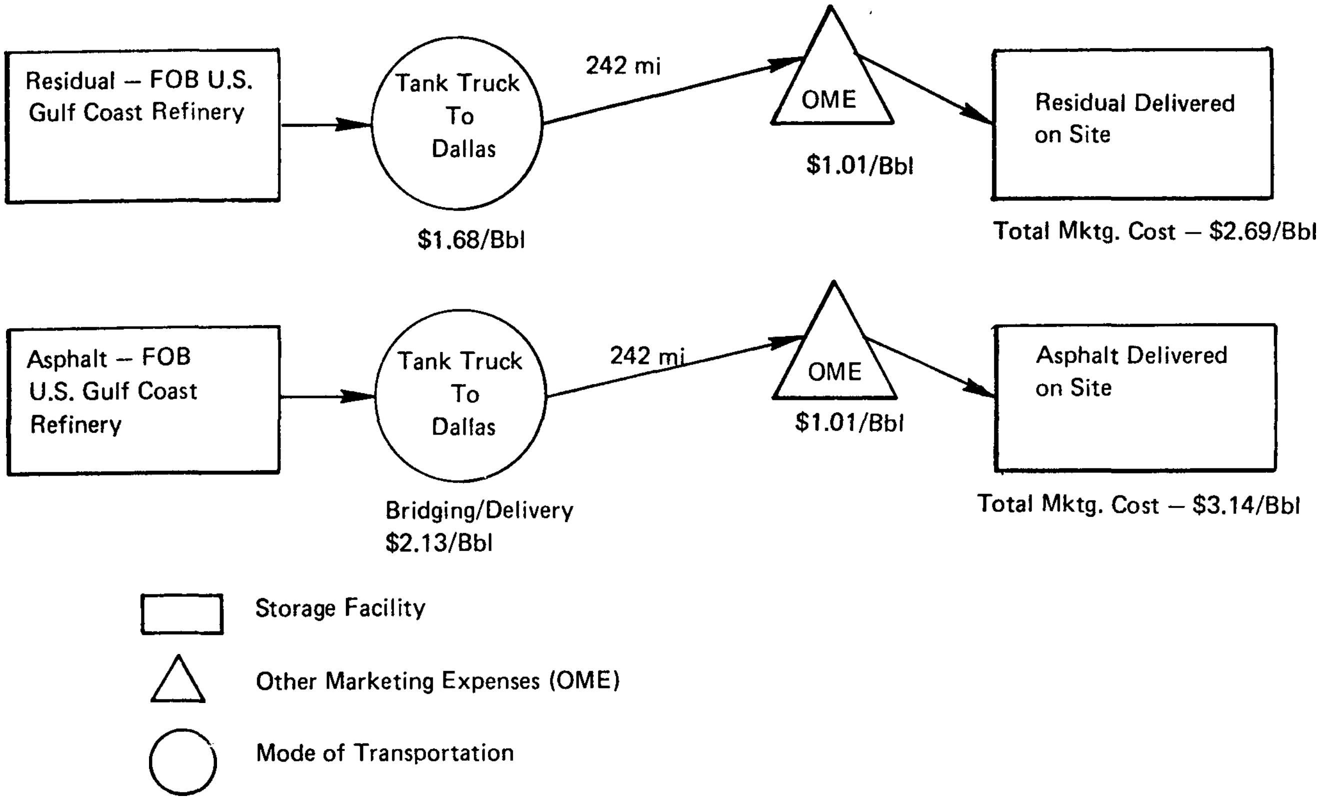

FIGURE A-1-24 DIRTY PRODUCTS - CENTRAL CONVERSION OPTION
LOGISTICAL LINKAGES AND COSTS

primary terminal costs. It has been assumed that fuel cell facilities in the Los Angeles area would be located approximately 45 miles from the local key refineries. Commodity clean products for fuel cells could be sent from these refineries to secondary bulk plants in the suburbs by the Southern Pacific and/or the San Diego pipelines (see Figure A-1-25). Due to quality or handling problems with pipelines, all of the other potential fuels would be directly shipped from a truck loading rack in a refinery to the fuel cell facilities (Figure A-1-26). The cost of both the primary products storage and the truck loading facilities in Los Angeles are considered to be part of the refinery costs.

As in the other three illustrative cases, dispersed fuel cells would require approximately 2 deliveries per day while the central conversion option would need over 250 daily truck loads of dirty products. Clean products could be handled by 2 or 3 dedicated trucks which are operated on a double shift basis. Fuel demand of a central conversion facility could be met with a dedicated dirty product fleet of 30 tractor trailers which are also double shifted.

1.2.5.5 Delivered Oil Products Prices

Summaries of the prices of refinery-delivered oil products prices delivered to either dispersed or central fuel cell locations in Hartford, Columbus, Dallas and Los Angeles are listed in Tables A-1-33 to A-1-36.

Region – West Coast
Sample City – Los Angeles

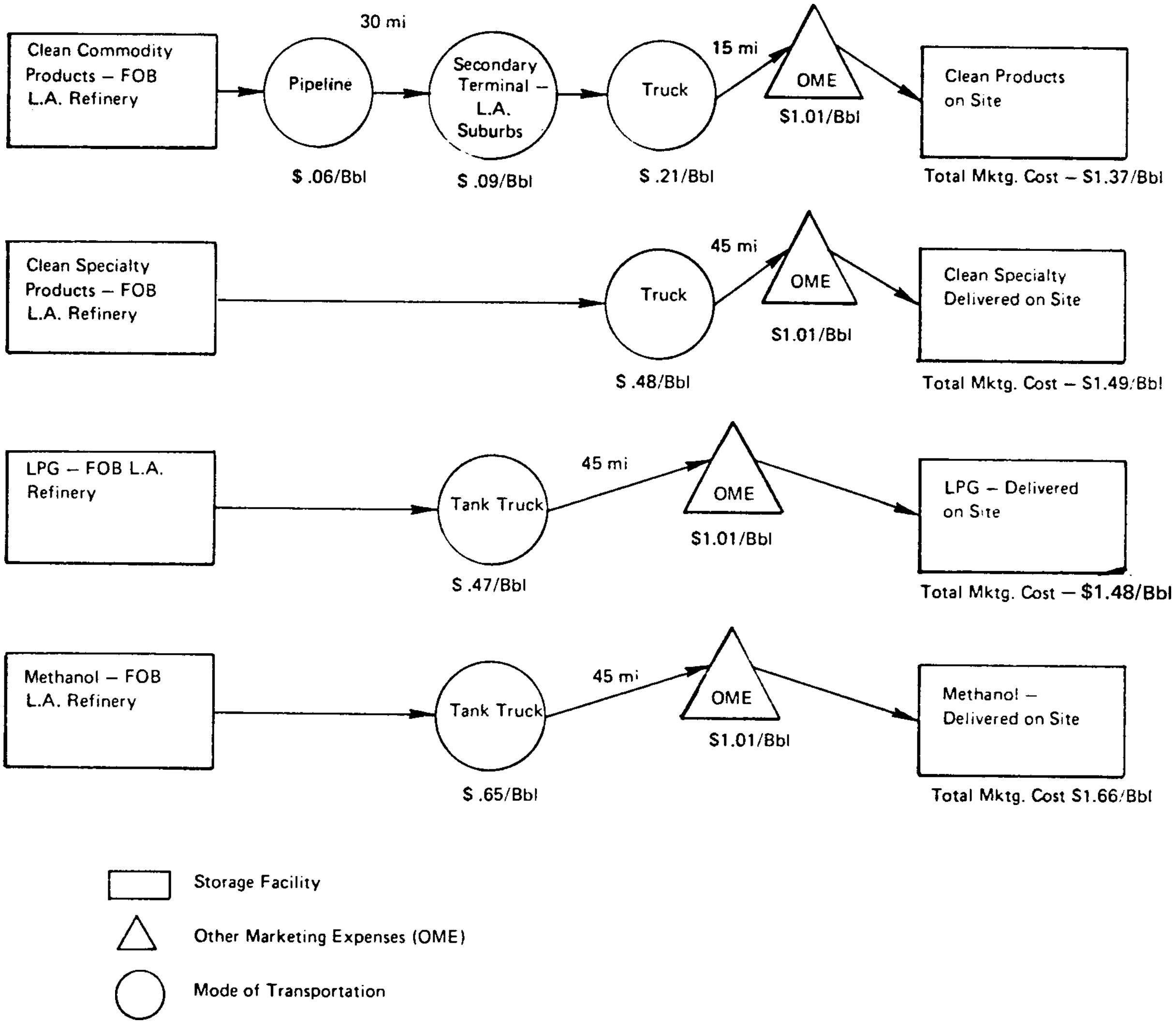

FIGURE A–1–25 CLEAN PRODUCTS – 10 DISPERSED FUEL CELLS LOGISTICAL LINKAGES AND COSTS

Region – West Coast
Sample City – Los Angeles

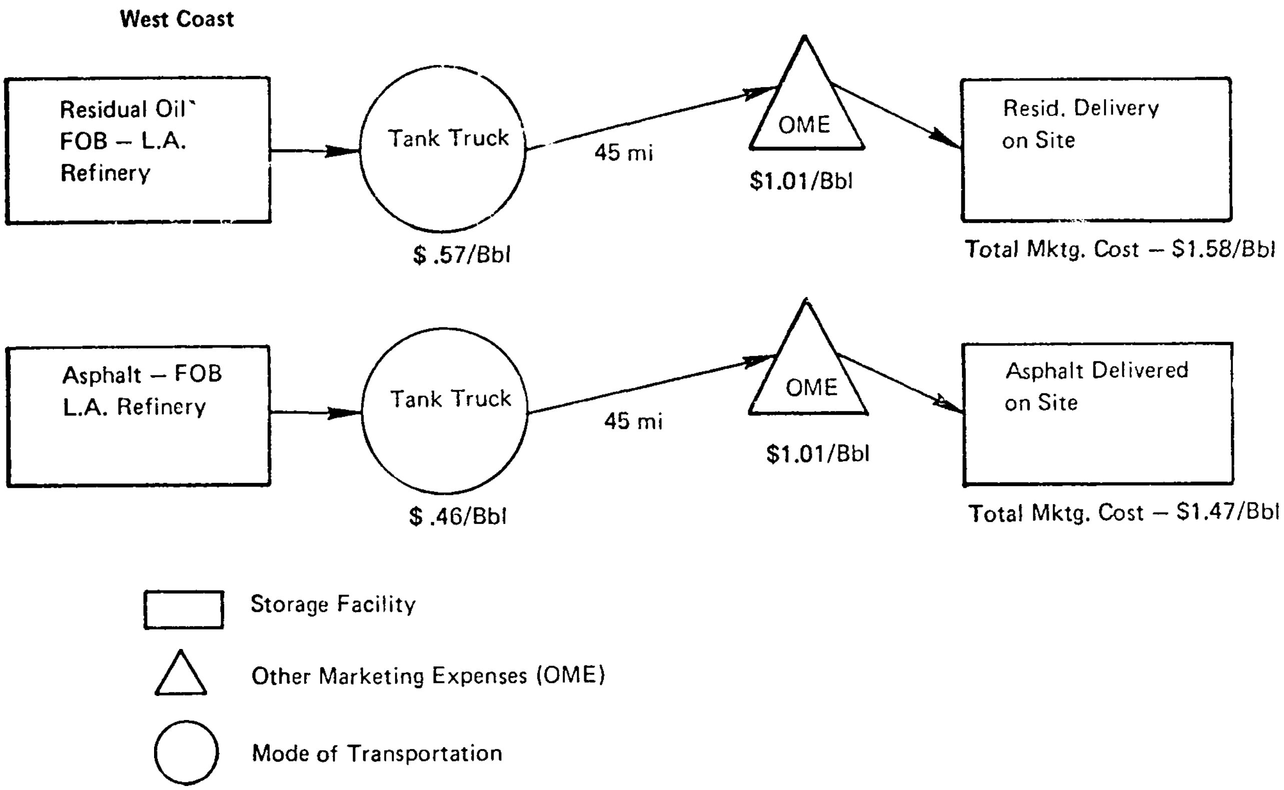

FIGURE A-1-26 DIRTY PRODUCTS - CENTRAL CONVERSION OPTION LOGISTICAL LINKAGES AND COSTS

TABLE A-1-33

DELIVERED OIL PRODUCTS PRICES - HARTFORD

(1975 $/Bbl)

CRUDE PRICE SCENARIO	1980		1985		1990	
	CONTROLS	DECONTROL	CONTROLS	DECONTROL	CONTROLS	DECONTROL
Products						
Unleaded Gasoline	19.55	20.73	22.63	27.39	23.34	28.45
0.5 ppm Naphtha	17.78	19.01	21.14	25.76	21.91	26.86
Virgin Naphtha	16.68	17.86	19.95	24.41	20.68	25.46
0.5 ppm Kerosine	18.63	19.98	22.50	27.51	23.79	29.17
200 ppm Kerosine	17.04	18.29	20.73	25.46	21.95	27.03
Virgin Kerosine	16.78	18.03	20.47	25.20	21.69	26.77
0.5 ppm No. 2 Fuel	19.48	20.94	23.85	29.26	25.67	31.48
200 ppm No. 2 Fuel	17.17	18.49	21.29	26.28	23.00	28.37
Virgin Gas Oil	16.87	18.19	20.99	25.98	22.70	28.07
0.5% S No. 6 Fuel	15.74	17.06	19.86	24.85	21.57	26.94
Asphalt	12.37	13.58	15.78	20.34	17.00	21.91

Source: A.D. Little Estimates

TABLE A-1-34

DELIVERED OIL PRODUCTS PRICES - COLUMBUS

(1975 $/Bbl)

CRUDE PRICE SCENARIO	1980 CONTROLS	1980 DECONTROL	1985 CONTROLS	1985 DECONTROL	1990 CONTROLS	1990 DECONTROL
Products						
Unleaded Gasoline	19.55	20.73	22.63	27.39	23.34	28.45
0.5 ppm Naphtha	18.28	19.51	21.64	26.26	22.41	27.36
Virgin Naphtha	16.68	17.86	19.95	24.41	20.68	25.46
0.5 ppm Kerosine	19.13	20.48	23.00	28.01	24.29	29.67
200 ppm Kerosine	17.04	18.29	20.73	25.46	21.95	27.03
Virgin Kerosine	16.78	18.03	20.47	25.20	21.69	26.77
0.5 ppm No. 2 Fuel	19.98	21.44	24.35	29.76	26.17	31.98
200 ppm No. 2 Fuel	17.17	18.49	21.29	26.28	23.00	28.37
Virgin Gas Oil	16.87	18.19	20.99	25.98	22.70	28.07
1.0% S No. 6 Fuel	15.55	16.84	19.52	24.43	21.14	26.41
Asphalt	12.94	14.15	16.35	20.91	17.57	22.48

Source: A.D. Little Estimates

TABLE A-1-35

DELIVERED OIL PRODUCTS PRICES - DALLAS

(1975 $/Bbl)

	1980		1985		1990	
CRUDE PRICE SCENARIO	CONTROLS	DECONTROL	CONTROLS	DECONTROL	CONTROLS	DECONTROL
Products						
Unleaded Gasoline	18.53	19.71	21.61	26.37	22.32	27.43
0.5 ppm Naphtha	17.42	18.65	20.78	25.40	21.55	26.50
Virgin Naphtha	15.66	16.84	18.93	23.39	19.66	24.44
0.5 ppm Kerosine	18.27	19.62	22.14	27.15	23.43	28.81
200 ppm Kerosine	16.02	17.27	19.71	24.44	20.93	26.01
Virgin Kerosine	15.76	17.01	19.45	24.18	20.67	25.75
0.5 ppm No. 2 Fuel	19.12	20.58	23.49	28.90	25.31	31.12
200 ppm No. 2 Fuel	16.15	17.47	20.27	25.26	21.98	27.35
Virgin Gas Oil	15.85	17.17	19.97	24.96	21.68	27.05
0.7% S No. 6 Fuel	14.92	16.23	18.98	23.94	20.65	25.99
Asphalt	11.61	12.82	15.02	19.58	16.24	21.15

Source: A.D. Little Estimates

TABLE A-1-36

DELIVERED OIL PRODUCTS PRICES - LOS ANGELES

(1975 $/Bbl)

	1980		1985		1990	
CRUDE PRICE SCENARIO	CONTROLS	DECONTROL	CONTROLS	DECONTROL	CONTROLS	DECONTROL
Products						
Unleaded Gasoline	18.67	19.44	20.96	25.28	21.50	25.89
0.5 ppm Naphtha	17.24	18.00	19.79	24.04	20.37	24.67
Virgin Naphtha	16.24	16.97	18.71	22.82	19.25	23.41
0.5 ppm Kerosine	18.40	19.20	21.39	25.87	22.59	27.12
200 ppm Kerosine	16.90	17.65	19.74	23.96	20.86	25.14
Virgin Kerosine	16.64	17.39	19.48	23.70	20.60	24.88
0.5 ppm No. 2 Fuel	19.56	20.40	22.99	27.69	24.80	29.56
200 ppm No. 2 Fuel	17.33	18.10	20.55	24.87	22.25	26.64
Virgin Gas Oil	17.03	17.80	20.25	24.57	21.95	26.34
0.5% S No. 6 Fuel	14.43	15.19	17.97	22.26	20.46	24.85
Asphalt	10.81	11.54	13.29	17.41	14.35	18.50

Source: A.D. Little Estimates

2.0 COAL AND COAL-DERIVED LIQUIDS

2.1 SUPPLY/DEMAND TRENDS

2.1.1 Coal

The United States has abundant supplies of coal which have been unexploited due to the historic availability of low-cost, cleaner burning petroleum and natural gas. Two-thirds of the nation's reserves are located in the Rocky Mountain area, where the coal tends to be low in sulfur and concentrated in thick seams near the surface so that it is amenable to strip mining. The immediate constraints on higher coal consumption involve delays in establishing new mines (environmental obstacles and the time to mobilize equipment and manpower), developing an adequate transportation network to enable the coal to reach the market areas economically, and resolving the significant air pollution problems associated with coal's use.

Forecasted coal production figures are shown in Table A-2-1 in natural units. They assume favorable government policies towards leasing federal lands and resolution of the current debates on sulfur emission control (either via acceptable technical innovations or relaxation of environmental control regulations). The use of coal for gasification and liquefaction is assumed to be insignificant in the period to 1990.

2.1.2 Coal-Derived Liquids

The emergence of a synthetic fuels industry would be a significant development for fuel cell power generation since the light hydrocarbon portions from coal liquefaction plants would be potential fuel for fuel cells. However, general availability of coal-derived liquids (CDL) before 1990 seems remote.

A recent world energy supply forecast (R1) projects all synthetics* at about 1% of world supply by 1990. With world supply projected at 151 $\overline{M}$Bbl/d fuel oil equivalent (FOE) by then, synfuels would account for about

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*tar sands, shale, coal and heavy oils.

TABLE A-2-1

U.S. COAL SUPPLY AND DEMAND 1976-1990

(Million Tons)

	1976	1980	1985	1990
DOMESTIC DEMAND				
Electric Utilities	463	610	820	1070
Commercial/Industrial	57	95	105	180
Conversion	-	5	10	25
Metallurgical/Coking	93	105	115	120
Miscellaneous	4	5	5	5
Subtotal	617	820	1055	1400
Exports	42	45	75	100
TOTAL PRODUCTION	659	865	1130	1500

Source: A.D. Little Estimates

1.5 $\overline{\text{M}}$Bbl/d. The U.S. energy needs by 1990 will be 54.3 $\overline{\text{M}}$Bbl/d FOE with about 0.5 $\overline{\text{M}}$Bbl/d FOE from synthetics assuming the same portion as for world supplied. However, this would be equivalent to ten nominal 50 KBbl/d syncrude plants (e.g., coal, oil shale, etc.) by 1990. This assumes no allowance for coal-based SNG facilities. This is clearly optimistic since the first full-scale commercial plants are not likely to be in operation before 1985. Subsequent projects will take 4-5 years for completion if the experience with the initial plants is satisfactory. It would be more realistic to assume that syncrude capacity will be in the range of 100-300 KBbl/d by 1990. However, even at 500 KBbl/d, this source of fuel would not set market prices for refined petroleum products. Conversely, the economics of coal liquefaction will be determined by the prevailing prices of petroleum-derived fuels.

A related issue is the impact that the arrival of coal-derived liquids will have on the availability of fuel cell grade fuels. This analysis focuses on syncrude since distillate range fuels are of most interest to dispersed fuel cells.

The initial determination to make is the likely market outlet for the fractions obtained from coal-derived syncrude. Some insight is obtained by considering the composition of various distillate range fractions of syncrude. In Table A-2-2 are presented the inspections of three boiling range fractions from H-Coal syncrude. Of particular interest is the high concentration of ring structures including both cycloparaffins (naphthenes) and aromatics. Also of interest is the quantity of heterocyclic and non-hydrocarbon constituents. In the case of the C_4 to 400 naphtha fraction, the high naphthenic content makes it potentially suitable for the production of gasoline-blending stock via catalytic reforming. However, hydrorefining of this fraction prior to catalytic reforming will be necessary to remove the heteroatoms and increase the saturates.

The high aromaticity of the 400-650°F gas oil fraction is a major consideration in converting this fraction to turbine fuels. The low hydrogen-to-carbon ratio of this fraction makes it impossible to meet smoke point specifications on jet fuel without extensive hydroprocessing. To

TABLE A-2-2

H-COAL PRODUCT INSPECTION FROM ILLINOIS NO. 6 COAL

Syncrude Mode

DISTILLATE FRACTION:	C_4-400°F (Naphtha)	400-650°F (Gas Oil)	650-920°F (Vacuum Gas Oil)
Constituents, wt. %			
Paraffins	11.99	6.5	1.4
Saturated Naphthenes	51.13	24.5	5.4
Unsaturated Naphthenes	11.20	4.3	1.6
Alkyl Benzenes	17.55	12.6	3.0
Other Aromatics*	7.03	47.0	73.3
Heterocyclics†	0.9	2.0	1.5
Other Non-Hydrocarbons	0.2	3.10	13.8
TOTALS	100.00	100.00	100.0

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*Indans, tetralins, and polycyclic aromatics.

†Principally phenols.

Source: Battelle

some extent these assays represent some of the better grades of coal liquids since H-coal is a more severe coal hydrogenation process than some of the other technologies. Since it may be quite costly to upgrade the 400-650°F fraction further to jet fuels, some process developers see the entire 400-650°F fraction going to low sulfur fuel markets.

The potential outlets for coal-derived liquids and their interaction with petroleum refining are shown in Figure A-2-1. This is simplified representation of selected process routes within a refinery. The process steps starting with crude oil show the conventional routes to gasoline, No. 2 fuel and low sulfur residual fuel. The steps starting with coal liquids show the potential routes for utilizing coal-derived liquids (CDL). The 400°F CDL naphtha is distilled into three factions. The C_5-160°F light end is taken directly to gasoline blending; the middle cut goes through hydrorefining and catalytic reforming to gasoline blending; and the 350-400°F heavy end goes to No. 2 fuel oil. The low sulfur CDL fuel oil fraction ($400°F^+$) would be sold directly to a large utility user on a contract basis. Based on these potential outlets, the price of coal-derived naphtha will be set by its value in gasoline via catalytic reforming. Furthermore, at best, the 400-650°F gas oil will be valued as No. 2 distillate fuel and more likely valued as low sulfur residual fuel. The LPG produced from coal liquefaction is a potential fuel cell fuel requiring no upgrading.

2.2 COAL AND DERIVED FUELS PRICE FORECASTS

2.2.1 Coal Prices

There are two basic approaches to estimating the price of coal delivered to utilities. One is based on estimating the resource development cost of incremental supplies assuming new contracts will reflect these costs. The other is to determine an indifference price for coal assuming price competition with other fuels; namely, oil or uranium. However, if President Carter's energy policy is implemented, there will be essentially no price competition between coal and oil in the utility sector for new generation.

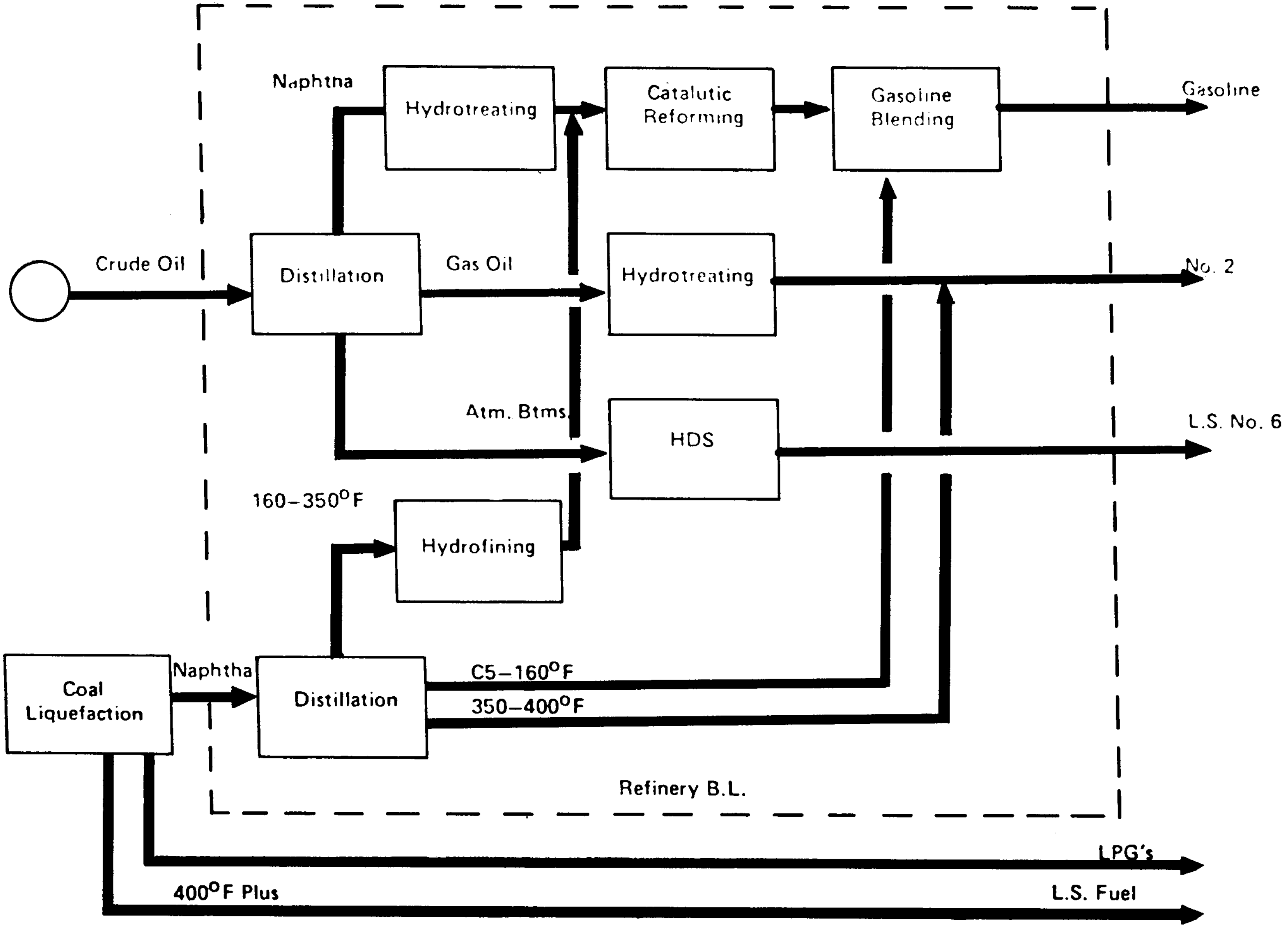
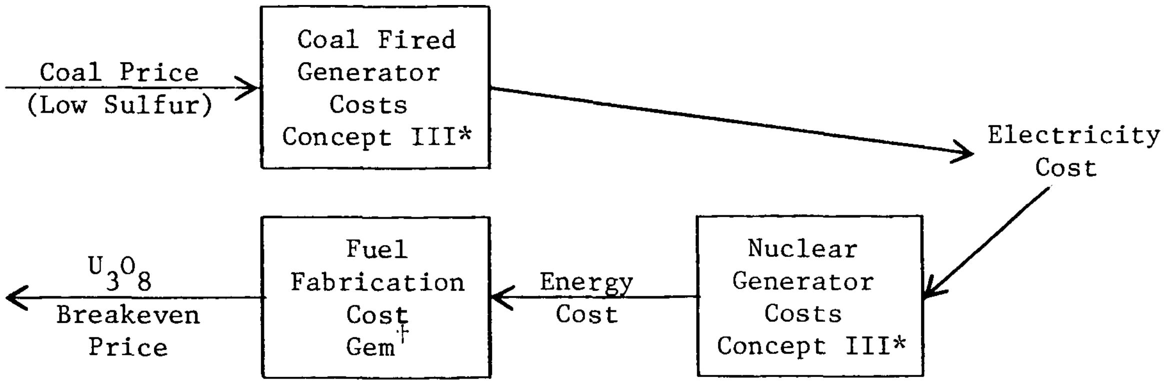


FIGURE A-2-1 POTENTIAL OUTLETS FOR COAL-DERIVED LIQUIDS

The cost elements for these two coal pricing mechanisms are generalized in Figure A-2-2 assuming nuclear power is the only real option for electric utilities. The cost-of-supply approach includes mine development and operating costs including transportation and a representative profit. The coal price based on inter-fuel price competition is derived by equating cost of electricity produced by nuclear and coal-fired plants and netting back the indifference price of coal; the price differential between low- and high-sulfur coal being the cost of flue gas desulfurization. In theory, the cost-based approach defines the minimum price required to justify investments in new mines. This was the method used by Arthur D. Little and Foster Associates for pricing coal in the RP-318 and RP-759-2 studies, respectively.

In regard to potential price competition between coal and nuclear power for electricity generation, a review of the Foster's Associates analysis of uranium prices in RP-759-2 is of interest. In their analysis of uranium prices Foster Associates developed the "breakeven" price of U_3O_8 with respect to low-sulfur coal as shown conceptually below:



The results of Foster Associates' evaluation are summarized in Table A-2-3 for the regions of interest in RP-1042. The "forecasted" price of uranium is also shown for medium and high price cases. This analysis indicates that the breakeven price of U_3O_8 based on Foster Associates'

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*reference 21

†reference 22

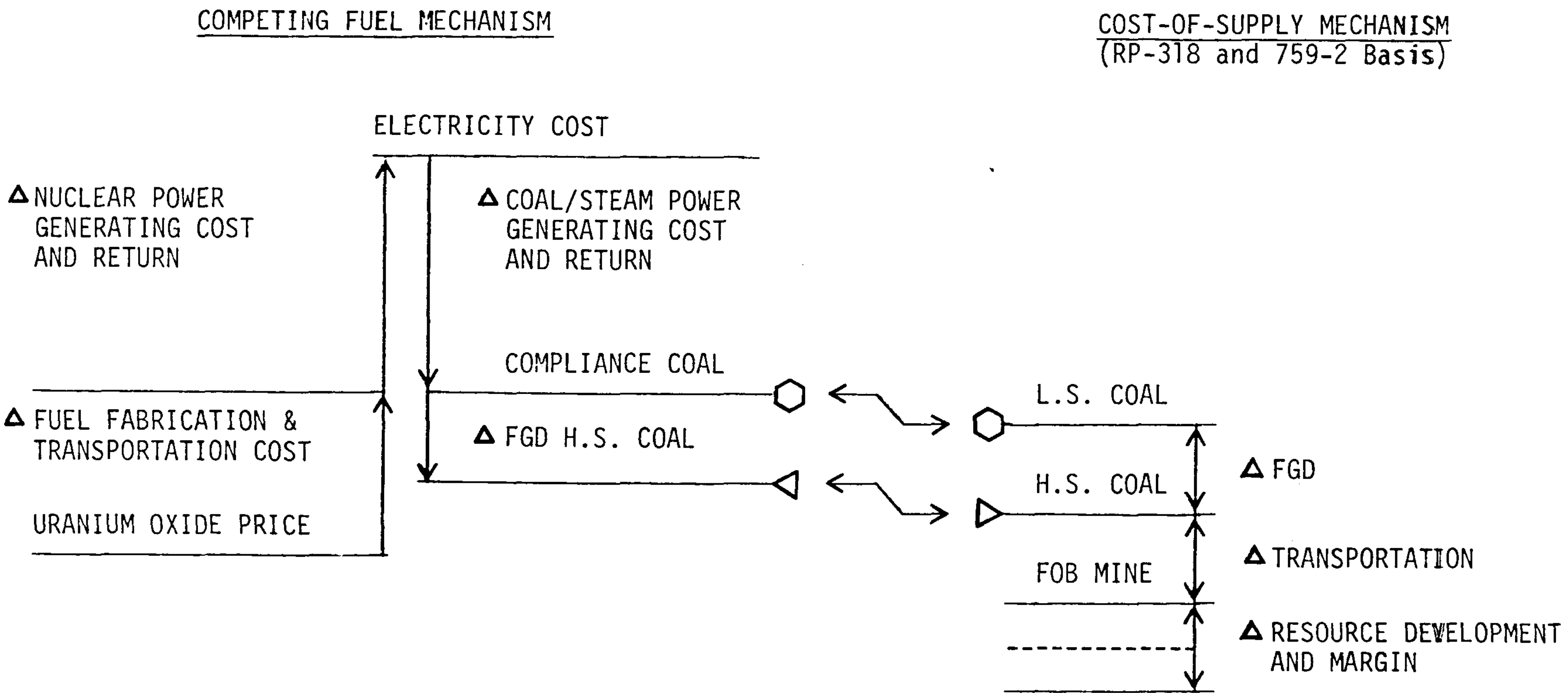

FIGURE A-2-2

COAL PRICING METHODOLOGIES

BASIS: NEW FACILITIES, UTILITIES SECTOR*

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*Utilities are under pressure from Department of Energy to convert to coal.

TABLE A-2-3

BREAKEVEN U_3O_8 PRICE WITH RESPECT TO LOW SULFUR COAL

(1975 Dollars)

Region:	N. England		E.N. Central		W.S. Central		Pacific/Mountain		Forecasted U_3O_8 Price \$/lb	
	Coal $\$/10^6$ Btu	U_3O_8 \$/lb	Coal $\$/10^6$ Btu	U_3O_8 \$/lb	Coal $\$/10^6$ Btu	U_3O_8 \$/lb	Coal $\$/10^6$ Btu	U_3O_8 \$/lb	Medium	High
1985	1.70	76	1.62	73	1.58	75	1.45/1.28	59/50	47	59.50
1990	1.81	74	1.74	78	1.70	84	1.56/1.41	61/57	52	88.75
1995	1.92	76	1.88	83	1.86	92	1.72/1.58	69/69	58	135.00

Source: Foster Associates for EPRI project 759-2.

forecasted coal prices is considerably above forecasted U_3O_8 prices; however, the differential is the smallest in the West. This suggests that only in the Western region is coal likely to be competitive with nuclear fuel.* With a clear need to develop all forms of energy, nuclear power represents, at best, a porous ceiling on the price of coal. Only for the case of very pessimistic (high) projections for uranium prices do the prices of coal and nuclear power come into equilibrium in the other geographical regions.

As a result of the above considerations, a cost-based approach to coal price forecasting was adopted. Normally an independent coal forecast would have been developed by Arthur D. Little on this basis. However, this aspect was not included in the scope of work under this contract and Foster Associates had recently developed price projections for EPRI (RP-759-2) using a similar methodology. Consequently, the Foster Associates' cost-based coal price projections were adopted as the basis for the RP-1042 assessment study. These projections for high and low sulfur coal are shown in Table A-2-4. Since basic coal conversion systems are evaluated in this study, the high sulfur coal values were used.

In the original forecast, Foster's high sulfur coal prices in the West South Central region were based on Texas lignite. The high sulfur coal values for this region shown in Table A-2-4 were reestablished by Arthur D. Little based on Foster's western sub-bituminous minemouth coal prices with appropriate freight to Dallas added. To calibrate this approach, the CIF coal prices in Los Angeles were first calculated assuming the same FOB minemouth prices. The reconstructed prices were consistent with the values reported by Foster for the Pacific region.

2.2.2 Coal-Derived Liquids Prices

2.2.2.1 Syncrude Cost

The cost of producing syncrude from coal was estimated for comparison with expected market values assuming competition with petroleum-derived

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*This is the present situation.

TABLE A-2-4

BASIC FUEL PRICE FORECAST SUMMARY HIGH SULFUR COAL AT BURNER TIP

($/$\overline{\text{M}}$Btu - 1975)

Region:	N. England		E.N. Central		W.S. Central		Pacific	
Year	LS	HS	LS	HS	LS	HS[¶]	LS	HS
1975*	1.24		0.82		0.23		0.57	
1980[†][§]	1.59	1.08	1.50	0.94	1.44	1.10	1.34	1.04
1985[†]	1.70	1.17	1.62	1.02	1.58	1.18	1.45	1.17
1990[†]	1.81	1.25	1.74	1.10	1.70	1.30	1.56	1.28
1995[†]	1.92	1.35	1.88	1.20	1.86	1.46	1.72	1.44

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*Average utility reportings to FPC.

[†] Cost based on incremental production from new mines.

[§] Extrapolated values.

[¶] Sub-bituminous coal, based on FOB mine cost plus unit train freight cost.

Source: EPRI EA-411 by Foster Associates.

fuels. The H-coal process was selected for this analysis since its development schedule is generally considered to be the most advanced liquefaction process. Representative cost information is also more readily available than for most other processes. The most detailed evaluation of H-coal economics is presented in ERDA report FE-200212 which was prepared by Fluor Engineering. (20) This reference was used as the basis for the economics developed herein.

The block flow diagram for a 50 Kbbl/d syncrude plant based on Illinois No. 6 coal is shown in Figure A-2-3. The major process units include coal preparation, coal hydrogenation, H_2 manufacture by partial oxidation of char, gas treating, waste treating and plant utilities. The flags on the process streams in Figure A-2-3 coincide with the material balance stream number designations shown in Table A-2-5. The overall utility summary for the facility is presented in Table A-2-6. The plant is in energy balance except for exported power (82 $\overline{M}$w) and imported water (842 Klb/hr).

Pro forma economics for the production of syncrude are displayed in Table A-2-7 for the E.N. Central and New England regions* based on delivered coal prices in 1990; costs are presented in 1975 dollars. The total investment for the coal plex is \$1387 $\overline{M}$, \$1233 $\overline{M}$ is depreciable investment in fixed plant and start-up expense and \$154 $\overline{M}$ is nondepreciable investment for working capital and interest during construction.† Operating costs include coal, other utility costs and credit for byproducts and power. Capital charges with and without federal income tax included are presented. The unit cost of producing syncrude with appropriate capital recovery is in the range of \$32-35/bbl with federal taxes and \$28-31.50/bbl without federal taxes. These cost-derived values are significantly above projected landed crude oil prices in the same time period, without taking account of quality differences. The effect of quality difference is considered in part in Section 2.2.2.2.

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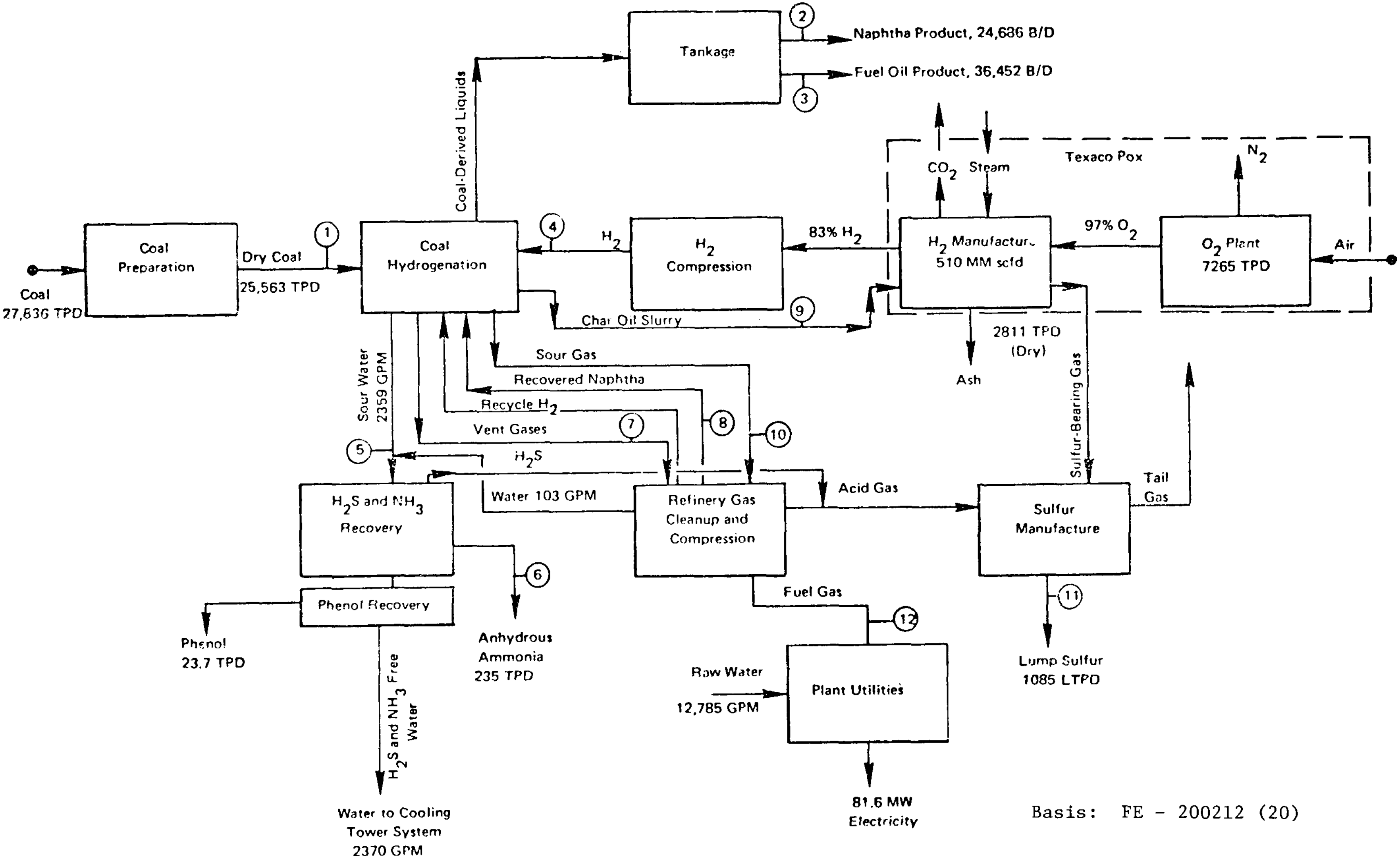
*most likely areas with access to high sulfur eastern coal

†based on accounting procedure used by Fluor.



Basis: FE - 200212 (20)

FIGURE A-2-3 H-COAL - BLOCK FLOW DIAGRAM (SYNCRUDE MODE) - ILLINOIS NO. 6 COAL

TABLE A-2-5

MATERIAL BALANCE
H-COAL LIQUEFACTION
(Illinois No. 6)

STREAM NO.	1		2		3		4		5	
	Coal		Naphtha		Syncrude		Hydrogen		Sour Water	
TEMP (°F)										
			24,686		36,452					
Component Flows	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr
$CO + N_2$							1,885		0.3	
H_2							53,683		4.5	
CH_4							250		0.3	
C_2H_6										
C_3H_3										
C_4H_{10}										
CO_2							142		363	
H_2S									657	
NH_3									1,006	
H_2O	2,363	42,540			46	826	48	856	65,290	1,175,223
Naphtha		--	2,823	296,018						
500 + F.O.		--			2,226	455,909				
Coal (DAF)		1,876,146				--				
Ash		211,554								
Total		2,130,240	2,823	296,618	2,272	456,735	56,008	177,868	67,321	1,230,755

Source: FLUOR.

TABLE A-2-5

H-COAL LIQUEFACTION Continued

STREAM NO.	6		7		8		9		10	
	Ammonia		Raw Gas		C_5 Plus		Char Slurry		Sour Gases	
TEMP (°F)										
BBL/D					18,823		45,017			
Component Flows	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr
CO + N_2			7,973		7.8				353	
H_2			71,759		28.5				1,995	
CH_4			14,100		70				660	
C_2H_6			2,408		61				373	
C_3H_8			1,115		81				274	
C_4H_{10}			797		149				304	
CO_2			1,405						28	
H_2S		5 ppm	2,415		126				399	
NH_3	1,148		833						--	
H_2O	5		9,069	163,242	1	18			118	2,123
Naphtha			1,760		1,296				140	
500 + F.O.			385		384		956.9	522,778		
Coal (DAF)								91,581		
Ash								211,554		
Total	1,153	19,640	114,019	1,323,645	2,222	224,634		825,913	4,644	92,959

TABLE A-2-5

H-COAL LIQUEFACITON Concluded

STREAM NO.	11		12							
	Lump Sulfur		Fuel Gas							
TEMP (°F)										
SCFD			144.1×10^6 (Dry)							
Component Flows	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr
$CO + N_2$			1,154							
H_2			6,761							
CH_4			4,115							
C_2H_6			1,726							
C_3H_8			997							
C_4H_{10}			735							
CO_2			--							
H_2S			--							
NH_3			--							
H_2O			4,782	14,068						
Naphtha			351							
500 + F.O.			2							
Coal (DAF)										
Ash										
Total	3,164	101,248	16,623	290,863						

TABLE A-2-6

UTILITY SUMMARY
SYNCRUDE MODE

		Coal Preparation Unit 100	Coal Hydrog-enation Unit 200	Product Recovery Unit 300	Gas Processing Unit 400	Sulfur Plant Unit 500	Oxygen Plant Unit 600	Hydrogen Plant Unit 700	Sour Water Stripping Unit 800	Product Storage & Loading Unit 900	Utility & Offsites Unit 1000	TOTALS	REMARKS
Electric Power, kw	P										303,059	303,059	81,600 Exported
	C	14,206	25,300	17,622	23,778	3,540	68	85,311	2,500	1,756 Avg.	47,378	221,459	
Cooling Water, gpm	P										387,900	387,900	Excess Capacity Provided
	C	250	3,440	4,347	2,571	13,800	103,000	37,700	9,800	393	158,550	333,851	
Fuel Oil, M̄Btu/hr	P												
	C									3		3	
Fuel Gas, M̄Btu/hr	P				5,533							5,533	
	C	486	635	161		35					4,216	5,533	
Raw Water, gpm	P												
	C										12,785	12,785	
Steam, 1500 psig, lb/hr	P							1,155,800			3,040,200	4,196,000	
	C						1,036,300	489,000			2,670,700	4,196,000	
Steam, 600 psig, lb/hr	P										42,200	42,200	
	C			27,500	14,700							42,200	
Steam, 150 psig, lb/hr	P			124,000		273,600		362,000			96,000	855,600	
	C			151,700				84,000	151,000		468,900	855,600	
Steam, 50 psig, lb/hr	P			197,300	236,000			257,400			481,500	1,192,400	
	C				221,600	2,900		497,100	147,000		323,800	1,192,400	
BFW, lb/hr	P										4,873,100	4,873,100	842,300 Make-up Water Required
	C			327,900	240,600	285,000		1,782,500			3,079,400	5,715,400	
Condensate, lb/hr	P			27,500	236,900	2,500	1,036,300	344,300	298,000		2,580,600	4,526,500	
	C							839,200			3,687,300	4,526,500	

LEGEND: P - produced
C - consumed

Source: Fluor Engineering

TABLE A-2-7

COST OF SYNCRUDE FROM COAL

Basis: H-Coal Process
Illinois No. 6 Coal
50,000 BPD Capacity - 8000 Operating Hr/Yr

(Mid-1975 Dollars)

Region:	N. England	E.N. Central
Coal Price, \$/Ton*	30	24
Depreciable Investment (DI),\$$\bar{M}$	1233	1233
Non-Depreciable Investment (NDI),\$$\bar{M}$	154	154
Total	1387	1387
Operating Cost, \$$10^6$		
Feed-Coal, 27,836 TPD	278.18	222.55
Other Operating Costs	105.15	105.15
Byproduct & Power Credits	(39.54)	(39.54)
Capital Charges		
(A) With Federal Tax		
DI - 18%	221.94	221.94
NDI - 14.7%	22.64	22.64
(B) Without Federal Tax		
DI - 13.3%	163.99	163.99
NDI - 10.0%	15.40	15.40
Total Operating Cost		
(A)	588.37	532.74
(B)	523.18	467.55
Unit Cost, \$/bbl†		
(A)	35.44 (5.09)	32.09 (4.61)
(B)	31.51 (4.52)	28.16 (4.04)

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*Compatible with \$20/bbl crude price in 1990 timeframe.

†Values in parenthesis are \$/$\bar{M}$Btu.

Source: Fluor Engineering (20)

2.2.2.2 Naphtha Value

One way to value coal-derived naphtha (CDN) is to establish a price for motor gasoline made from petroleum and then netback the value of naphtha feed through catalytic reforming and hydrorefining economics. As previously shown (Figure A-2-3), the processing routes to gasoline for straight-run naphtha and coal-derived naphtha are similar. However, the degree of hydrorefining required is an important difference as discussed in Section 2.1.2.

An evaluation of coal-derived naphtha value has recently been conducted for ERDA and EPRI based on the Doner Solvent Process. (18) To determine the netback price of coal-derived naphtha, a consistent set of petroleum product values including gasoline was established by L.P. model based on conventional petroleum refining economics. These values in conjunction with refinery processing costs were used to back-calculate the value of CDN as a feedstock for catalytic reforming.

Using a similar approach, we establish a representative set of petroleum product prices using the refinery model discussed in Appendix section 1.2.4. These values are shown in Table A-2-8 for various products consistent with Arabian light crude valued in our forecast at $20/bbl.* Note that all the major fuels are fairly similar in value. The netback prices of naphtha based on gasoline at $24.91/bbl are also developed in Table A-2-8 for naphtha from the Doner Solvent process and straight-run naphtha from Arabian light crude. The indifference prices of the naphtha feeds are $22.70 for Doner Solvent naphtha and $23.47 for the Arab light naphtha--a difference of $0.77/bbl in favor of the Arab light naphtha. This is of particular interest since the Arab light naphtha is paraffinic and a relatively poor feedstock for motor gasoline production. There are two factors which contribute importantly to this result. The first is that with all the petroleum products having similar value, the superiority (high ring structure) of the coal-derived naphtha as a catalytic reformer feedstock is diminished. This is because the higher yield of gasoline from the CD naphtha is of little economic advantage if gasoline values are substantially

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*Consistent with A. D. Little 1990 crude forecast.

TABLE A-2-8

CONSISTENT PETROLEUM VALUES

(1975 Dollars)

	$/Bbl
Arab Light Crude	20.00
Motor Gasoline	24.91
No. 2 Distillate	25.15
C_1-C_4 Gas L.S. Fuel Oil	24.73 (FOE)*

NETBACK PRICE OF NAPHTHA

Basis: Gasoline at $24.91/Bbl

Item	Value $/Bbl Naphtha Feed CD Naphtha	Value $/Bbl Naphtha Feed Arab Light
Mogas	17.99	21.86
Hydrogen	0.43	1.25
Distillate	6.60	0
C_1-C_4 Gas	0.62	2.29
Fuel	(1.09)	(0.69)
Total Value of Products	24.55	24.71
Operating Costs[†]		
Capital Recovery	1.10	0.63
O & M	0.75	0.61
Total Operating Cost	1.85	1.24
Value of Feed Naphtha	22.70	23.47
Delta	0.77	

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*6.32 x 10^6 Btu/FOE Bbl (HHV).

[†] ERDA study values deflated at 5% p.a.

Source: Exxon R&E and A.D. Little

the same as other fuels. Secondly, the operating cost associated with hydrorefining the coal-derived naphtha is substantially greater than for Arab light naphtha. The economics for the CD naphtha case might improve if H-coal naphtha were considered since the quantity of heteroatoms requiring removal is less. However, at best, it would not be much better than Arab light naphtha and certainly not as good as naphthenic naphthas such as obtained from Nigerian or Venezuelan crude.

This analysis indicates that coal-derived naphtha would bring a value equivalent to paraffinic straight-run naphtha. Hence, its price in the 1980-1990 timeframe would be comparable to the forecasted petroleum naphtha prices presented in this report.

It is also interesting that if the naphtha portion of the total coal liquid and the 400^+ fraction are valued at the fuel values previously shown in Table A-2-8, the weighted value of syncrude shown in Table A-2-9 becomes \$23.64/bbl. This is \$3.64/bbl above the value of imported Arabian light crude. In fact, if the naphtha fraction was valued at fuel value, the premium over Arabian light would increase to approximately \$4/bbl. A similar result was obtained by Bonner and Moore in 1973 for crude valued at \$6.50/bbl. (19) This analysis suggests that the real value of syncrude is associated with its low sulfur content. A comparison of this imputed value with the cost-based valuation, presented earlier, indicates that the economics of coal liquefaction are still not favorable in this timeframe.

The introduction of coal-derived syncrude also has implications regarding the availability of naphtha within the refinery. Typical yields of motor gasoline and low sulfur fuel oil from CDL and Arabian light crude are shown in Table A-2-10. Assuming fuel oil production is the primary objective, 125 barrels of Arabian light crude can produce 50 barrels of low sulfur fuel oil and 28.8 barrels of motor gasoline. To produce an equivalent quantity of low sulfur fuel oil requires only 100 barrels of syncrude. However, the yield of gasoline is only 26 barrels resulting in a net loss of 2.8 barrels of motor gasoline per 50 barrels of low sulfur fuel oil produced. Presumably this deficit would be made up by diverting petroleum naphtha into gasoline production with a net effect of putting pressure on

TABLE A-2-9

SYNCRUDE VALUES

Syncrude Fraction	Yield Vol %	Value $/Bbl
	Arab Light = $20/Bbl	
C_5 - 400°F	42.4	22.70
400^+ Fuel Oil	57.6	24.34*
Syncrude	100.0	23.64 (weighted)

Bonner & Moore study results: Arabian light = $7/bbl

Syncrude = $8+/bbl

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*$24.73/FOE adjusted for HHV content.

TABLE A-2-10

TYPICAL YIELDS

(Vol %)

	Arabian Light Crude	CDL's	CDL-ALC
Mogas	23	26	
LSFO	40	50	
Total HC's, bbl	125	100	-25
LSFO, bbl	50	50	-
Mogas, bbl	28.8	26	$\boxed{-2.8}$

the availability of straight-run naphtha. Admittedly, the substitution of 100 bbl of CDL's did free up 125 barrels of Arabian light crude. However, if the refinery capacity is not available to process additional crude, this does not improve the availability of naphtha. This rather simplistic analogy is included to illustrate that coal-derived fuels and petroleum products are interrelated and the answer to the availability issue is not straightforward.

The conclusions drawn from this analysis are summarized below for naphtha and distillate obtained from coal-derived syncrude:

Coal-Derived Naphtha

- Raw CD naphtha must be upgraded to be suitable for catalytic or steam reforming (need to remove heteroatoms, etc.)
- Hydrorefined CDN is a desirable catalytic reformer feed due to its high ring structure (naphthenes).
- Hydrorefined CDN may also be suitable as a feedstock for steam reforming, but not as good as a paraffinic SR naphtha.
- The arrival of a coal liquefaction industry does not automatically insure a greater abundance of reformable naphtha if fuel oil is the principal outlet.
- The price of refined CDN will be set by its value in motor gasoline production and its value will be roughly comparable to SR naphtha in the 1980-1990 timeframe.

Coal-Derived Distillates

- Due to their extreme aromaticity, CD distillates will require severe hydroprocessing in order to obtain specification turbine fuels.
- The most valuable property of the distillate fraction is a low sulfur content. It is more likely that this fraction will be left in the 400°F plus product and sold as low sulfur fuel oil for industrial and utility uses.
- Consequently in the 1980-1990 timeframe, the value of CD distillate will be comparable to petroleum-derived low sulfur fuel oil.

2.2.2.3 Coal-Derived Methanol

Methanol was considered as a fuel for fuel cells very early in the development of commercial systems. Methanol offers several advantages over other fuel cell fuels. Pure methanol can be reformed at much lower temperatures than light hydrocarbons and optimization of a low temperature reformer design for methanol conversion would result in improved efficiencies. In addition, methanol is sulfur-free, hence, no hydrodesulfurization (HDS) system is required. A sulfur removal system (ZnO) is still required since sulfur contamination may occur during shipping.

The production cost of coal-derived methanol was developed based on a process flow scheme using technically proven process technology. Methanol is synthesized via the ICI process from synthesis gas produced from gasified coal using the Texaco partial oxidation process. Texaco gasification was chosen over other gasification processes because it has been commercially demonstrated with a variety of feedstock and has been piloted with different coals. Moreover, there are no limitations of the Texaco gasifier to the use of caking coals as there are for the Lurgi gasifier. In addition, the process can be operated at high pressures. This eliminates compression of the synthetic gases prior to methanol synthesis. The ICI methanol synthesis system is a highly developed process and is operated in about 14 locations. A newly developed liquid phase methanol synthesis process,* offers a potentially higher thermal efficiency and lower capital cost than the ICI System. However, since the process has not yet been commercially proven, it was not used in this analysis.

The overall process flow scheme, depicted by Figure A-2-4, includes five major operations including coal preparation, gasification, water-gas shift, acid gas removal, and methanol synthesis. In addition a number of auxiliary operations such as oxygen production, sulfur recovery, waste heat recovery, and steam generation are required.

Coal Preparation. The coal is ground and slurried with water. The slurry (50-55% coal) is pumped to 1150 psi and is preheated to 600°F in a furnace, fired by vent gas from the methanol synthesis loop.

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*Offered by Chem Systems.

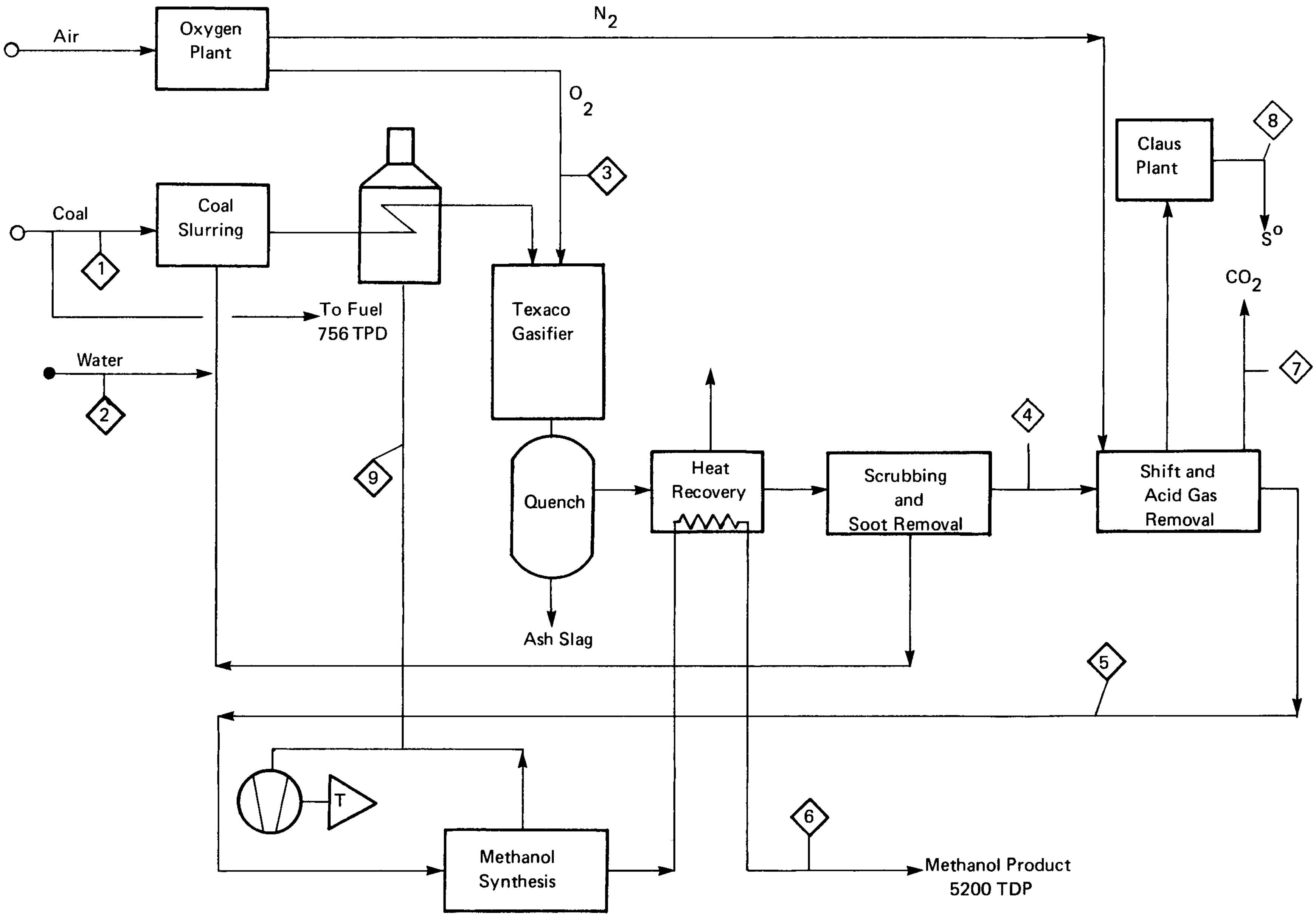

Figure A-2-4 METHANOL FROM COAL WITH TEXACO GASIFICATION

Gasification. The preheated slurry is injected into the combustion zone of the gasifier. The coal is partially oxidized at 1100 psi and 2300°F. The synthesis gas produced is quenched to 1700°F by direct contact with water. The quenched slag and soot is removed from the bottom of the gasifier vessel through a lock-hopper. The quenched gas is cooled by exchange to 400°F; the waste heat is recovered for steam generation. The saturated gas is further scrubbed with water for soot removal.

Water-Gas Shift. The CO/H_2 mole ratio of the gasifier effluent is 1.14. In the shift reactor water reacts with CO until the CO/H_2 mole ratio is 0.5, the stoichiometric ratio of CO and H_2 in the methanol synthesis reactions. The gas entering the shift reactor contains sufficient water for the required conversion, hence steam is not injected into the reactor. The gas entering the shift is brought to temperature (644°F) by heat recovered from the reactor effluent gas.

Acid Gas Removal. The shifted gas is cooled to 100°F before it enters the Rectisol acid gas removal unit where hydrogen sulfide and carbon dioxide are removed by selective absorption and desorption in methanol at 40°F. Nitrogen, a byproduct of the oxygen plant, is used to strip the CO_2 from the H_2S.

Methanol Synthesis. Methanol is synthesized from CO and H_2 at 130°F and 975 psi via the ICI process. Heat recovered from the reactor effluent gas is used to warm the feed from 100°F. Since approximately six percent of the reactant gases are converted during its residence in the catalyst bed, a significant recycle stream is required. A bleedstream from the recycle controls the accumulation of inerts within the loop. This vent stream is fired in the coal-slurry furnace. The methanol produced contains 0.2% water and 3.2% higher alcohols, which can be removed by distillation.

Oxygen Plant. Liquefied oxygen is produced by cryogenic separation from air. Liquid oxygen is pumped to 1150 psi, vaporized and preheated to 80°F.

Sulfur Plant. Sulfur is produced by reduction from hydrogen sulfide recovered by the Rectisol unit. Sulfur is produced by the Claus process.

A portion of the H_2S is oxidized to SO_2 so that the H_2S/SO_2 mole ratio at the reactor inlet is maintained at 2:1 to maximize sulfur production.

Heat Recovery/Steam Generation. Heat is recovered from the product gases at three locations: after the gasifier quench, after the water-gas shift, and after the methanol synthesis.

Heat recovered from the post-shift gases is utilized to preheat the coal slurry prior to its entrance to a fire heater. The remainder of the post-shift heat is used to heat the oxygen to 650°F prior to injection into the gasifier. Heat from the post-quench gases is recovered at two temperature levels. High pressure steam (600 psi) is generated by the heat recovered at the highest temperature level. Heat recovered by the subsequent condensation of the water vapor contained in the post-quench gases is used to generate medium pressure steam (250 psi) and to preheat feedwater to the high pressure boiler. The condensate from the post-shift gases and from the post-quench gases is used for slurrying the coal fed to the gasifier.

Heat recovered from the post-methanol synthesis gases is used to preheat boiler feedwater for both the high and medium pressure boilers. The remaining heat is removed by cooling water.

A coal-fired boiler is required to superheat the high pressure steam by 650°F and to produce additional medium pressure steam. The high pressure, superheated steam is delivered to the turbines which drive compressors in the oxygen plant. The medium pressure steam is delivered to various locations for turbine drive, heating and other process requirements.

Material balances and utility summaries for the production of 5,200 tons per day of methanol from Illinois No. 6 coal and from New Mexico coal are presented in Tables A-2-11 through A-2-14. The thermal efficiency based on the higher heating value of the methyl fuel was determined to be 52.3% for the Illinois coal and 55.4% for the New Mexico coal. This difference results from the fact that more heat can be recovered from the gasified New Mexico coal at a higher temperature because of the higher partial pressure of water in the gasifier effluent. The major system component sizes for both methanol-from-coal plants are presented in

TABLE A-2-11

METHANOL FROM COAL (TEXACO GASIFICATION) - ILLINOIS NO. 6 COAL
MASS BALANCE

STREAM NO.	1		2		3		4		5	
	Coal		Water		Oxygen		Inlet to Shift		Inlet to Synthesis	
TEMP (°F)										
GPM/ACFM										
Component Flows	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr
CO							23,733	664,513	14,875	416,490
CO_2							9,721	427,744	--	--
CH_4							446	7,131	446	7,131
H_2							20,864	41,727	29,749	59,499
H_2S							724	24,624	--	--
N_2					84	2,340	334	9,359	195	5,460
O_2					16,019	512,600				
MAF Coal	--	520,000								
Ash	--	58,000								
Water	1,389	25,000	25,944	467,000			17,159	308,858		
Methanol										
Total		603,000	25,944	467,000	16,130	514,940	72,981	1,483,956	45,265	488,580

TABLE A-2-11 Concluded

METHANOL FROM COAL (TEXACO GASIFICATION) - ILLINOIS NO. 6 COAL
MASS BALANCE

STREAM NO.	6		7		8		9			
	Methanol Product		CO_2 Vent		Sulfur		Synthesis Vent			
TEMP (°F)										
GPM/ACFM										
Component Flows	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr
CO							462	12,947		
CO_2			17,019	748,857			231	10,173		
CH_4							--			
H_2							919	1,838		
H_2S							--			
N_2			139	3,900			195	5,560		
O_2										
MAF Coal										
Ash										
Water	56	1,010					3	60		
CH_3OH	13,135	420,325					14	447		
Higher Alcohols	254	13,475								
Sulfur					699	22,360				
Total	13,445	434,810	17,158	752,757	699	22,360	1,824	31,025		

TABLE A-2-12

METHANOL FROM COAL (TEXACO GASIFICATION) - NEW MEXICO COAL
MASS BALANCE

STREAM NO.	1		2		3		4		5	
	Coal		Water		Oxygen		Inlet to Shift		Inlet to Synthesis	
TEMP (°F)										
GPM/ACFM										
Component Flows	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr
CO							20,897	585,120	14,090	394,520
CO_2							13,351	584,444	--	--
CH_4							158	2,528	211	3,376
H_2							21,346	42,692	28,179	56,358
H_2S							185	6,290	--	--
N_2					87	2,444	396	11,088	396	11,088
O_2					16,650	532,800	--			
MAF Coal		568,800								
Ash		169,800								
Water		48,200	28,920	520,600			13,667	246,006	--	--
Methanol										
Total		786,800	28,920	520,600	16,737	535,244	70,000	1,481,168	42,876	465,342

TABLE A-2-12 Concluded

METHANOL FROM COAL (TEXACO GASIFICATION) - NEW MEXICO COAL
MASS BALANCE

STREAM NO.	6		7		8		9			
	Methanol Product		CO_2 Vent		Sulfur		Synthesis Vent			
TEMP (°F)										
GPM/ACFM										
Component Flows	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr	#mols/hr	#/hr
CO							462	12,947		
CO_2			20,158	886,952			231	10,173		
CH_4										
H_2							919	1,838		
H_2S										
N_2			198	5,544			198	5,544		
O_2										
MAF Coal										
Ash										
Water	56	1,010								
CH_3OH	13,135	420,325					3	60		
Higher Alcohols	254	13,475					14	447		
Sulfur					184	5,900				
Total	13,445	434,810	20,356	892,496	184	5,900	1,827	31,009		

TABLE A-2-13

METHANOL FROM COAL (TEXACO GASIFICATION) - ILLINOIS NO. 6 COAL

UTILITY SUMMARY

Plant Section		Coal Grinding and Slurry Preparation	Gasifier	Post Quench Heat Recovery	Oxygen Plant	Post Shift Heat Recovery	Rectisol	MeOH Synthesis	MeOH Product Heat Recovery	Cooling Water	Steam Plant*	Net Imported Utilities
ELECTRICITY	P											
Kw	C	4,190										4,190
FUEL	P											
Coal, klb/hr	C										63	
STEAM												
600 psi, 650°F	P			1,465†							†	
K lbs	C				1,465							
200 psi, sat	P			202							255	
K lbs	C	44	148				216	35			14	
65 psi, sat	P					82						
K lbs	C						82					
	P											
	C											
COOLING WATER	P											
GPM	C	5,108	17,181		169,695		25,514	31,900			1,625	
RAW WATER	P											
GPM	C	606									12,551	13,157
	P											
	C											
	P											
	C											

P = Produced

C = Consumed

*includes coal-fired heater

†coal-fired heater provides superheat only

TABLE A-2-14

METHANOL FROM COAL (TEXACO GASIFICATION) - NEW MEXICO COAL

UTILITY SUMMARY

Plant Section		Coal Grinding and Slurry Preparation	Gasifier	Post Quench Heat Recovery	Oxygen Plant	Post Shift Heat Recovery	Rectisol	MeOH Synthesis	MeOH Product Heat Recovery	Cooling Water	Steam Plant*	Net Imported Utilities
ELECTRICITY	P											
Kw	C	5,430										5,430
FUEL	P											
Coal, klb/hr	C										77.4	
STEAM												
600 psi, 65°F	P			1,528†							†	
K lbs	C				1,528							
250 psi, sat	P			256							93	
K lbs	C	57	170				69	35			18	
65 psi, sat	P					22						
K lbs	C						22					
	P											
	C											
COOLING WATER	P									231,790		
GPM	C	4,902	14,619		178,071		6,048	26,603			1,548	
RAW WATER	P											11,896
GPM	C	306								11,590		
	P											
	C											
	P											
	C											

P = Produced

C = Consumed

*includes coal-fired heater

†coal-fired heater provides superheat only

Tables A-2-15 and A-2-16. Using New Mexico coal requires slightly larger capacities for the coal handling, gasification, and oxygen production sectors of the plant. However, significantly smaller capacities for the sulfur related processes are required for the New Mexico case since the New Mexico coal contains only 0.8 pounds of sulfur per million Btu's while the Illinois coal contains 3.0 pounds of sulfur per million Btu's.

The estimated capital investment in 1975 dollars is shown in Tables A-2-17 and A-2-18 for the Illinois and New Mexico coals, respectively. The production cost for methanol produced with Illinois coal is $5.36/M̄Btu's and is $5.52/M̄Btu's when produced from New Mexico coal. The detailed production cost breakdown is summarized in Tables A-2-19 and A-2-20.

The capital and production costs reflect the basic differences between the two cases--that of gasifying a high Btu, high sulfur Eastern coal and a low Btu, low sulfur Western coal.

TABLE A-2-15

METHANOL FROM COAL
MAJOR SYSTEM COMPONENTS

Illinois No. 6 Coal

ITEM NO.	Name	Size	Comment
Process Units			
P-1	Coal Preparation	603 Klb/hr Coal (A.R.)	A. D. Little Estimate
P-2	Gasification*	664 $\overline{\text{M}}$Scfd Gas Product	A. D. Little Estimate
P-3	Rectisol	724 lb mol/hr H_2S	A. D. Little Estimate
P-4	Oxygen Plant	6150 t/d Oxygen	Lotepro Estimate
P-5	Sulfur Plant	240 LT/d Sulfur	A. D. Little Estimate
Reactors			
R-1	Water-Gas Shift	37% CO Conversion	A. D. Little Estimate
R-2	Methanol Synthesis	5200 t/d Methanol	Bureau of Mines
Fired-Heaters			
F-1	Gas-Fired Furnace (Coal/water slurry preheat)	200 $\overline{\text{M}}$Btu/hr	--
F-2	Coal-Fired Boiler/Superheater	900 $\overline{\text{M}}$Btu/hr	--
Heat Exchangers			
E-1	Gas Cooler/Waste Heat Boiler (600 psi)	99,000 ft^2	\$1.30/ft^2
E-2	Gas Vapor Condensor/Waste Heat Boiler (250 psi)	16,100 ft^2	\$9.60/ft^2
E-3	Gas Vapor Condensor/Boiler Feedwater Preheat	138,000 ft^2	\$9.60/ft^2
E-4	Gas Cooler/Cooling Water	22,000 ft^2	\$9.60/ft^2

---

*Includes coal slurry storage, handling and preheat; ash handling; gas cleaning.

TABLE A-2-16

METHANOL FROM COAL
MAJOR SYSTEM COMPONENTS

New Mexico Coal

ITEM NO.	Name	Size	Comment
Process Units			
P-1	Coal Preparation	787 Klb/hr Coal (A.R.)	A.D. Little Estimate
P-2	Gasification*	800 MScfd Gas Product	A.D. Little Estimate
P-3	Rectisol	185 lb mol/hr H_2S	A.D. Little Estimate
P-4	Oxygen Plant	6394 t/d Oxygen	Lotèpro Estimate
P-5	Sulfur Plant	60 Lt/d Sulfur	A.D. Little Estimate
Reactors			
R-1	Water-Gas Shift	33% CO Conversion	A.D. Little Estimate
R-2	Methanol Synthesis	5200 t/d Methanol	Bureau of Mines
Fired-Heaters			
F-1	Gas-Fired Furnace (Coal/water slurry preheat)	200 $\overline{\text{M}}$Btu/hr	--
F-2	Coal-Fired Boiler/Superheater	900 $\overline{\text{M}}$Btu/hr	--
Heat Exchangers			
E-1	Gas Cooler/Waste Heat Boiler (600 psi)	92,000 ft^2	\$1.30/ft^2
E-2	Gas Vapor Condensor/Waste Heat Boiler (250 psi)	12,800 ft^2	\$9.60/ft^2
E-3	Gas Vapor Condensor/Boiler Feedwater Preheat	78,800 ft^2	\$9.60/ft^2
E-4	Gas Cooler/Boiler Feedwater Preheat	140,150 ft^2	\$1.30/ft^2

---

*Includes coal slurry storage, handling and preheat; ash handling; gas cleaning.

TABLE A-2-17

EPRI FUEL ASSESSMENT
CONVERSION SYSTEM CAPITAL INVESTMENT

Process Methanol - Illinois Coal Product Methyl Fuel
Daily Capacity: Feed 7240 t/d Product 5200 t/d

CATEGORY	$\$10^6$ (mid-1975)
INSTALLED PROCESS UNITS	
Coal Preparation	4.5
Gasification*	49.3
Water-Gas Shift	1.0
Acid Gas RemovaL	23.8
Sulfur Recovery	3.2
Oxygen Plant	50.3
Gas Cooling/Steam Generation	28.1
Methanol Synthesis	23.2
Total Process Plant	183.4
OFFSITES	
- Storage (30 days)	3.6
- General Facilities & Utilities	36.7
Total Direct Plant	223.7
Contingency @ 15%	33.6
Total Fixed Capital	257.3
Interest During Construction @ 15%[†]	38.6
Owner's Cost @ 3%	7.7
Total Depreciable Capital	303.6
Start-Up Cost[§]	23.0
Working Capital[¶]	15.2
TOTAL CAPITAL INVESTMENT	341.8

---

*Price includes coal slurry storage and handling, ash dewatering, carbon recycle system and coal slurry preheat furnace.

[†] 4 year construction period, 8% interest.

[§] 20% of Annual Gross Operating Cost.

[¶] 5% of Depreciable Capital

TABLE A-2-18

EPRI FUEL ASSESSMENT
CONVERSION SYSTEM CAPITAL INVESTMENT

Process Methanol - New Mexico Coal Product Methyl Fuel

Daily Capacity: Feed 9442 t/d Product 5200 t/d

CATEGORY	$\$10^6$ (mid-1975)
INSTALLED PROCESS UNITS	
Coal Preparation	5.3
Gasification*	55.5
Water	1.0
Acid Gas Removal	9.9
Sulfur Recovery	1.3
Oxygen Plant	51.6
Gas Cooling/Steam Generation	25.6
Methanol Synthesis	23.2
Total Process Plant	173.4
OFFSITES	
- Storage (30 days)	3.6
- General Facilities & Utilities	34.3
Total Direct Plant	211.3
Contingency @ 15%	31.7
Total Fixed Capital	243.0
Interest During Construction @ 15%[†]	36.5
Owner's Cost @ 3%	7.3
Total Depreciable Capital	286.8
Start-Up Cost[§]	24.2
Working Capital[¶]	14.3
TOTAL CAPITAL INVESTMENT	325.3

---

*Price includes coal slurry storage and handling, ash dewatering, carbon recycle system and coal slurry preheat furnace.

[†] 4 year construction period, 8% interest.

[§] 20% of Annual Gross Operating Cost.

[¶] 5% of Depreciable Capital.

TABLE A-2-19

METHANOL FROM COAL
OPERATING COST SUMMARY

Process Methanol, Illinois Coal Product Methyl Fuel

Daily Capacity: Feed 7240 t/d Product 5200 t/d

Capital Investment: Total $341.8 M̄ Depreciable $303.6 M̄

Stream Factor: 8000 hr/yr Depreciation Period: ____ Cost Basis: 1975

Category	Unit Consumed/Yr	$/Unit	Annual Cost $\$10^3$
VARIABLE COST			
Coal (tons)	2664 K	26.95	71,795
Purchased Water (kgal)	6024 K	0.40	2,408
Purchased Electricity (kwh)	33.5 M̄	0.018	603
Catalysts & Chemicals (lb)	--	--	--
Ash Disposal (dry tons)	232 K	3.00	696
Total Variable			75,502
SEMI-VARIABLE COST			
Operating Labor (person-hours)	1992 K	9.10	9,027
Maintenance Labor (60% of L&M)	--	--	6,175
Supervision (20% of O&M Labor)	--	--	3,040
Total Labor	--	--	18,242
G and A (60% of Total Labor)			10,945
Maintenance Materials (40% of L&M)			4,116
Total Semi-Variable Cost			33,303
FIXED COST			
Taxes & Insurance @ 2% of Total Depreciable Capital			6,072
Total Gross Operating Cost			114,877
Byproduct Credits			
Sulfur (LT)	85 K	45.00	(3,825)
Total Byproduct Credit			(3,825)
TOTAL NET OPERATING COST			111,052
CAPITAL CHARGES			
20.5% of Total Investment			70,070
TOTAL ANNUAL COST			181,122
Unit Cost, $/M̄Btu			5.36

TABLE A-2-20

METHANOL FROM COAL
OPERATING COST SUMMARY

Process Methanol, New Mexico Coal — Product Methyl Fuel
Daily Capacity: Feed 9442 t/d — Product 5200 t/d
Capital Investment: Total $325.3 M̄ — Depreciable $286.8 M̄
Stream Factor: 8000 hr/yr — Depreciation Period: ____ — Cost Basis: 1975

	Unit Consumed/Yr	$/Unit	Annual Cost $$10^3$
VARIABLE COST			
Coal (tons)	3457 K	22.53	77,886
Purchased Water (kgal)	5710 K	0.40	2,284
Purchased Electricity (kwh)	43.4 M̄	.015	652
Catalysts & Chemicals (lb)	--	--	--
Ash Disposal (dry tons)	679 K	3.00	2,037
Total Variable			82,859
SEMI-VARIABLE COST			
Operating Labor (person-hours)	992 K	9.10	9,027
Maintenance Labor (60% of L&M)	--	--	5,780
Supervision (20% of O&M Labor)			2,961
Total Labor			17,768
G and A (60% of Total Labor)			10,661
Maintenance Materials (40% of L&M)			3,853
Total Semi-Variable Cost			32,282
FIXED COST			
Taxes & Insurance @ 2% of Total Depreciable Capital			5,736
Total Gross Operating Cost			120,877
Byproduct Credits			
Sulfur (LT)	20 K	45.00	(900)
Total Byproduct Credit			(900)
TOTAL NET OPERATING COST			119,977
CAPITAL CHARGES			
20.5% of Total Investment			66,687
TOTAL ANNUAL COST			186,664
Unit Cost, $/M̄Btu			5.52

3.0 IMPORTED LNG AND NATURAL GAS

3.1 GAS SUPPLY AND DEMAND

The U.S. faces a major gas supply problem. Since 1954, the FPC has maintained the wellhead price of interstate natural gas below the prices of competing fuels, which tended to encourage rapidly growing consumption of gas, often in relatively low value end uses. At the same time incentives for new gas exploration and development were reduced.

Domestic production peaked in 1972, and has declined every year since then. Even with the higher $1.42/MCF wellhead price for new interstate gas, and the unregulated prices in the intrastate market, natural gas production from the lower 48 states is expected to decline continuously from 19.6 TCF in 1976 to 14 TCF in 1990. Proven reserves at January 1, 1977, totalling 216 TCF, are expected to decline at around 9% p.a., while new reserve additions (which have averaged around 9 TCF per year over the past 10 years -- excluding the North Slope reserves) are projected at 10 TCF per year in the absence of major new exploration incentives. By 1990, gas production from existing fields (excluding the North Slope) will be about 6 TCF, with 8 TCF of production resulting from new gas discoveries, plus revisions and extensions to existing fields. A large proportion of this new gas production could come from offshore fields in the Gulf of Mexico and other OCS areas (providing accelerated leasing of new permits is granted by regulatory agencies).

Table A-3-1 shows that in order to offset declining reserves in the lower 48 the U.S. will increasingly turn to higher cost alternatives. These include Alaskan gas, LNG imports, coal gasification and SNG from light hydrocarbons. The timing and quantities of these alternative sources depends in part on government policy. For example, LNG imports could be increased (with a corresponding reduction in oil imports, provided this were deemed to be in the national interest from a cost and security point of view.

TABLE A-3-1

PROJECTED U.S. SUPPLY OF NATURAL GAS AND GAS SUPPLEMENTS

(Trillion Cubic Feet)

	1975	1980	1985	1990
Lower 48 Production*	19.91	18.0	16.0	14.0
Alaskan Production	-	-	0.7	1.0
Pipeline Imports	0.95	0.9	1.3	1.3
LNG Imports	-	0.6	1.1	1.6
Coal Gasification	-	-	0.1	0.5
Light Hydrocarbon SNG	0.20	0.3	0.4	0.4
Total Supply	21.06	19.8	19.6	18.8

---

*Marketed production -- gross withdrawals less gas used for repressuring and quantities vented and flared.

Source: A. D. Little Estimates

Since the early 1970's gas demand has been supply limited -- a situation highlighted by increasing curtailments of low priority gas users (principally large industrial and utility). Annual curtailments (reflecting potential for increased consumption) reported by the Federal Power Commission since 1970 are shown in Table A-3-2.

Supply will continue to limit natural gas demand in the future with the available supply allocated preferentially to high priority markets (domestic, commercial, critical industrial use, etc.) with any residual gas availability being burned in the industrial market. Electric power generation is a special case with much of current consumption being used in the intrastate market by utilities that face difficulties converting to other fuels. This low priority use, however, is projected to be phased out by 1990 except for minor use in gas turbine peaking units.

Table A-3-3 shows projected demand by end use against the available supply and shows the balance available for the industrial market.

If sufficient gas were available in the future at competitive prices, a large proportion of the low priority consumers currently being curtailed would use that gas supply. As one measure of the potential industrial market for natural gas we added the projected level of curtailments in 1976 (3.57 TCF) to actual industrial use (6.83 TCF) to arrive at a theoretical industrial demand of 10.4 TCF if supply had not been limiting. This theoretical industrial gas demand was then added to the non-industrial gas demand projection shown in Table A-3-3 to determine the total potential gas requirement shown in Table A-3-4.

Table A-3-4 indicates that industrial users will be subject to additional curtailments unless further gas supplements can be introduced into the market at acceptable cost. Potential curtailments in the 1980's are on the order of 5 TCF. Projected gas supply is always sufficient (in the period from 1970 to 1990) to supply high priority non-industrial uses of gas, as well as to provide more than adequate coverage for critical industrial uses of gas (fertilizers, heat treatment, metal finishing, baking, etc.), whose total consumption amounts to about 1.0 - 1.5 TCF. This effect is shown in Figure A-3-1.

TABLE A-3-2

NATURAL GAS CURTAILMENTS

(Trillion Cubic Feet)

YEAR (April-March)	CURTAILMENTS (TCF)
1970/71	0.1
1971/72	0.5
1972/73	1.1
1973/74	1.6
1974/75	2.0
1975/76	2.9
1976/77 (Provisional)	3.5-4.0

Source: A.D. Little Estimates

TABLE A-3-3

PROJECTED SUPPLY/DEMAND OUTLOOK FOR NATURAL GAS

(Trillion Cubic Feet)

	Actual		Forecast		
	1970	1976	1980	1985	1990
Supply					
Marketed Production					
-- Lower 48	21.92	19.80	18.0	16.0	14.0
Supplements	0.82	0.97	1.8	3.6	4.8
Total	22.74	20.77	19.8	19.6	18.8
Demand					
Residential/Commercial	6.90	7.71	8.3	9.2	10.2
Electric Power	3.89	3.07	2.7	1.5	0.3
Others	0.34	0.24	0.2	0.2	0.2
Extraction Loss	0.91	0.80	0.8	0.8	0.8
Lease & Plant Fuel	1.40	1.38	1.2	1.0	1.0
Pipeline Fuel	0.72	0.57	0.6	0.6	0.6
Exports	0.07	0.07	0.1	0.1	0.1
Stock Changes	0.40	(0.17)	0.3	0.3	0.3
Unaccounted for	0.23	0.27	0.3	0.4	0.4
Subtotal	14.85	13.94	14.5	14.1	13.9
Balance Available for Industrial Use	7.89	6.83	5.3	5.5	4.9

TABLE A-3-4

POTENTIAL U.S. NATURAL GAS SUPPLY AND DEMAND

(Trillion Cubic Feet)

	1970	1976	1980	1985	1990
Gas Demand					
Non-Industrial	14.85	13.94	14.5	14.1	13.9
Industrial Plus Curtailments	7.99	10.40	10.4	10.4	10.4
Potential Gas Requirements	22.84	24.34	24.9	24.5	24.3
Gas Supply	22.74	20.77	19.8	19.6	18.8
Shortfall	0.10	3.57	5.1	4.9	5.5

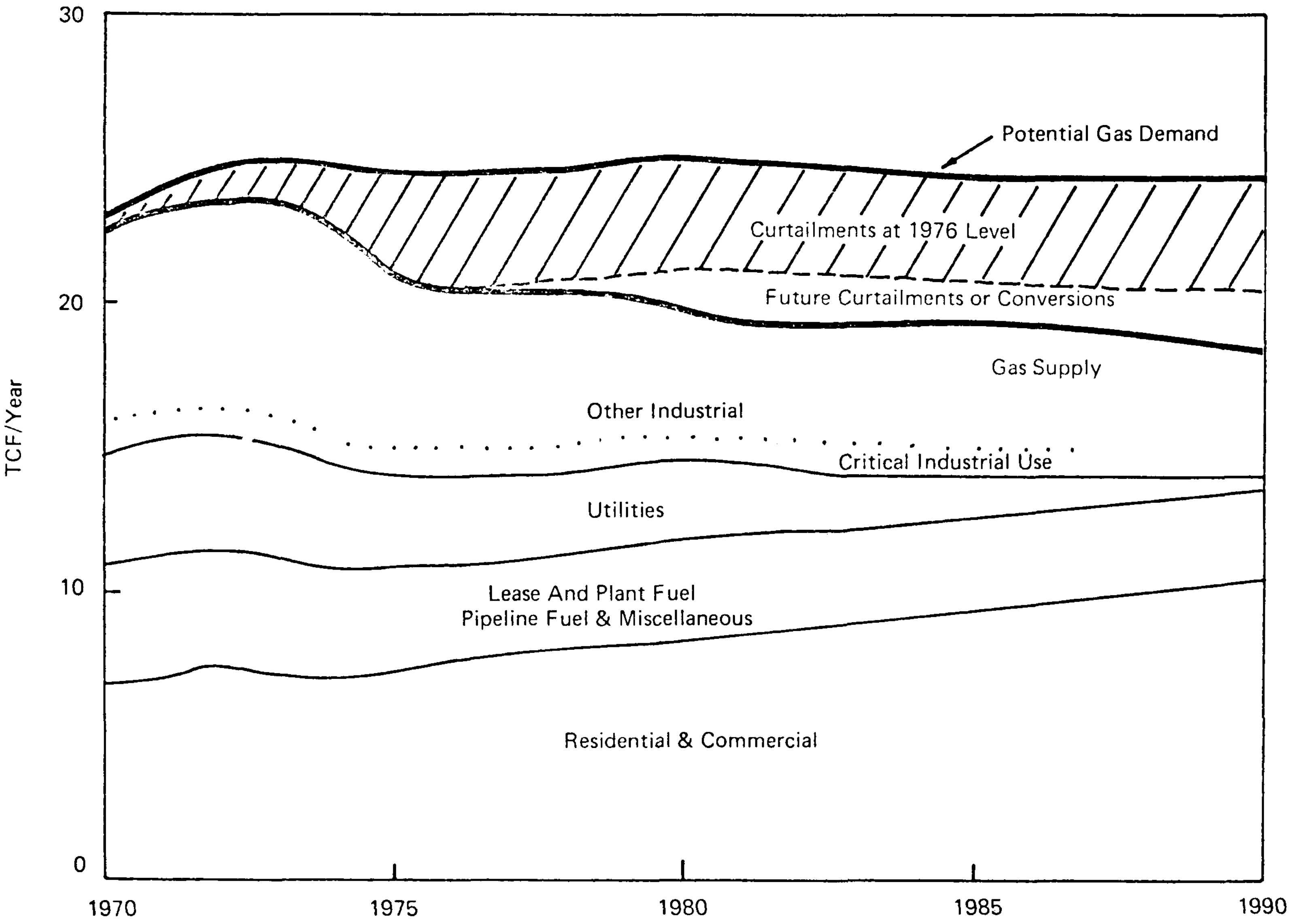

FIGURE A-3-1 U.S. NATURAL GAS SUPPLY AND DEMAND

3.2 DOMESTIC GAS SUPPLEMENTS

Domestic gas supplements include Alaskan gas, SNG from light hydrocargons, and coal gasification.

3.2.1 Alaskan Gas

Plans for developing the 22.5 TCF of proven Alaskan North Slope associated-gas reserves finally appears to be crystallizing. Of the three proposals for moving North Slope gas to the U.S., that which has gained both U.S. and Canadian favor is the Alcan Project which involves a 48" pipeline paralleling the Alyeska pipeline as far as the southern extent of the Brooks Range. There it would branch off following the Alcan Highway to Alberta, where extensions would move the gas again to the U.S. West Coast and Midwest. This project could be in operation by 1982, and could deliver from 2.4 to 3.4 BCF/d. Our forecast assumes a production level of 2 BCF/d in 1985, rising to 3 BCF/d by 1990. Current cost estimates for the Alcan pipeline and associated facilities are about $13 billion, which would infer a delivered cost at the Canadian border of at least $3.00-3.50/M̄Btu.

No allowance has been made in this forecast for gas supplies from Southern Alaska to the U.S. West Coast via the Pacific Alaska Project. The viability of the project (amounting to some 200 M̄CF/d in the initial stages) is dependent on further gas discoveries in Southern Alaska, and thus is not considered to have a high probability of success at this point.

3.2.2 SNG From Light Hydrocarbons

Currently, 13 light hydrocarbon SNG plants (based on naphtha or LPG) are in operation in the U.S. They are all located (with one exception) in New England, the Mid-Atlantic and East North Central regions. Total send-out capacity is of the order 1.3 BCF/d.

A further 12 SNG plants totalling nearly 1.5 BCF/d in send-out capacity, are awaiting approvals. Construction of further light hydrocarbon SNG plants has been held up partly because of the pressure from petrochemical producers anxious to retain preferential access to allocations of secure supplies of feedstocks, and partly by the FEA which currently regards naphtha and NGL-based SNG plants as being a thermally inefficient mode of supplemental energy supply. President Carter, in his Energy Message, however has indicated that greater emphasis would be placed on SNG developments, without indicating any preference for individual feedstocks. Our gas projection is based on SNG from light hydrocarbons contributing on average 0.4 trillion cubic feet of gas in 1985 -- about twice the current send-out.

Because SNG from light hydrocarbons is not highly capital intensive, it lends itself to load balancing operations which will become increasingly important as gas consumption shifts to low load factor (highly seasonal) domestic and commercial uses and away from interruptible industrial usage. Here SNG competes with other load balancing options such as underground storage, propane/air, pipeline LNG, etc. SNG also has the advantages of being adaptable to individual local utility systems, relatively short lead times for construction, and less FPC regulatory problems if confined to intrastate use.

3.2.3 Coal Gasification

FPC certification is pending on six coal gasification plants having a total capacity of about 1.5 BCF/d. These plants face a host of regulatory, environmental, technical, and economic problems which are unlikely to be resolved early. Our gas supply projection is based on having one plant in operation by 1985 with accelerating development thereafter as other gas supply alternatives are exhausted or high in cost and as oil prices rise in real terms.

Coal gasification is considerably more capital intensive than SNG from light hydrocarbons and thus will be used for base-load gas supply. Coal gasification has the advantage of being completely based on indigenous energy and in this sense reduces U.S. dependency in foreign energy supplies.

3.3 GAS IMPORT SUPPLEMENTS

Imported gas supplements include pipeline gas from Canada and Mexico and LNG from Africa, South America and eventually the Middle East.

3.3.1 Pipeline Gas

Pipeline gas exports to the U.S. from Alberta and Saskatchewan are expected to continue in the near term at or around present levels. Until recently there were concerns in Ottowa about the declining reserves in the Prairie Provinces, which were expected to lead to a phasing-out of gas exports to the U.S. by around 1980. More recently, however, there has been an improvement in the reserve position in these areas (due to new discoveries) which is expected to enable a continuing, if declining, flow of gas exports to the U.S. In addition, plans are in hand for the long-range development of the 17 TCF of gas reserves in the Canadian Arctic Islands, either by means of a temporary 250 M̄CF/d LNG scheme, or via the Polar Gas pipeline project (which could move up to 4.5 BCF/d into the Trans-Canadian system). Development of the MacKenzie Valley gas reserves has been postponed due to this improvement in the Canadian domestic gas outlook, and has influenced Canadian thinking in favor of the Alcan Gas Project (which does not access the MacKenzie Valley gas fields) for moving North Slope gas through Canada to the U.S. A recently proposed 2 BCF/d pipeline project from the Reforma Fields of Tabasco and Chiapas states in Mexico to McAllen, Texas, could be brought on stream by the early 1980's. Our estimate assumes that slowly declining Canadian pipeline imports will be bolstered by 1 BCF/d of Mexican gas in 1985 and 2 BCF/d in 1980.

3.3.2 LNG Imports

At present, the only LNG imports to the U.S. come from Algeria under the Distrigas I contract (10.2 BCF in 1976). Six further LNG projects, of which five are from Algeria, are at advanced stages of the construction/approvals process. Table A-3-5 summarizes the status of these projects. In principle, given full FPC approval within the next year, all of these projects could be in operation by 1985. At the

TABLE A-3-5

STATUS OF NEGOTIATED U.S. LNG IMPORT PROJECTS

Source	Contract	LNG Volumes (BCF/Year)		FPC Approvals	Construction Status	Comments[3]
		Loaded	Delivered[1]			
Algeria	Distrigas II	43	39	Approved	Operational	Replaces equity Distrigas contracts with SONATRACH from Skikda.
Algeria	El Paso I	365	332	Approved	Under Construction	Start-up projected for January 1978 from Arzew I.
Algeria	Eascogas[4]	238	216	Provisional Approval	Under Construction	Start-up provisionally mid-1979 from Arzew II.
Algeria	Panhandle (Trunkline)	164	149	Provisional Approval	Under Construction	Start-up provisionally early 1981 from Arzew III
Algeria	El Paso II	365	332	Pending	–	–
Algeria	Tenneco	390	354[2]	Pending	Under Construction	Start-up provisionally early 1981 from Arzew III
Indonesia	Pacific Lighting	200	182	Pending	Under Construction	–
		1765	1604			

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[1] Allowing 10% for ocean boil-off, losses and regasification plant fuel.

[2] Excludes gas retained in Canada for Canadian Lowell Gas, but includes gas which Tenneco may sell in Canada (up to 40% of volume shown). Siting of the reception terminal in St. John, NB now appears unlikely. Alternative sites in New England are being investigated.

[3] Allocation of contracts to specific LNG plants in Algeria is provisional and may be subject to change.

[4] Algonquin Gas Transmission's withdrawal from the Eascogas consortium is likely to create further delays in the implementation of this project

present time, however, the only contracts which have cleared all the U.S. regulatory hurdles are the Distrigas and El Paso I contracts. The El Paso I contract will be serviced from the Arzew I LNG plant now nearing completion by Bechtel (as a substitute for Chemico). Start-up is currently estimated for December, 1977, with first tanker loading in January, 1978. The current Distrigas contract is being serviced from three LNG trains in operation in Skikda, while the somewhat expanded quantities provided in the Distrigas II deal will be obtained from Trains 4-6, which are currently under construction. (Due to the dismissal of Pritchard Rhodes as prime contractor, commissioning of two of these trains will be further delayed.)

The delay in obtaining FPC approvals of the remaining Algerian contracts, has also slowed construction of Arzew II and III by Kellogg-Pullman and Foster Wheeler. Given the LNG contracts SONATRACH has currently negotiated with U.S. and European buyers, at least 2.5 BCF/d of additional liquefaction capacity would be required, which may be located at Skikda, Arzew or possibly, Dellys. At this time no construction contracts have been let on this additional liquefaction capacity. In addition to delays in the FPC certification process and in construction in Algeria, there have also been delays experienced by the receiving terminals as a result of environmental opposition relating to siting. This has become a major problem in California and New England. The only advanced project other than those from Algeria is the Pacific Lighting contract for Indonesian gas. Implementation of this agreement has been delayed due to lack of FPC approvals and uncertainty over the siting of the reception terminal.

Our estimates of future LNG import levels assume that by 1980 the Distrigas, El Paso I and (perhaps optimistically) Eascogas contracts will be in operation, at a combined level of 0.6 TCF/year. By 1990, all currently pending LNG projects are projected to be in operation. No allowance has been made for LNG supplies from countries other than Algeria and Indonesia although there would be room for additional supplies of LNG to maintain U.S. gas supply even at current levels.

Because of the high cost of transporting LNG by sea, gas exporters located nearest to consuming markets have a considerable economic advantage. Figure A-3-2 shows an estimate of the supply position of the most significant "short haul" gas producers to assess their ability to provide additional volumes of gas to the U.S. market. This estimate was made by applying a 20:1 reserve life index to published gas reserve figures.

In Algeria, currently proven gas reserves of around 120 TCF (at a 20:1 R/P ratio) just support the maximum potential demand in 1990 placed by current gas pipeline and LNG export commitments as well as a rapid expansion of local market demand. Indications are that additional potential gas reserves in Hassi R'Mel and the Southern Sahara fields may raise total reserves to 150 TCF or higher. In spite of this, however, we do not believe that significant additional volumes over and above that now committed to the U.S. market will become available.

Of the remaining short haul gas producers, Nigeria shows the greatest potential as a supplier of gas to the U.S. Practically all gas currently produced in Nigeria is flared. Plans for two LNG export deals totalling 2 BCF/d (Shell/BP and ENI/Phillips) have been approved by the Nigerian Government. It is unlikely that these projects will come on stream before 1985. A major proportion of this gas could be directed toward the U.S. market. Even allowing for these two LNG projects, substantial additional gas supplies are theoretically available from Nigeria. This hypothetical representation probably overstates Nigeria's supply capability, since the majority of reserves are in widely dispersed associated gas fields. Not only would the gathering of this gas be costly, but underproduction of the oil fields would result in cutbacks in potential gas production.

Two other short haul gas producers, Venezuela and Libya, are also unable to realize their apparent gas production potential. Libya has stated that, as a matter of policy, no new LNG schemes will be undertaken. Gas will be reserved for local industrial and residential needs. The Exxon plant operating at Marsa el Brega is thus expected to continue

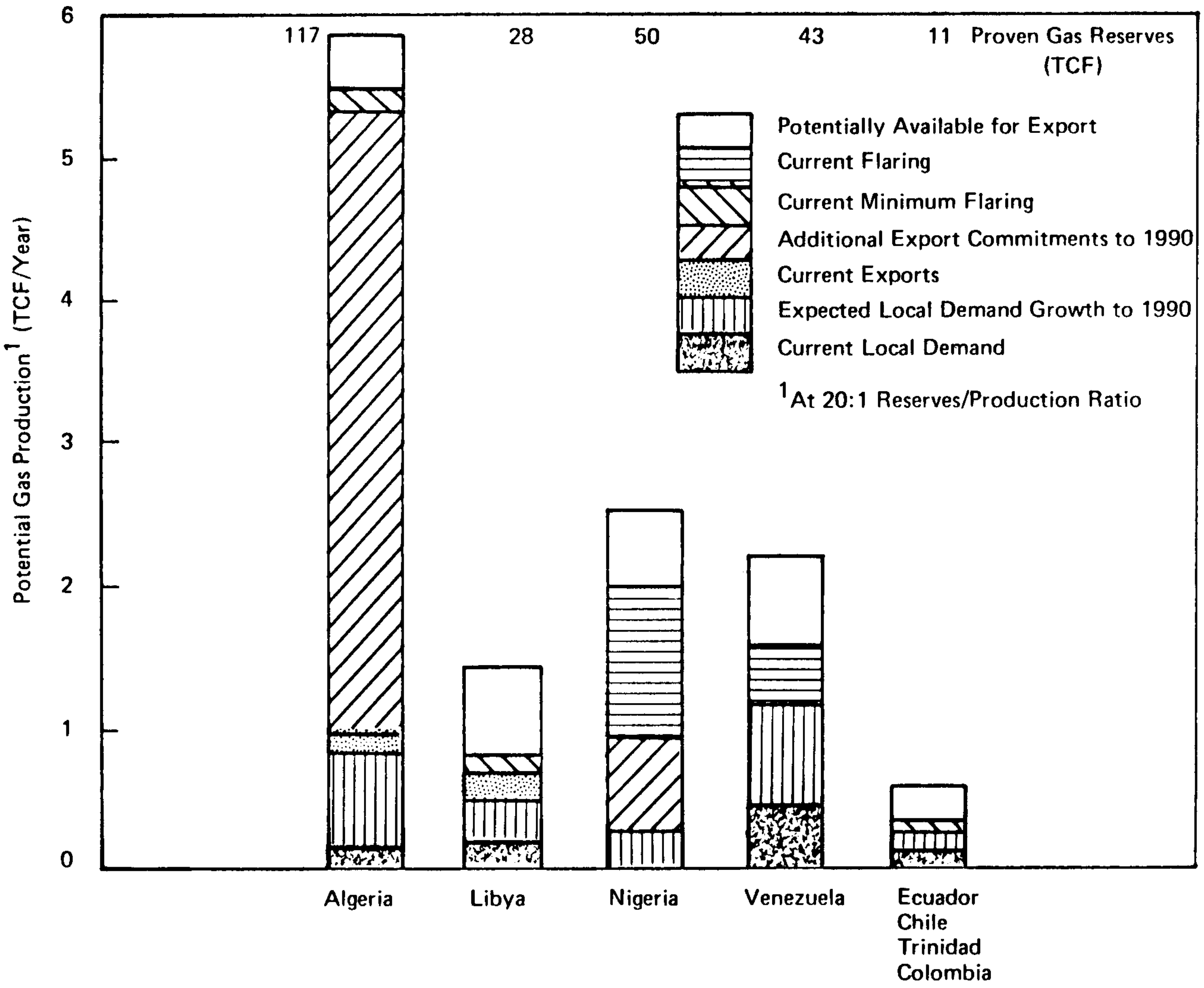


FIGURE A-3-2 NATURAL GAS SUPPLY POSITION OF POTENTIAL SHORT HAUL GAS PRODUCERS as the only LNG project in Libya. Venezuela had, at one time, considered the possibility of LNG exports from the El Tablazo complex in Western Venezuela. More recently, Venezuelan Government policy has shifted to a policy of gas conservation and, as in Libya, the country is anxious to preserve its gas reserves for internal needs.

Other potential Latin American gas exporters are:

- Chile (from ENAP's Magallanes fields in Tierra del Fuego)
- Trinidad (from Amoco's offshore Teak and Samaan fields)
- Ecuador (from CEPE's Oriente gas fields)
- Colombia (from Texaco's Guajira discovery)

None of these projects is expected to be capable of supplying significantly in excess of 250 M̄CF/d. As Figure A-3-2 shows, the combined export potential of these countries is relatively insignificant. In addition, government policy in Colombia and Trinidad may constrain gas exports.

In conclusion, the supplies of shorthaul gas which may be available to the U.S. before 1990 appear to be very limited for reasons of resource constraints, full commitment of production or government policy retaining gas reserves for local needs. Thus, if the U.S. is to import supplemental gas supplies over and above those already committed, it must increasingly look to supplies from more distant sources.

Figure A-3-3 shows that more than 3.5 TCF of uncommitted gas Projects directed to the U.S. market are currently under discussion. Of this total, 2.6 TCF would be supplied from distant sources (the Arabian Gulf, the Far East or the U.S.S.R.). The Siberian North Star and Yakutsk projects are the largest contracts pending (each at 2 BCF/d), but major political, technological, financial and economic barriers must be overcome before either of these projects can approach commercial viability. Even if these barriers were overcome in the relatively near future, gas supplies from the U.S.S.R. would probably not be available

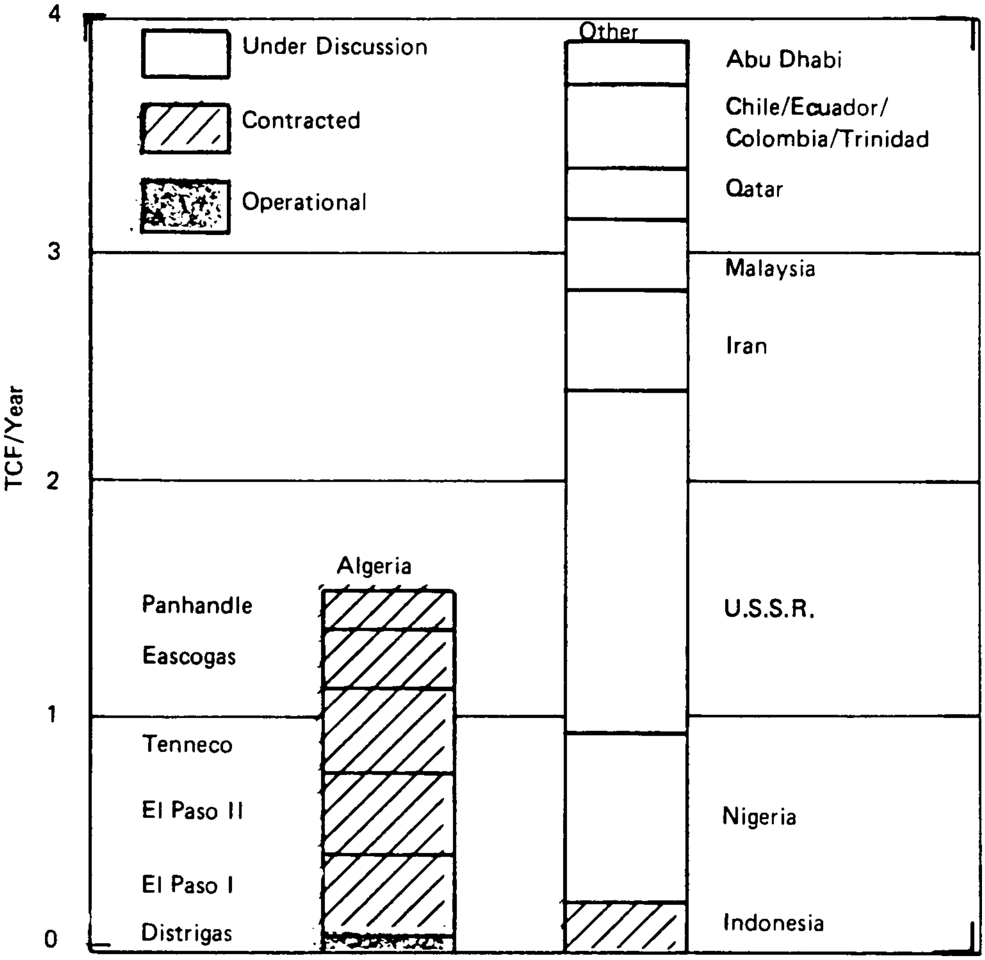


FIGURE A-3-3 POTENTIAL U.S. LNG IMPORT PROJECTS BY 1990 much before 1990. A further consideration is the apparent reduction in Eastern Bloc energy self-sufficiency in the late 1980's which may make the U.S.S.R. reluctant to export its gas in exchange for more costly oil imports.

After the U.S.S.R., Iran has the world's largest gas reserves (around 330 TCF). Associated gas from the onshore oil fields is used principally for pressure maintenance, although some is exported via the IGAT pipeline to the U.S.S.R. Pipeline exports of associated gas will increase when the trilateral Iran/U.S.S.R./Ruhrgas deal becomes operational. Major non-associated gas reserves are located in the Kangan structure (at least 20 TCF), and various consortia have been established in attempts to commercialize these resources. The most advanced of these projects is the Kalingas scheme which would move up to 1.6 BCF/d to Japanese and U.S. markets. However, project financing constraints have caused several changes in the participation structure of the consortium and further changes are probable. It is unlikely that any Iranian LNG will flow to the U.S. much before 1985 even if prompt regulatory approvals were granted by the FPC.

As yet, none of the U.S.S.R., Iranian or other long haul LNG projects have approached sufficiently close to contractual agreement to permit filing with the FPC. Though a potentially very large "overhang" of uncommitted LNG import projects may exist, it is quite possible that none of these may be in operation by 1990.

3.4 LNG REGULATORY FACTORS

Probably the most important element controlling the market penetration of LNG will be the regulatory environment. The impact of regulations will be felt both directly in terms of actions which will inhibit or facilitate the import of LNG or indirectly because of pricing standards or other measures which affect the availability and pricing of other fuels. Some of the principal regulatory factors impacting on LNG imports are described in more detail below.

3.4.1 Inter- and Intrastate Prices for Natural Gas

The Federal regulatory system involves maintaining the price of interstate gas at a value which is considerably lower than intrastate gas, which is unregulated. If President Carter's program leads to equalization of prices at $1.75/MCF the interstate pipelines may again be able to compete effectively for acquisition of new supplies (although there may still be some reluctance by producers to sell to interstate pipeliners if the prices are exactly equal). Nevertheless, the result may be greater supplies flowing to the more distant markets, particularly along the East and West Coasts, and thus some moderation of the need for supplemental supplies by distribution companies operating in these areas. Deregulation of natural gas prices would have an even more striking effect.

3.4.2 Terms of Purchase and Resale of LNG

Most of the existing contracts for import of LNG to the U.S. embody pricing provisions which were designed to allow its entry on a pricing basis essentially equivalent to fuel oil. The contract terms between Sonatrach and El Paso, originally negotiated in 1972, called for an FOB Price of 30.5¢ per MCF, and after adding shipping and terminalling costs, the delivered values are estimated to be about $1.03-$1.15 per MCF. In the El Paso II contract, however, as well as contracts with Panhandle Eastern, Distrigas, and Eascogas, the starting price was set at $1.30, and after shipping and regasification costs, the delivered values were targeted at $2.25/$\bar{\text{M}}$Btu -- judged to be roughly competitive with fuel oil. (Cost escalation and changes in the transportation agreements have since escalated ex-regasification plant costs above this value.) The $1.30 FOB price is tied to the reported values for fuel oil on the U.S. East Coast, or Algerian cost experience, whichever gives the higher value. Similar terms, with some minor modifications, were negotiated by Pertamina with Pacific Lighting, starting at a price of $1.25 per MCF, escalated with the value for crude oil and the wholesale energy price index. The use of a market-based indexation system was recently challenged by the FPC in its decision on the application of Trunkline Gas Company (Panhandle Eastern) for import of LNG from Algeria.

3.4.3 Incremental Versus Roll-In Pricing

Each of the applicants requesting approval of an LNG import project from the FPC has requested that it be allowed to roll in the price of regasified LNG with other sources of gas. While the FPC originally denied this request in the case of Columbia Gas at Cove Point, Maryland, its decision was overruled in the courts and a roll-in basis was allowed. In the recent Trunkline Gas decision, however, the FPC at first stipulated the use of incremental pricing, arguing that the higher price of delivered gas ($3.37 per MCF in 1980) as contrasted with the El Paso I project could mean a greater burden on the ratepayers, and incremental pricing would force users to make a decision as to the viability of the imports of higher-priced gas. Subsequently, on appeal from Trunkline which argued that incremental pricing would indeed force the cancellation of the project (which had at least $1 billion in committed or spent finances), the FPC withdrew its earlier decision and approved the project on the basis of roll-in pricing.

3.4.4 Imported Limits

During 1976, the Energy Resources Council, an interagency committee of organizations concerned with energy policy in the Federal Government, reviewed the question of possible import limits for LNG to avoid excess dependency on foreign sources. They recommended that the FPC consider authorization of projects only if they fell within a limit of one trillion cubic feet per year from any single country and two trillion cubic feet in total. This would amount to about 10% of the domestic supply, and was judged to be a reasonable exposure. Adoption of these limits would not have any significant effect on planned imports except in the case of Algeria where projects totalling about 1.4 trillion cubic feet were under negotiation. President Carter has indicated that he plans to review this import limitation, however, and would probably implement a policy with greater flexibility. It appears likely, therefore, that none of the presently envisioned projects will be inhibited by this limitation.

3.4.5 Terminal Approvals

One of the most significant delay factors in the implementation of LNG projects has been the process of securing approvals for import terminals. At present, the only two major import terminals are at Cove Point, Maryland, and Elba Island, Georgia, although there is a small terminal operating at Everett, Massachusetts, to receive Distrigas imports. Operation and/or construction of terminals at other sites, including Providence, Rhode Island, Staten Island, New York, Gloucester County, New Jersey and all of the sites in California, have been inhibited by various levels of governmental intervention. For some, concern has been expressed about the safety of ship movements and deliveries in congested areas, whereas for others opposition has arisen from environmental forces that do not want industrialization of pristine areas. Terminals on the Gulf Coast at Lake Charles, Louisiana, and Corpus Christi, Texas, appear subject to less opposition because of the general acceptability of energy-related activities in those areas, but considerable difficulty is foreseen in securing the necessary approvals for construction of terminals at other sites. Tenneco, for its project, plans to import LNG via Canada, with a new terminal at St. John, New Brunswick, although this site now appears to be doubtful.

3.4.6 Controls on End Use of Gas

Because of limitations in the supply of gas to the interstate market, the pipeline companies and, correspondingly, their customers have been forced to adopt curtailment schedules which eliminate given classes of customers from use of gas at times of peak demand in the winter. The FPC has promulgated a set of curtailment priorities which call for first eliminating large-volume users who employ gas as boiler fuel, followed by smaller-volume boiler fuel users, then by large users who employ gas for process purposes, and finally by residential and commercial users. Some pipeline systems, such as Transcontinental Gas Pipeline, were required to make 45-50% curtailments of their total deliveries during the recent cold-weather period, and virtually all large industrial users without alternative fuel capabilities were required to shut down. While service to such users has now been restored, many of these companies have taken steps to provide alternative fuel supplies, either in the form of LPG for those who need clean-burning fuels, or fuel oil for the others. One result has been that many of these companies are no longer basing their plans on continued access to natural gas, and are particularly indifferent to offers from importers who can provide natural gas service only on an incremental-pricing basis.

3.5 LNG COSTS

The potential for additional LNG imports from relative nearby sources is felt, in the longer term, to be limited. The U.S., if it wishes to import substantially greater LNG volumes than the 1.6 TCF shown in our forecast for 1990, must turn to more distant sources, such as the Middle East. In this section, therefore, we have developed the costs of both Algerian and Middle Eastern LNG supplies delivered to a reception terminal on the U.S. Gulf Coast. This LNG can then be moved through the interstate transmission system (either directly or by replacement) in order to supply users in Dallas, Columbus or Hartford.

There is much uncertainty over the likelihood of further LNG import projects for the West Coast, given the environmental opposition to terminal construction. Our analysis reviews the cost of LNG delivered from a Middle East project, as well as that from Indonesia under current LNG pricing terms.

3.5.1 Algerian LNG Contracts

All LNG contracts signed with U.S. companies since mid-1975 contain similar pricing provisions. These are briefly summarized below:

- A base FOB price of \$1.30/$\overline{\text{M}}$Btu;
- Indexation of the FOB price with a No. 2 and No. 6 fuel oil "market" price index;
- Pass-through of inflation construction and operating costs of into a "floor" price;

- A review of the price and escalation terms every four years;
- 20-year contract life.

A more detailed presentation of the indexation formulae is given in Table A-3-6. The purpose of the floor price is to protect Algerian investment in the liquefaction projects against inflation and currency variations, while the market-based formula is established to relate Algerian LNG values to those of competing fuels. Since these agreements have been signed, cost inflation in construction has probably exceeded fuel oil and distillate price increases such that during the early years of operation the FOB price will reflect the cost-based floor price rather than the oil market price index.

The only Algeria-U.S. LNG contract not to use these pricing terms is the El Paso I contract, which was originally signed in late 1969. This contract will be serviced by the Arzew I LNG plant due to commence operation in January 1978. The pricing formula used in this contract is considerably less favorable to Algeria than the more recent contracts because of:

- Lower FOB price (base price of only 30.5¢/$\overline{\text{M}}$Btu);
- Only 20% price indexation;
- Exclusion of oil product price equivalency.

We expect that the El Paso I pricing terms will be eventually renegotiated since the indexed price (using recent values for the index terms) would only be 37¢/$\overline{\text{M}}$Btu FOB Arzew. To sell at this price would result in substantial losses for SONATRACH, which has experienced major cost overruns and delays on the Arzew plant.

In determining future costs for Algerian LNG we have applied the projected values for No. 2 and No. 6 oil in Hartford (as calculated in the report section on oil products prices) to the indexation formula shown in Table A-3-6. These calculations are illustrated in Table A-3-7.

TABLE A-3-6

ALGERIAN LNG CONTRACT INDEXATION TERMS

(Excluding El Paso I)

1. Market Price

$$P_m = 1.30\ [0.5(\frac{G}{G_o}) + 0.5(\frac{F}{F_o})]$$

where P_m = Invoice Price (never less than 1.30 nor less than P_f) in $/MMBtu.

G = "No.2 Fuel Oil" price as calculated from the highest prices published by Platt's Oilgram under the heading "Atlantic and Gulf Coast, New York Harbor District" for each day during the six months ending one month prior to the beginning of the semester for which the price is computed.

G_o = U.S. $12.642/bbl

F = "No.6 Fuel Oil, low pour, max. 0.3% sulfur" price listed under "Atlantic and Gulf Coast, New York Harbor District, No.6 fuel rack" in Platt's Oilgram for each day . . .

F_o = U.S. $13.505/Bbl

2. Floor Price*

$$P_f = 1.30\ [A(\frac{C}{C_o}) + B(\frac{I}{I_o}) + W]\ (X)$$

where P_f = Floor Price (never less than 1.30)

$\frac{C}{C_o}$ = Escalation of capital costs of gas processing, pipeline and LNG plant in Algeria compared with mid-1975 estimate.

$\frac{I}{I_o}$ = Indexation of operating costs in accordance with a variety of U.S. Bureau of Labor statistics (mid-1975 base).

A,B,W = Proportions of the 1.30 Floor Price represented by capital items, operating costs and wellhead value for gas (confidential).

X = Currency adjustment factor (mid-1975 base) reflecting changes in the parity of a basket of six European currencies against the U.S. Dollar.

*This formula is explicitly contained only in the El Paso contract. The other U.S./Algeria LNG contracts contain a review clause which will be implemented on plant start-up to ensure that SONATRACH is adequately protected against investment cost inflation

TABLE A-3-7

PROJECTED COSTS OF ALGERIAN LNG ON U.S. GULF COAST

(1975 $/$\overline{\text{M}}$Btu)

	1976	1980	1985	1990
FOB Price	1.30	1.75	2.53	2.74
Transportation (Arzew - U.S. Gulf)	0.85	0.87	0.92	0.93
Regasification	0.27	0.27	0.27	0.27
Regasified LNG Value	2.42	2.89	3.72	3.94

3.5.2 Indonesian Contract

The Pacific Lighting contract signed in 1973, and revised in 1975 is somewhat different from the Algerian contracts. A brief summary of the terms is given below:

- A minimum FOB price of \$1.25/$\overline{\text{M}}$Btu;
- Escalation against Indonesia crude prices and a U.S. energy cost index;
- A most favored nation clause;
- Quarterly price determination.

Details of the price escalation formula are given in Table A-3-8.

Applying estimated international crude oil and domestic fuels escalation to the Indonesian contract, regasified LNG values obtained for the Pacific Lighting contract are shown in Table A-3-9.

3.5.3 Middle East LNG

Since no contracts for delivering Middle East LNG are either in operation or at an advanced stage of either construction or even negotiation, we have calculated the costs of a 1 BCF/d LNG project from the Persian Gulf via Suez to the U.S. Gulf Coast, and to the U.S. West Coast.

Salient aspects of the project cost are:

- A natural gas inlet price of 40¢/MCF in the Persian Gulf (in 1975 dollars);
- A liquefaction plant cost of \$1.3 billion in 1976 dollars for 1 BCF/d of LNG sendout (natural gas equivalent);
- Capital cost of \$120 million in 1976 dollars for each 125,000 M^3 LNG tanker used;
- Regasification terminal costs which are similar to those used for the Algerian and Indonesian ventures.

A summary of the projected costs for regasified LNG on both the U.S. Gulf and West Coasts is shown in Table A-3-10.

TABLE A-3-8

PACIFIC LIGHTING LNG CONTRACT INDEXATION TERMS

Pacific Lighting (Indonesia to U.S. West Coast)

$$P = 1.25\ [0.5\ (\frac{A}{A_o}) + 0.5\ (\frac{I}{I_o})]$$

where P = FOB sales price in $\$/\overline{\mathrm{M}}\mathrm{Btu}$

A = Volume weighted average price of all Indonesian FOB crude oils sold for export under long term contracts.

A_o = U.S. $\$11.0$/bbl

I = Latest value for Code 05 - "Wholesale Price Index for Fuels and Related Products and Power" published by the Bureau of Labor Statistics, U.S. Department of Labor.

I_o = 230.0

TABLE A-3-9

PROJECTED COSTS OF INDONESIAN LNG ON U.S. WEST COAST

(1975 $/$\overline{\text{M}}$Btu)

	1976	1980	1985	1990
FOB Price	1.30	1.56	2.13	2.28
Transportation (Indonesia - Pt. Conception)	1.35	1.38	1.45	1.47
Regasification	0.27	0.27	0.27	0.27
Regasified Value	2.92	3.21	3.85	4.02

TABLE A-3-10

LANDED COSTS OF MIDDLE EAST LNG

(1975 $/$\overline{\text{M}}$Btu

	Middle East to U.S. Gulf Coast	Middle East to U.S. West Coast
Inlet Gas Value	0.31	0.31
Liquefaction Cost	0.81	0.81
Transportation Cost	1.59	2.10
Regasification Cost	0.27	0.27
Regasified LNG Value	2.98	3.49

By 1985, the projected landed cost of Middle East LNG is close to that of Algerian and Indonesian supplies. As the price of imported crude oil forces up real oil products prices, the economic rent in LNG contracts will continue to grow. We expect that the majority of this economic rent will be captured (as it will be by the provisions of the Algerian and Indonesian contracts) by the producer country. Consequently, we have escalated the 1985 and 1990 FOB cost of Middle East LNG by the oil products price escalation on the U.S. East Coast between 1980 and 1990 as a measure of the increased economic rent captured by the Persian Gulf LNG producers under conditions of continued real growth in energy prices.

3.6 GAS TRANSMISSION AND DISTRIBUTION COSTS

In order to assess the costs of mainline gas transmission from the regasification plant to the city gate of the gas utility which would distribute the gas to its users, published tariffs offered by the major interstate gas transmission systems were examined. In addition, assessments of the costs of constructing and operating large (36-48") gas pipelines in the continental U.S. were made. These analyses yielded an estimate of close to 4¢/MCF (in 1976 dollars) for each 100 miles the gas is transmitted.

Gas distribution costs to large volume industrial and utility users vary widely in the U.S. depending on:

- Utility cost allocation practice and state regulations governing rate base cost pass-throughs;
- Proportion of interruptible to non-interruptible users;
- Gas penetration of the industrial energy market;
- City gate cost of gas.

Figure A-3-4 shows that gas distribution costs vary widely on a national basis. They are highest in New England which has a combination of the highest cost of gas and the lowest gas penetration of any major region in the U.S. Distribution and marketing costs are also highest (as may be expected) for residential consumers. Even within these regions there may be wide variations in effective distribution costs (defined as the gas utility tariff minus the city gate cost of gas acquired from the transmission utility). These are shown in Table A-3-11.

In order to generalize the analysis of gas distribution costs, data were obtained from the AGA publication "Gas Facts" for 1975, from which the average industrial gas distribution cost for the states in which the projected fuel cells would be located (Table A-3-12) was calculated. It should be stressed once again that these "distribution costs" do not, and cannot, necessarily represent the cost of gas distribution to a specific user, such as an electric utility.

To illustrate the projected cost of LNG supplies to the four specified fuel cell locations, it was assumed that this gas would be priced incrementally. In fact, further LNG import contracts may well obtain roll-in pricing approval from the FPC which would lower the delivered cost of gas to the utility. As further gas import contracts are approved, however, it is expected that roll-in pricing may become more and more difficult to obtain. It will become academic if new gas deregulation is eventually passed.

Tables A-3-13 and A-3-14 show the projected costs of imported LNG delivered to the electric utility fuel cell locations. It is assumed that deliveries under the specified contractual terms from the three different LNG sources would become available in 1976 for Algerian LNG, 1980 for Indonesian LNG and 1985 for Middle East LNG.

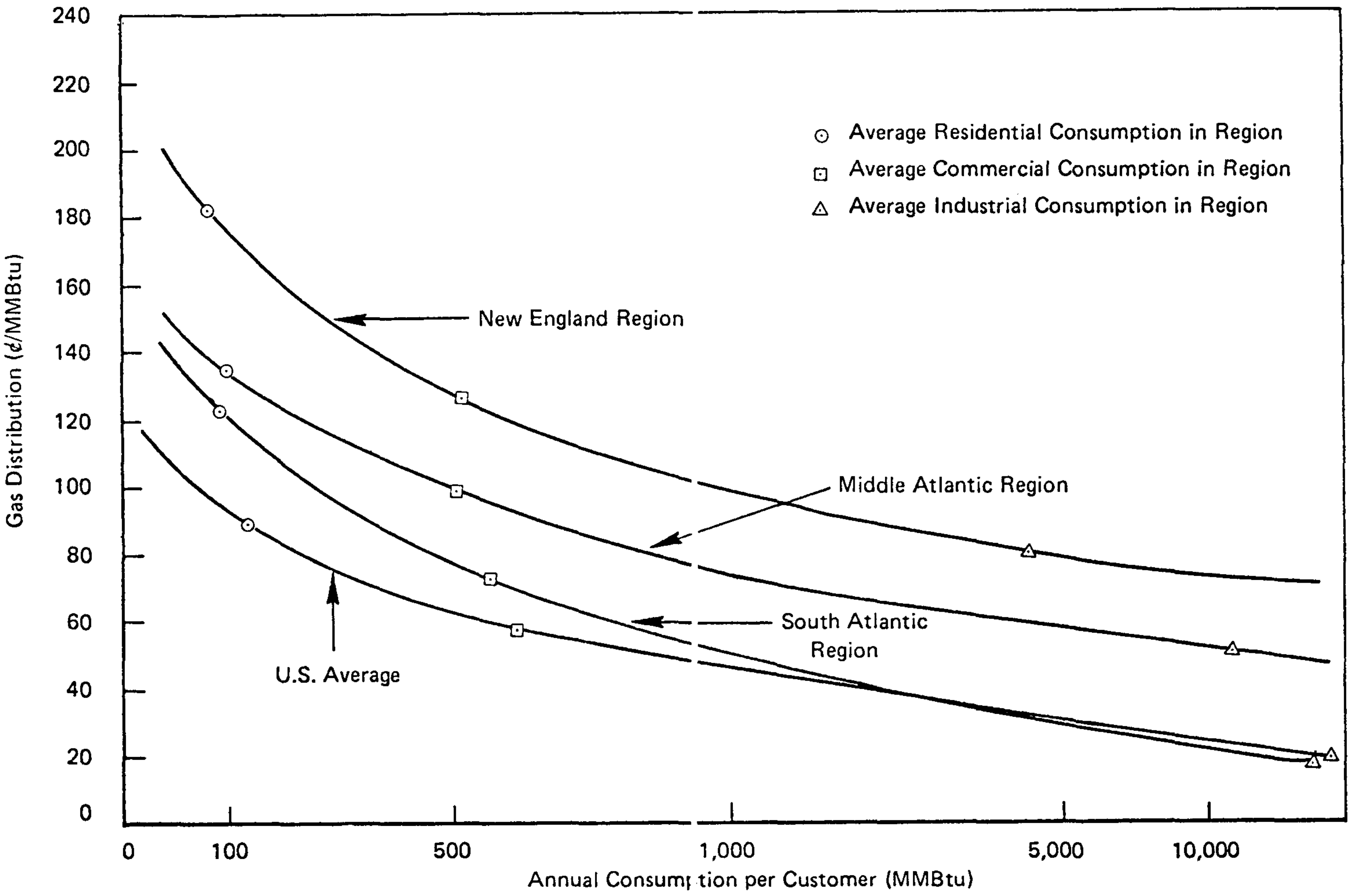


FIGURE A-3-4 U.S. GAS DISTRIBUTION COSTS BY CUSTOMER SIZE - 1975

TABLE A-3-11

GAS UTILITY DISTRIBUTION COSTS

(Tariff - $/MCF)

<table>
<tr><th>Utility</th><th>Industrial Firm†</th><th>Industrial Interruptible†</th><th>Average All Customers</th></tr>
<tr><td>Boston Gas</td><td>1.33*</td><td>0.47</td><td>1.67</td></tr>
<tr><td>So. Conn. Gas Co.</td><td>1.18</td><td>0.44</td><td>1.64</td></tr>
<tr><td>Brooklyn Union Gas</td><td>1.39</td><td>0.43</td><td>1.85</td></tr>
<tr><td>Consolidated Edison</td><td>1.69*</td><td>0.43</td><td>1.98</td></tr>
<tr><td>So. Jersey Industries</td><td colspan="2">0.53</td><td>0.92</td></tr>
<tr><td>W. Kentucky Gas Co.</td><td>0.22</td><td>0.18</td><td>0.35</td></tr>
<tr><td>Atlanta Gas Light Co.</td><td colspan="2">0.16</td><td>0.38</td></tr>
<tr><td>Alabama Gas Co.</td><td colspan="2">0.34*</td><td>0.41</td></tr>
<tr><td>Piedmond Nat'l Gas Co.</td><td>0.68</td><td>0.46</td><td>0.79</td></tr>
<tr><td>Public Service of NC</td><td>0.58</td><td>0.50</td><td>0.89</td></tr>
<tr><td>Michigan Consolidated Gas</td><td colspan="2">0.28</td><td>0.46</td></tr>
<tr><td>No. Illinois Gas Co.</td><td>0.33</td><td></td><td>0.53</td></tr>
<tr><td>Indiana Gas Co.</td><td colspan="2">0.23</td><td>0.54</td></tr>
<tr><td>Wisconsin Gas Co.</td><td colspan="2">0.19</td><td>0.48</td></tr>
<tr><td>Minnesota Gas Co.</td><td>0.47</td><td>0.17</td><td>0.61</td></tr>
<tr><td>Laclede Gas Co.</td><td>0.34</td><td>0.21</td><td>0.80</td></tr>
<tr><td>Lone Star Gas Co.</td><td colspan="2">0.22</td><td>0.26</td></tr>
<tr><td>So. California Gas Co.</td><td colspan="2">0.19</td><td>0.49</td></tr>
<tr><td>So. Union Gas Co.</td><td colspan="2">0.09</td><td>0.44</td></tr>
<tr><td>Pacific Gas & Elec. Co.</td><td colspan="2">0.28</td><td>0.43</td></tr>
<tr><td>Washington Natural Gas Co.</td><td>0.50</td><td>0.17</td><td>0.64</td></tr>
<tr><td>Intermountain Gas Co.</td><td>0.78</td><td>0.02</td><td>0.55</td></tr>
</table>

* Includes commercial

† Where individual figure is not given in each category, the figure shown is the average of firm and interruptible industrial.

TABLE A-3-12

INDUSTRIAL GAS DISTRIBUTION COSTS

(1975 $/MCF)

Fuel Cell Location	State	Industrial Gas Price	Cost of Gas	Industrial "Distribution Cost"
Hartford	Conn.	2.27	1.51	0.76
Columbus	Ohio	1.12	0.84	0.28
Dallas	Texas	1.05	0.64	0.41
Los Angeles	Calif.	1.04	0.85	0.19

TABLE A-3-13

DELIVERED COST OF GULF COAST LNG IMPORTS TO FUEL CELL LOCATIONS

(1975 $/MBtu)

	Algerian LNG				Middle East LNG	
	1976	1980	1985	1990	1985	1990
To U.S. Gulf Coast						
FOB Price	1.30	1.75	2.53	2.74	1.62	1.75
Transportation	0.85	0.87	0.92	0.93	1.69	1.72
Regasification	0.27	0.27	0.27	0.27	0.27	0.27
Regasified LNG Cost	2.42	2.89	3.72	3.94	3.58	3.74
Delivery to Hartford						
- Transmission	0.49	0.49	0.49	0.49	0.49	0.49
- Distribution	0.76	0.76	0.76	0.76	0.76	0.76
Delivered Cost	3.67	4.14	4.97	5.19	4.83	4.99
Delivery to Columbus						
- Transmission	0.38	0.38	0.38	0.38	0.38	0.38
- Distribution	0.28	0.28	0.28	0.28	0.28	0.28
Delivered Cost	3.08	3.55	4.38	4.60	4.24	4.40
Delivery to Dallas						
- Transmission	0.12	0.12	0.12	0.12	0.12	0.12
- Distribution	0.41	0.41	0.41	0.41	0.41	0.41
Delivered Cost	2.95	3.42	4.25	4.47	4.11	4.27

TABLE A-3-14

DELIVERED COST OF WEST COAST LNG IMPORTS TO FUEL CELL LOCATIONS

(1975 $/MBtu)

	Indonesian LNG			Middle East LNG	
	1980	1985	1990	1985	1990
To U.S. West Coast					
FOB Price	1.56	2.13	2.28	1.62	1.75
Transportation	1.38	1.45	1.47	2.23	2.27
Regasification	0.27	0.27	0.27	0.27	0.27
Regasified LNG Cost	3.21	3.85	4.02	4.12	4.29
Deliveries to Los Angeles					
- Transmission	0.06	0.06	0.06	0.06	0.06
- Distribution	0.19	0.19	0.19	0.19	0.19
Delivered Cost	3.46	4.10	4.27	4.37	4.54

4.0 LPG SUPPLY/DEMAND AND PRICES

4.1 SUPPLY/DEMAND

4.1.1 Domestic

The United States is by far the world's largest consumer of LPG and currently accounts for more than half the LPG consumed in the world outside the Eastern Socialist countries. The driving forces for the development of this very large market have been the presence of abundant natural gas liquids and the regulated higher prices available for liquid fuels compared to natural gas. LPG has therefore been extracted from natural gas and has been sold in competition with higher priced oil products. The low production cost has allowed LPG to overcome higher distribution costs and penetrate markets which in other developed countries have been monopolized by oil products.

Domestic natural gas production is now declining. In addition, the gap between gas and oil prices is narrowing. These two factors are likely to have a significant impact on the domestic LPG market in that:

- Domestic LPG production will stagnate or decline.
- Future growth in LPG demand will depend on imports, which is causing some concern on security of supplies.
- Ethane extraction will continue to grow and ethane will gradually substitute for propane as a petrochemical feedstock.
- Natural gas distribution systems will not expand significantly. Growth in propane use as an interim fuel in outer suburbs in anticipation of expansion of the gas grid will thereby suffer.
- Opportunities will emerge for propane sales into markets undergoing curtailments of natural gas supplies, particularly in industry.
- Further opportunities should begin to develop with gas utilities for manufacture of supplemental gas supplies through SNG, propane/air or Btu enrichment.

We have analyzed the prospects for growth of propane consumption in the different end-use sectors, based on the hypothesis that LPG prices at Mont Belvieu, Texas remain competitive with prices for No. 2 fuel on a $/$\overline{\text{M}}$Btu basis in the U.S. Gulf. This implies a significant premium for LPG over distillate at the burner tip, particularly in the Northeast and Midwest, since LPG pipelining, distribution and marketing costs are significantly higher than those for conventional petroleum fuels. The propane demand forecast is summarized in Table A-4-1.

In the domestic and commercial sectors LPG growth prospects are limited. LPG use is primarily a feature of rural areas and of outer suburbs which lie outside urban gas grids and where oil distribution networks do not exist. We do not expect any further penetration of LPG in the domestic sector with the consequence that demand will track growth in the stock of homes. However, one sub-sector where LPG consumption is growing rapidly is in mobile homes for which there are no satisfactory alternative fuels. Overall, however, we forecast low growth rates of the domestic and commercial sectors, averaging 2.8% p.a. in the late 1970's, falling to 2.0% in the 1980's. Due to the large base of demand in this sector, this will create by 1990 an additional 65 million barrels/year of demand.

In the industrial sector, many large users facing natural gas curtailments are considering converting their base-load requirements to No. 6 fuel oil, rather than to LPG, on grounds of cost. Some industries, particularly those with open-flame processes, require gaseous fuels. However, natural gas curtailments are most often only partial so that those processes requiring gaseous fuels can continue to operate on natural gas, while steam and power is obtained by burning liquid fuels. There is clearly great scope for large increases in LPG sales to the industrial sector at prices competitive with heavy fuel oils. However, as explained above, the base forecast is predicated on significantly higher propane prices at the burner tip. The principal opportunity for increasing LPG sales to the industrial sector thus lies in its ability to act as a standby fuel for curtailed industrial users, particularly in process-specific applications requiring clean fuels.

LPG is of interest to gas utilities for three different uses:

TABLE A-4-1

PROJECTED U.S. PROPANE CONSUMPTION

(Million Barrels)

	1975	1980	1985	1990
End-Use				
Residential/Commercial	159.2	183	202	224
Internal Combustion	14.3	15	15	15
Agriculture	42.9	46	50	55
Industrial	22.4	27	32	39
Petrochemical	54.3	90	82	82
Utility				
- Propane	9.5	15	23	36
- SNG	3.9	11	12	27
- Enrichment	-	-	3	3
	13.4	26	38	66
TOTAL	306.5	387	419	481

- Peak-shaving with propane/air systems.
- As a feedstock for SNG plants.
- Btu enrichment of pipeline gas.

Propane/air systems represent a flexible, low cost and quickly constructed peak-shaving alternative to such capital intensive supplements as SNG or imported LNG. Therefore, a high growth rate in this sector, is anticipated. We also believe that limited interest may re-emerge in propane as a feedstock for SNG in the early 1980's, if the FEA is pressured into acceptance of high cost natural gas supplements, as appears possible from the current versions of the Carter Energy Plan.

In the petrochemical industry, propane is used as a feedstock for ethylene manufacture. In this use it is particularly interchangeable with ethane. Because of the expected decline in domestic ethane-propane availability for olefin production, very few new gas liquids-based olefin plants have been built in recent years. Most new plants are based on gas-oil or naphtha, and this trend is expected to continue. One new plant based on ethane cracking is currently under construction, and it is possible that one additional plant based on ethane could be justified. However, due to the wide gap between natural gas and oil product prices in the United States, ethane extraction from natural gas has been growing rapidly. In the petrochemical industry, ethane can be valued as a feedstock in relation to oil products, while as a component of natural gas its value is strictly controlled. The increase in ethane extraction will partially displace domestic propane from the petrochemical feedstock sector.

Other sectors in which propane is used include agriculture, where the replacement of propane-fueled tractors by higher horsepower diesel-engined tractors will limit growth. Propane use in internal combustion engines, mainly for fork-lift trucks, also has limited growth possibilities due to competition from battery-powered vehicles.

Three quarters of U.S. butane consumption, which totalled 243 million barrels in 1975, is for gasoline blending; another 20 percent is used as a petrochemical feedstock. Neither of these two uses are likely to grow significantly in the forecast period, and increased butane production from U.S. refineries should more or less keep pace with domestic requirements.

On the supply side, due to the decline in domestic natural gas production, as well as the increasing leanness of new non-associated gas fields which will be brought into production during the next five years from OCS fields in the Gulf of Mexico, production from gas liquids extraction plants will decline (Figure A-4-1). The two short periods in which this trend will briefly be reversed are due first to the extraction of LPG associated with production from Naval Petroleum Reserve 1 (Elk Hills) and secondly from extraction of LPG from North Slope Alaskan associated gas, although plans for the latter are still fluid.

LPG production from U.S. refineries will increase steadily over the period with increased crude oil runs. The proportion produced from catalytic reformers will increase due to higher severity operations needed to meet the lower lead content of gasoline. The proportion produced from residue conversion processes will decrease in response to the heavier product demand per barrel which is foreseen. In addition, butane produced from natural gasoline will decline with decreased domestic availabilities.

The difference between forecast consumption and domestic supply must be met from imports. It is anticipated that overland imports from Canada will decline over the period 1975-1990. The balance of demand must be supplied by sea, and Table A-4-2 shows the projected growth of imports as domestic LPG availability declines over time.

This table shows that seaborne net imports of LPG, which were only 1.0 million barrels in 1975 (gross seaborne imports were 10 million barrels), will rise to nearly 240 million barrels by 1990. The great majority of this additional import requirement results from the projected growth in domestic demand (rather than from the relatively smaller reduction in domestic LPG supply), and is thus susceptible to government regulation -- either in terms of import controls, tariffs or other means of import limitations.

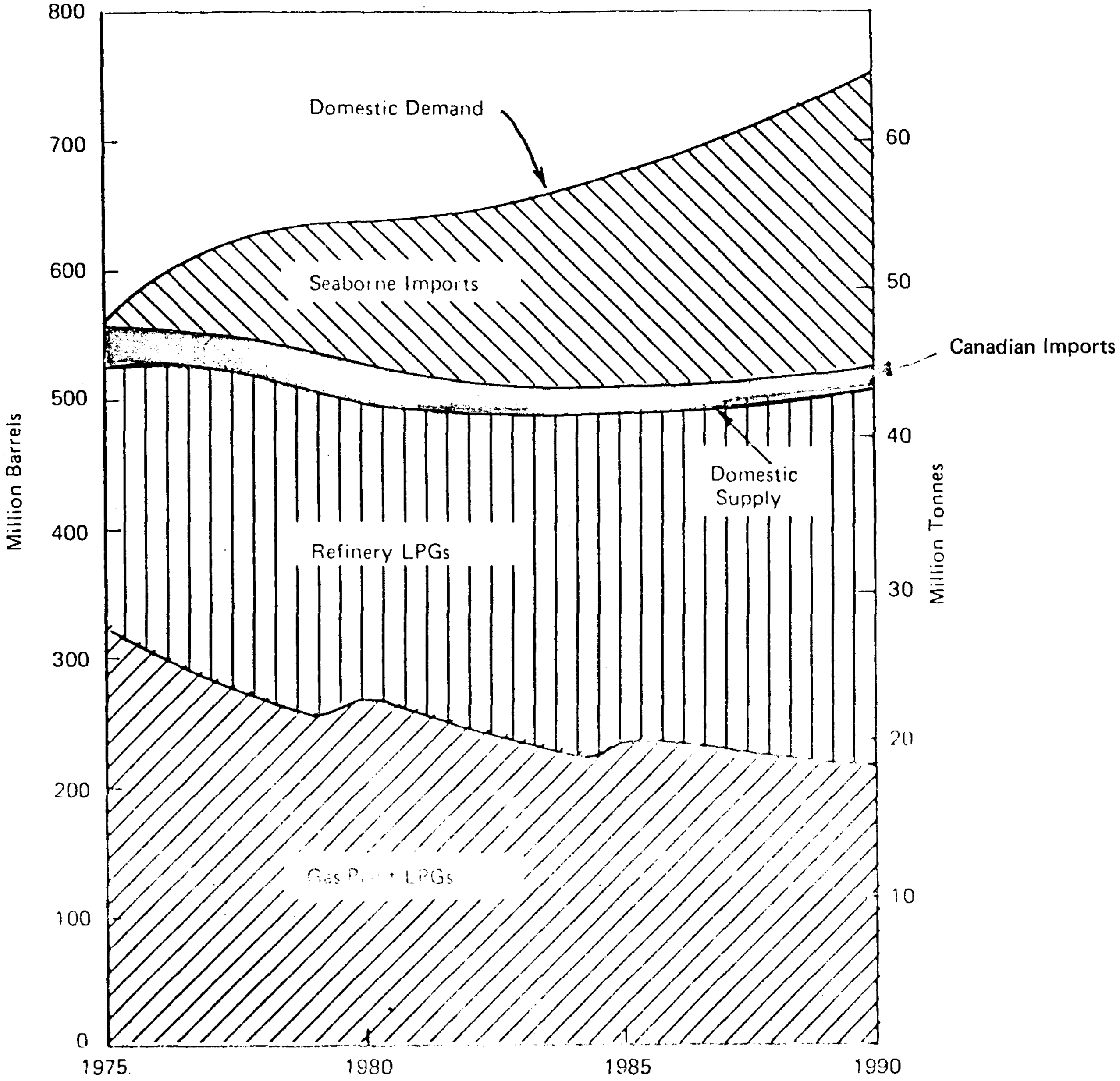


FIGURE A-4-1 U.S. LPG SUPPLY & DEMAND, 1975-1990

TABLE A-4-2

PROJECTED U.S. LPG SUPPLY AND DEMAND

(Million Barrels)

	1975	1980	1985	1990
PROPANE				
Demand	306.5	387	419	481
Stock Change	11.0	3	2	2
Domestic Production	289.4	271	260	261
Canadian Imports	14.9	17	12	10
Net Seaborne Imports	2.5	102	149	212
Statistical Difference	10.7	-	-	-
BUTANES				
Demand	243.1	244	248	260
Stock Change	(3.4)	-	-	-
Domestic Production	235.0	227	227	236
Canadian Imports	15.4	17	11	8
Net Seaborne Imports	(1.5)	-	10	16
Statistical Difference	(9.2)	-	-	-

This also opens up the question of logistics. The major deficit areas for LPG are in the Northeastern quadrant of the United States. These areas have historically been supplied, largely by pipeline, from the main producing areas in the south.

It is generally believed that the two main ways in which imported LPG will move to the deficit areas are via the Gulf Coast, and through Sun's terminal near Philadelphia. There are plans to reverse the Northeast segment of the Texas Eastern pipeline which would permit LPG to move from Sun's terminal inland into deficit areas. It is estimated that if the new terminals on the Gulf planned by Conoco and the Texas Eastern/Northern Natural group are implemented, there will be sufficient import terminal capacity to handle the predicted volume of imports. There are a large number of suitable salt dome structures in the Gulf Coast area, which tend to favor this area for future import terminals.

By contrast to the forecast adequacy of import terminal capacity, we have some concern on available pipeline capacity. Apart from the project to reverse the upper part of the Texas Eastern line, we know of no plans to increase the capacity of the major interstate pipelines. This has led and will continue to lead, to supply bottlenecks and could ultimately act as a brake on the expansion of LPG consumption in inland areas.

4.1.2 The World

Despite the large increase projected in U.S. LPG import requirements between 1975 and 1990, the world supply/demand picture is expected to remain in balance (with a tendency towards surplus), principally as a result of the major new gas processing projects that are underway in almost all of the OPEC countries. A summary of the major trends in LPG supply and demand from 1975-1990 is shown in Table A-4-3.

In Europe, a reduction in historic growth rates for LPG in domestic, commercial and industrial uses is projected due to strong competition from natural gas. There are growth prospects in the petrochemical sector, but these will probably not be realized given a continuation of current price relationships.

TABLE A-4-3

FORECAST WORLD LPG SUPPLY AND DEMAND

(Million Barrels)

	1975	1980	1985	1990
U.S.A.				
Supply	524	498	487	497
Demand	555	634	669	743
Surplus/(Deficit)	(31)	(136)	(182)	(246)
W. Europe				
Supply	143	228	268	299
Demand	141	182	207	238
Surplus/(Deficit)	2	46	61	61
Japan				
Supply	51	68	82	93
Demand	124	184	230	282
Surplus/(Deficit)	(73)	(116)	(148)	(189)
OPEC				
Supply	98	300	508	615
Demand	18	36	54	69
Surplus/(Deficit)	80	264	454	546
Rest of World				
Supply	60	269	304	351
Demand	38	209	298	379
Surplus/(Deficit)	22	60	6	(28)
WORLD				
Supply	876	1363	1649	1855
Demand	876	1245	1458	1711
Surplus/(Deficit)	-	118	191	144

High growth rates for Japanese consumption of LPG are projected, particularly in the industrial sector where a high value is attached to low sulfur fuels. OPEC countries will also show high growth rates in LPG consumption as they emphasize the use of gaseous fuels in internal markets. Other countries will continue to show growth rates in LPG consumption which are rather higher than those of other products.

On the supply side, extraction of LPG associated with North Sea natural gas and oil production will lead to a substantial increase in European supplies over the period considered. There will also be a contribution from natural gas liquids extraction projects in Brunei, Australia, Mexico and Brazil. Elsewhere, supply outside the OPEC countries will increase with refinery capacity.

For the OPEC countries LPG development plans have been analyzed on a project-by-project basis. A summary of projected capacity for the major OPEC producers is shown in Table A-4-4. These capacity forecasts take into account expected project delays as well as an allowance for a slower than normal buildup to full capacity in the face of possible start-up difficulties. They do not account for rationalization by OPEC of the apparent worldwide LPG surplus that will be caused by such developments, nor for the possibility of under-utilization caused by cutbacks in associated gas production linked to OPEC oil exports.

The apparent "surplus" of world LPG production reaches a peak of nearly 200 million barrels in 1985, declining somewhat thereafter. The 1985 "surplus" of 191 million barrels amounts to 42% of the exportable OPEC surplus in that year. One major influence on the statistical surplus of LPG's shown in Table A-4-3 is the potential for NGL re-injection into the producing formations. Two major schemes are under construction in Saudi Arabia which would ultimately have a re-injection potential of some 80 million barrels/year. These schemes would, if used, substantially reduce the overall LPG surplus by LPG re-injection, however, will not be undertaken lightly since it may result in a low present value for the re-injected LPG produced over the life of the gas processing plant. (This depends on the financial time preference of individual OPEC member country.)

TABLE A-4-4

PROJECTED OPEC LPG PRODUCTION CAPACITY

(Million Barrels)

	1980	1985	1990
Saudi Arabia	119	165	209
Abu Dhabi	13	30	47
Kuwait	33	55	55
Iran	26	63	74
Other OPEC Middle East	11	49	62
TOTAL OPEC Middle East	202	362	447
Algeria	46	89	109
Venezuela	32	32	32
Other OPEC Outside Middle East	20	25	27
TOTAL OPEC	300	508	615

The other alternatives to re-injection are to burn or flare it (this effect would be achieved by further project delays) or to lower the price in the hope of developing further markets. It is this last course of action, which may be the least unattractive for the OPEC countries, that is most likely to result in a re-equilibration of world LPG supply and demand.

Currently, imported propane prices (as well as domestic spot prices) lie in the range of \$3.00-\$3.50/$\overline{\mathrm{M}}$Btu at Mont Belvieu, and are about 25% higher (in calorific terms) than No. 2 oil prices. Over time, however, it is expected that imported propane prices will weaken relative to other oil products as the OPEC NGL plants begin to come on stream. The production "surplus" will become most unmanageable in the early to mid-1980's and is expected to result in a bottoming out of the LPG price (in relative terms) in the early 1980's.

4.2 LPG PRICE FORECAST

LPG price movements in the U.S. market, particularly those on the spot market, have always been difficult to predict. This stems from a number of issues:

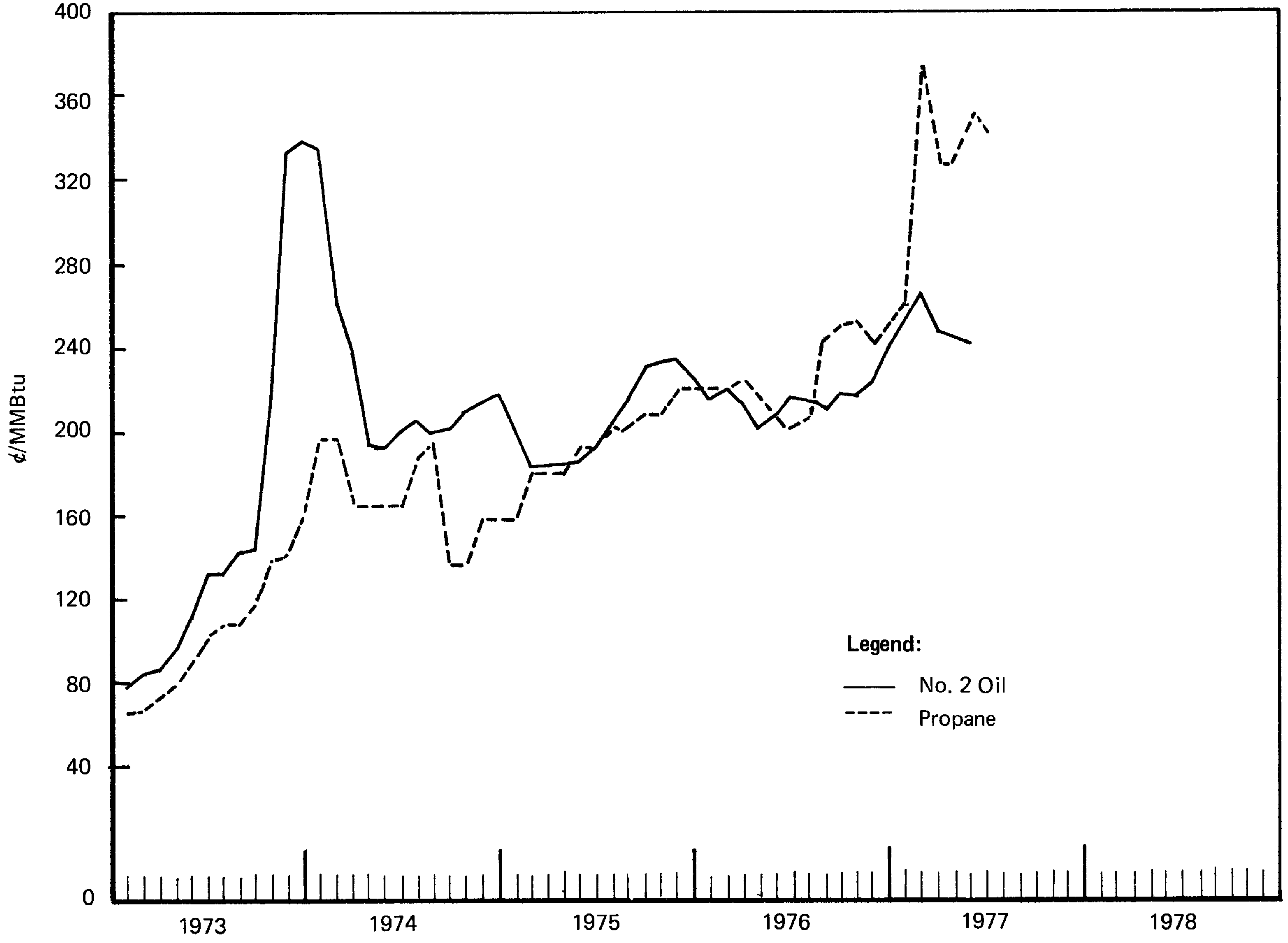
- LPG extraction from natural gas may vary according to the price relationships between gas and oil products
- Price controls and other legislative restrictions
- Increasing natural gas curtailments creating temporary opportunities for premium standby fuels such as LPG
- The great sensitivity of LPG prices to relatively small changes in the supply/demand balance
- Provisions of the current FEA regulations, permitting the banking of unrecovered product costs

Many of these factors have only become apparent in the past five years. Prior to the cold winter of 1976/77, prices for domestic spot propane were in the region of 22-24¢/gal at Mont Belvieu, while imported propane prices ranged from 30-35¢/gal landed. The existence of such a discrepancy was in part due to FEA's price and allocation regulations and in part due to the weakness of the LPG market at that time. As Figure A-4-2 shows however, spot propane prices moved sharply upwards during the past cold winter and have remained close to import parity. At these price levels (shown in terms of calorific equivalence with No. 2 distillate in the U.S. Gulf), the price premium for propane at the end-users' location will be even higher than that shown for the Gulf Coast since the pipelining, distribution and marketing costs associated with moving propane from the Gulf to the deficit Midwest and East Coast areas are significantly higher than for mainline oil products. Currently, therefore, LPG users are paying high premia for propane as compared to the cost of competing fuels.

This has not been the case historically (Table A-4-5). Propane prices in the U.S. Gulf have, until recently, been intermediate between distillate and residual in heating value terms, although the consumer-delivered prices of propane would probably be around or somewhat above distillate parity.

In the future, a number of trends will influence the development of U.S. and international propane prices:

- Growing world surpluses of LPG will probably cause some real price weakness in international prices, particularly in the mid-1980's.
- Marginal LPG movements required to balance world LPG supply and demand will probably be to the U.S., which is the most price-elastic market.
- Continued FEA regulation of propane will direct cheaper price-controlled domestic product to high priority users, leaving gas utilities and industry to compete for uncontrolled imported product which must be priced incrementally. This issue is crucial for the electric utility industry which, under current legislation, would be forced to import higher-priced non-Canadian LPG with the attendant security of supply risks.
- Propane's current premium over distillate will diminish. Given the uncertainty over future U.S. propane requirements, imported LPG prices will fall no lower than parity with low sulfur residual fuel oil on the Gulf Coast (at which price a very large volume of propane would be absorbed by industrial and utility users), nor rise higher than a 25% premium over No. 2 distillate, which is the case today.



Source: Oil Buyer's Guide, Platt's Oilgram.

FIGURE A-4-2 COMPARATIVE COSTS OF PROPANE AND NO. 2 FUEL U.S. GULF COAST

TABLE A-4-5

U.S. AVERAGE FUEL PRICES

	Propane*		No. 2 Fuel[†]		No. 6 Fuel[§]	
	¢/Gal	¢/M̄Btu	¢/Gal	¢/M̄Btu	$/Bbl	¢/M̄Btu
1955	3.25	35.5	8.91	64.6	n.a.	n.a.
1956	4.92	53.8	9.16	66.4	2.32	38.7
1957	5.03	55.0	9.99	72.4	2.93	48.8
1958	5.39	58.9	9.12	66.1	2.31	38.5
1959	5.54	61.0	9.24	67.0	2.30	38.3
1960	5.30	57.9	8.61	62.4	2.46	41.0
1961	4.64	50.7	9.17	66.4	2.49	41.5
1962	4.16	45.5	8.61	62.4	2.31	38.5
1963	4.42	48.3	8.76	63.5	2.29	38.2
1964	3.93	42.9	8.13	58.9	2.35	39.2
1965	4.50	49.2	8.58	62.2	2.35	39.2
1966	5.23	57.2	8.74	63.3	2.35	39.2
1967	6.17	67.4	9.48	68.7	2.22	37.0
1968	4.42	48.3	9.40	68.1	2.24	37.3
1969	4.16	45.5	9.24	67.0	2.03	33.8
1970	5.93	64.8	9.41	68.2	3.01	50.2
1971	5.86	64.0	9.80	71.0	3.72	62.0
1972	5.71	62.4	10.10	73.2	3.69	61.5
1973	9.56	104.5	21.74	157.5	4.48	74.7
1974	16.06	175.5	30.69	222.4	12.38	206.4
1975	18.32	200.2	28.43	206.0	10.21	170.2
1976	19.94	217.9	29.88	216.5	10.14	169.0

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*For pipeline input, minimum 10,000 bbl, Mt. Belvieu, Texas. 1965 and before, prices are for Baton Rouge.

[†] Gulf Coast cargo.

[§] Gulf Coast cargo, maximum 0.6-1.0% sulfur.

Gross heating values for fuels are (M̄Btu/bbl): Propane - 3.927; No. 2 Oil - 5.8; No. 6 Oil - 6.0.

Source: Platt's Oilmanac.

- The future evolution of international crude oil prices, freight rates and refinery margins affecting the cost of competing oil products.

By 1980, therefore, the premium for imported propane on the U.S. Gulf Coast is projected to decline to around 10% over the Btu equivalent price of gas oil. As the world LPG "surplus" develops during the early 1980's, international LPG prices will continue to fall relative to other oil products. In order to dispose of this surplus, new large volume markets for LPG must be developed where LPG has little or no form premium over products such as low sulfur fuel oil (such as in the industrial base-load fuel market). By 1985 it is expected that the value of imported propane will decline to the Btu equivalent of 0.5% No. 6 oil on the U.S. Gulf Coast. By 1990, however, the "surplus" of LPG production will be reduced, particularly if major new markets have been developed in the early eighties. Consequently an upward movement is anticipated once again, so that in 1990 propane would have a value equal to gas oil (on a heating value basis). An illustration of the calculation of the values of LPG which result from these assumptions is given in Table A-4-6. In Table A-4-7 are listed the delivered prices of LPG to the four regional locations, using the marketing and distribution costs for LPG previously developed.

TABLE A-4-6

PROJECTED EX-TERMINAL VALUES FOR IMPORTED PROPANE

(Prices in 1975 $/bbl Unless Otherwise Stated)

Basis: 1980 - 10% premium over calorific equivalent of ex-refinery No. 2 oil price
1985 - Calorific equivalent of ex-refinery 0.5% S No. 6 oil price
1990 - Calorific equivalent of ex-refinery No. 2 oil price.

	Gulf Coast						West Coast					
	1980		1985		1990		1980		1985		1990	
	Controls	Decontrol	Controls	Decontrol	Controls	Decontrol	Controls	Decontrol	Controls	Decontrol	Controls	Decontrol
Virgin Gas Oil Price	14.22	15.54	–	–	20.05	25.42	15.66	16.43	–	–	20.58	24.97
0.5% S No. 6 Fuel Price	–	–	16.64	21.63	-	–	–	-	16.39	20.68	–	-
Value ($/ MBtu	2.452	2.679	2.750	3.575	3.457	4.383	2.700	2.833	2.709	3.418	3.548	4.305
Propane Premium	10%	10%	–	–	-	–	10%	10%	–	-	–	–
Propane Value												
($/MBtu	2.697	2.947	2.750	3.575	3.457	4.383	2.970	3.116	2.709	3.418	3.548	4.305
($/bbl)	10.55	11.52	10.75	13.98	13.52	17.14	11.61	12.18	10.59	13.36	13.87	16.83

Assumed Heating Values (Million Btu's/bbl)		
	- Propane	3.91
	- Gas Oil	5.80
	- 0.5% S No. 6 Fuel	6.05

TABLE A-4-7

DELIVERED LPG (PROPANE) PRICES

(1975 $/bbl)

LOCATION:	Hartford	Columbus	Dallas	Los Angeles
1980				
Controls	15.15 (3.87)	14.00 (3.58)	13.04 (3.34)	13.09 (3.35)
Decontrol	16.12 (4.12)	14.97 (3.83)	14.01 (3.58)	13.66 (3.49)
1985				
Controls	15.35 (3.93)	14.20 (3.63)	13.24 (3.39)	12.07 (3.09)
Decontrol	18.58 (4.75)	17.43 (4.46)	16.47 (4.21)	14.84 (3.80)
1990				
Controls	18.12 (4.63)	16.97 (4.34)	16.01 (4.09)	15.35 (3.93)
Decontrol	21.74 (5.56)	20.59 (5.27)	19.63 (5.02)	18.31 (4.68)

Note: Values in parenthesis are $/$\overline{\text{M}}$Btu.

Source: A.D. Little Estimates

APPENDIX B

FUEL-TO-POWER SYSTEM DEFINITION

1.0 DISPERSED GENERATION SYSTEMS

1.1 CONVERSION TECHNOLOGY DESCRIPTION

1.1.1 Steam Reforming

In the United States, hydrogen has generally been produced by steam methane reforming due to the historically low price of natural gas. The familiar steam methane reforming reaction is:

$$CH_4 + H_2O = CO + 3H_2$$

This reaction is carried out in a furnace containing vertical tubes filled with a nickel catalyst to promote the reaction. Heat is applied externally by fuel firing to maintain an outlet temperature which is usually in the range of 1300°F to 1650°F.

Hydrogen is also produced from the reaction of steam with carbon monoxide as follows:

$$CO + H_2O = CO_2 + H_2$$

This reaction is commonly known as the CO shift and occurs to a limited extent in the reforming furnace. The CO shift reaction is favored at temperatures lower than those encountered in the reforming furnace and is generally carried out downstream of the furnace in a separate converter. A typical process flow schematic for this process is shown in Figure B-1-1.

Steam reforming of heavier hydrocarbons up to and including desulfurized naphtha is also possible. The generalized steam hydrocarbon reaction is as follows:

$$C_nH_m + nH_2O \rightarrow nCO + \frac{(2n + m)}{2} H_2$$

For feedstocks heavier than naphtha, coking of the reformer catalyst becomes a severe operating problem. This is due in part to the difficulty in removing sulfur from the heavier distillates to 0.5 ppm. One way to counter the coking problem with steam reforming is to use higher process

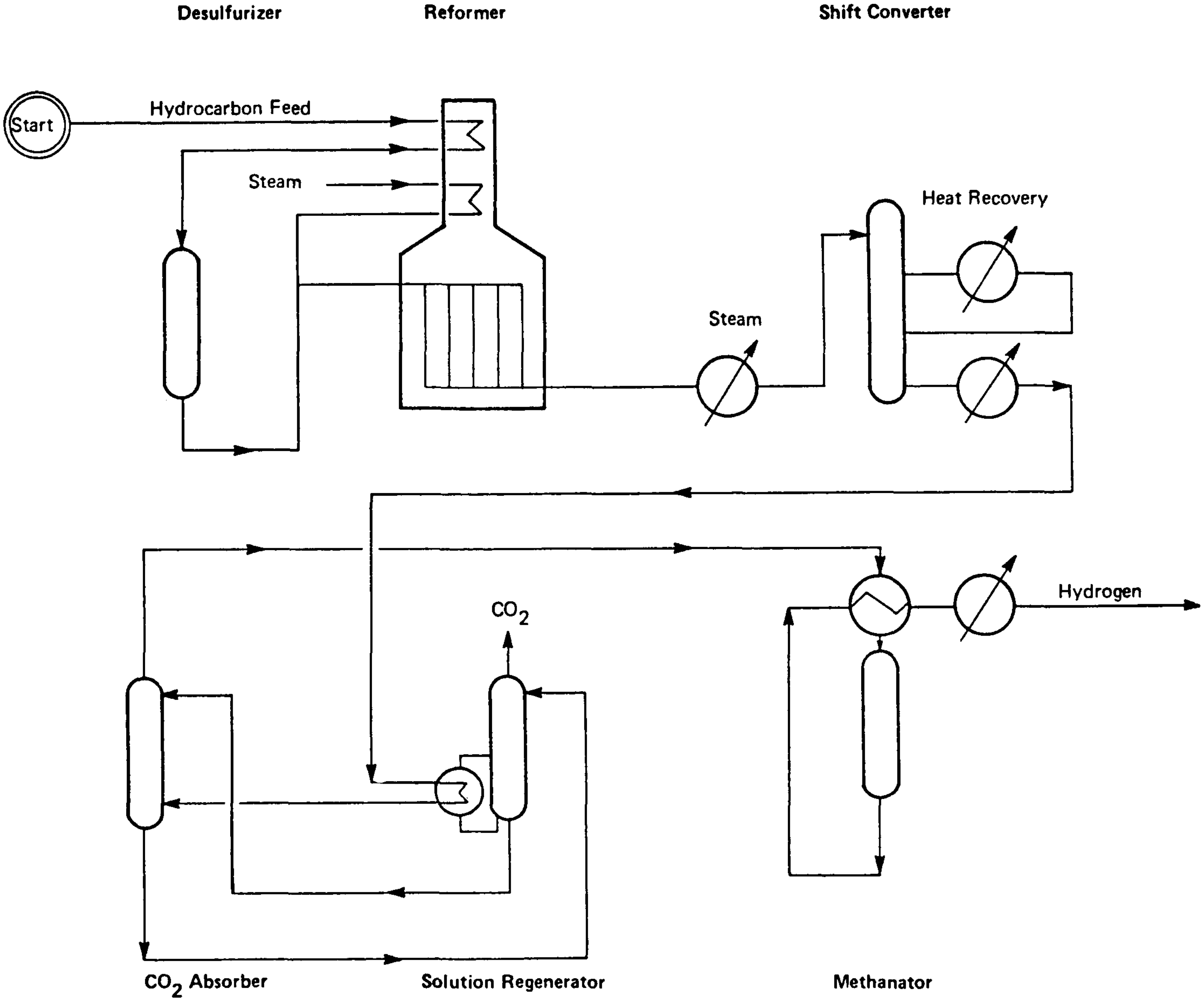


Figure B-1-1 Hydrogen Production by Steam Reforming temperatures. Ultimately, a different process is utilized for producing hydrogen from heavy hydrocarbons.

1.1.2 Autothermal Reforming

For gas oil and heavier feeds, a process known as partial oxidation is commercially available to produce hydrogen. As the name implies, the feedstock is burned with sub-stoichiometric oxygen to produce a gas containing hydrogen, carbon monoxide and carbon dioxide. Enough oxygen is supplied to maintain the flame temperature within the desired range to permit cracking of the remaining feedstock. Pure oxygen or air may be used to support combustion, however, the use of air produces synthesis gas diluted with a large amount of nitrogen.

Autothermal reforming (ATR) is a hybrid technology which combines partial oxidation and catalytic reforming of hydrocarbon fuels to produce synthesis gas (H_2 + CO). The process utilizes the exotherm of partial oxidation to preheat the reactants and to supply the heat of reaction for the reforming reactions, which are still conducted over nickel catalyst.

The advantage of autothermal reforming is the elimination of high temperature chrome-nickel alloy furnace tubes which allows more latitude to increase reforming temperatures. This is particularly important in regard to using conventional distillate fuels, since increasing the temperature improves the reforming operability in terms of sulfur tolerance and elimination of coking.

1.2 FUEL CELL INTEGRATION ASPECTS

Two dispersed fuel cell generating systems using steam reforming of low sulfur distillate fuels were evaluated. In one system, a conventional phosphoric acid fuel cell was matched with a conventional steam reformer and shift converter. In the other system, a more advanced molten carbonate fuel cell was matched with a high temperature steam reformer. A third generating system which combines the molten carbonate fuel cell with autothermal reforming was also evaluated. The designs and performance of these three systems are discussed below.

1.2.1 Conventional Steam Reforming

1.2.1.1 Process Description and Flow Sheet

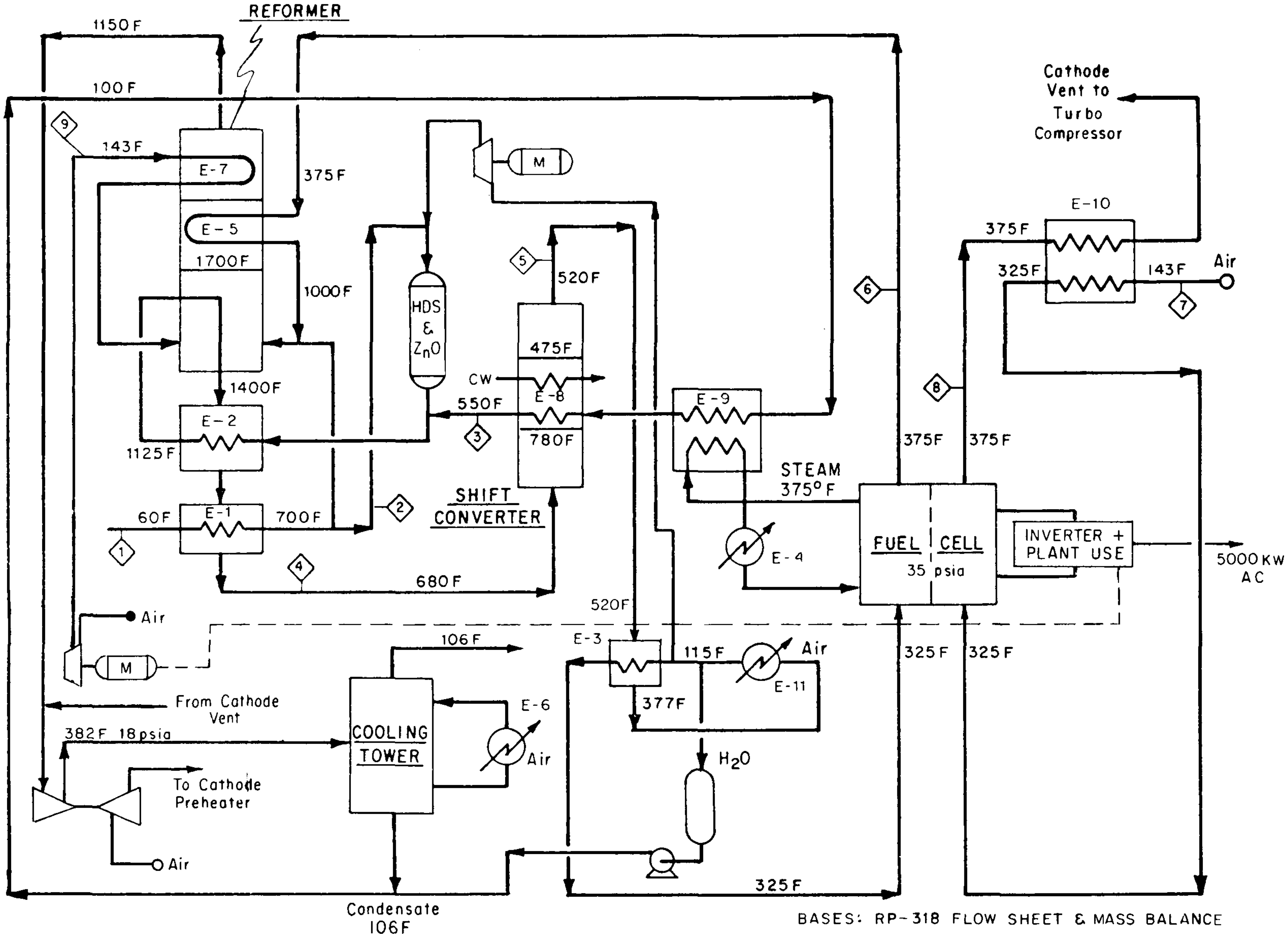
The more conventional steam reforming concept combines low temperature (i.e., state-of-the-art naphtha reforming) steam reforming with shift conversion to produce impure hydrogen for fuel in a phosphoric acid fuel cell. This system is illustrated schematically in Figure B-1-2. The schematic was derived trhough minor modification of the integrated steam reforming phosphoric acid fuel cell presented in EPRI Report RP-318. The modifications reduced the number of heat exchangers required, recognized the need for a turboexpander to improve energy efficiency and reduce process component costs, and adjusted the heating rate to the range of the first generation acid cell system (FCG-1).

As shown in the schematic, naphtha is vaporized and split into two streams. The smaller stream is used for supplemental fuel in the reforming furnace. The major portion of the vaporized oil is hydrodesulfurized, mixed with steam in a 4:1 steam-oil mol ratio, and reformed to synthesis gas at 1400°F in the reforming furnace. In the shift converter, this synthesis gas is enriched in hydrogen at the expense of carbon monoxide and water, as in the reaction:

$$CO + H_2O \rightarrow H_2 + CO_2$$

The impure hydrogen stream leaving the shift converter is cooled to condense water, preheated to near the phosphoric acid fuel cell operating temperature of 375°F, and supplied as fuel to the anode side of the fuel cell.

In the cell, most of the hydrogen is oxidized to release electrical energy. In a first generation phosphoric acid fuel cell, the cell operating voltage is only 0.65 volts, as opposed to the expected potential of circa 0.8 volts for second generation fuel cells. The energy loss implied by this voltage difference appears as waste heat, which is removed by heating the anode and cathode feed streams and by auxiliary waste heat exchangers.



BASES: RP-318 FLOW SHEET & MASS BALANCE

Figure B-1-2. ACID CELL INTEGRATED PROCESSOR CONVENTIONAL STEAM REFORMING.

The anode vent contains some combustible hydrogen, carbon monoxide, and methane and is therefore used as fuel in the steam reforming furnace. Since the reformer furnace and the fuel cell are operated above ambient pressure, useful work can be extracted from the high temperature gases leaving the furnace and cathode by expanding them in a turboexpander as shown in the schematic. This expansion not only yields useful work, but also reduces the temperature of the stack gases. Water is recovered from the expander exhaust in a cooling tower. Efficient recovery of water at this point and in the anode feed condenser, results in self-sufficiency in process water for the integrated steam reformer/phosphoric acid fuel cell system.

To optimize energy requirements, the various inputs to the integrated generating system are heated to process temperature by heat exchange with various hot process streams. As shown in the schematic, most of the heat of vaporization for the reformer steam feed is derived from fuel cell waste heat. The air supplies to the reforming furnace and the fuel cell are preheated by exchange with the respective exhaust gases. The oil feed to the system is vaporized in another heat exchanger, simultaneously cooling the reformer effluent to suitable temperatures for the shift converter.

1.2.1.2 Mass Balance and Utility Summary

The flow of material through a 5 mw (net AC power) integrated generating system is summarized in Table B-1-1. For the system illustrated, the overall heat rate (fuel oil to net AC power) is 9000 Btu/kWh. As shown in the utility summary (Table B-1-2), most of the work required to compress the process air requirement is recovered by the turboexpander. The major loss shown in the utility summary is the 4% loss in the inverter, which converts the DC power output of the fuel cell to AC current. More extensive optimizing studies might allow lower cooling requirements and a lower heat rate.

Catalyst requirements shown in Table B-1-2 are based on 20,000-hour life expectancy. The stack replacement allowance is based on 40,000 hours

TABLE B-1-1

CONVENTIONAL STEAM REFORMING/ACID FUEL CELL

5.0 mw Capacity (Net)

MASS BALANCE - 0.65 v
(lb mols/hr)

Stream No.	1	2	3	4	5	6	7	8	9
Component									
H_2	--	--	--	326.2	378.0	37.8	--	--	--
CO	--	--	--	55.6	3.7	3.7	--	--	--
CO_2	--	--	--	65.9	117.7	117.7	--	--	--
H_2O	--	--	491.7	298.5	246.7	26.4	--	340.3	--
CH_4	--	--	--	1.4	1.4	1.4	--	--	--
O_2	--	--	--	--	--	--	378.1	207.9	89.8
N_2	--	--	--	--	--	--	1422.3	1422.3	337.0
TOTAL	--	--	491.7	747.6	747.6	187.1	1800.3	1970.5	426.8
lb/hr	2,250	1,756	8,851	10,608	10,608	5,858	29,082	29,763	12,334

Exchangers	E-1	E-2	E-3	E-4	E-5	E-6	E-7	E-9	E-10	E-11	E-12	E-13
Duty, MMBtu/hr	1.09	3.47	0.85	5.48	1.21	8.22	2.24	9.30	2.28	4.18	0.94	0.66

TABLE B-1-2

STEAM REFORMING/PHOSPHORIC ACID FUEL CELL

(5.0 mw Net Output)*

UTILITY BALANCE

Gross Power, kw		
Fuel Cell	5,319	
Turboexpander[†]	293	
Subtotal		5,612
Plant Power Requirements, kw		
Pumps and Compressors	(399)	
Inverter Loss	(213)	
Subtotal		(612)
Net Output		5,000
Fuel Input, $\overline{\text{M}}$Btu/hr		45.00
Reformer Catalyst, ft^3/yr	2	
Shift Converter Catalyst, ft^3/yr		
Low Temperature	9.8	
High Temperature	9.3	
ZnO, ft^3/yr	90	
Labor, man hr/yr	750	
Stack Replacement, mil/kwh	2.0	
Other Maintenance, mil/kwh	0.2[§]	

Source: A.D. Little estimates.

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*for 3,000 hr/yr operation

[†] shaft work, expressed in kw

[§] 5% of CI for remainder of plant per year

TABLE B-1-3

FUEL PROCESSOR MAJOR EQUIPMENT
L.T. STEAM REFORMING/PHOSPHORIC ACID FUEL CELL

5 mw Net Capacity

Name	Description	Size	Comments
Exchangers			
E-1	Naphtha Vaporizer	46 ft^2	
E-2	Reformer Feed Preheat	674 ft^2	
E-3	Cell Feed Preheat	4,569 ft^2	
E-4	Condenser	1,599 ft^2	
E-5	Anode Vent Reheat	117 ft^2	
E-6	Cooling Tower	12,518 ft^2	
E-7	Reformer Air Preheat	363 ft^2	
E-9	Reformer Steam Vaporizer	59 ft^2	
E-10	Cathode Waste Heat	9,365 ft^2	
E-11	Anode Heat	868 ft^2	
E-12	Converter	635 ft^2	
E-13	Converter Waste Heat	571 ft^2	
		31,384 ft^2	
Pumps			
P_1	Naphtha Feed Pump	0.4 HP	
P_2	Water Recycle Pump	0.9 HP	
Vessels			
V-1	Knock-out Drum	33 ft^3	
V-2	Knock-out Drum	33 ft^3	
V-2	ZnO Vessel	15 ft^3	
Reactors			
R-1	Steam Reformer	4.82 MMBtu/hr fired duty	
R-2	Shift Converter	3,622 cfm space velocity	
Supplies			
S-1	Reforming Catalyst	12 ft^3	
S-2	H.T. Shift Catalyst	62 ft^3	
S-3	L.T. Shift Catalyst	65 ft^3	
S-4	ZnO Charge	22 ft^3	

of operation. Zinc oxide requirements are based on a sulfur loading of 15 pounds per cubic foot.

1.2.1.3 Major System Components

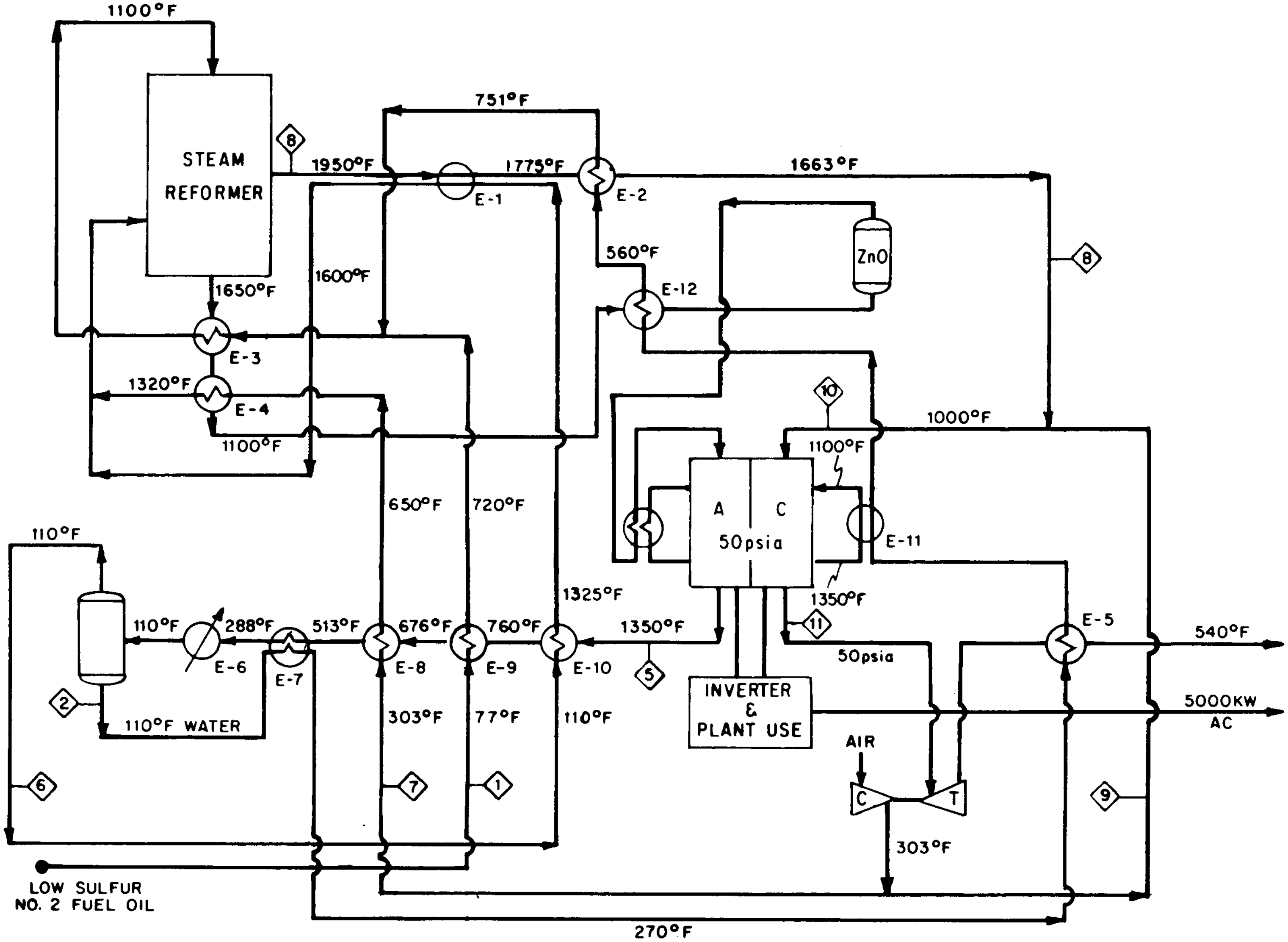
The major components of the steam reforming/phosphoric acid fuel cell system are summarized in Table B-1-3. These major components were sized using typical handbook design methods. No attempt was made to optimize heat exchanger designs or compression requirements, however, discussions were undertaken with vendors of form plate exchanges to determine the magnitude of heat transfer coefficients that might be expected. Heat exchangers were sized and costed on the basis of conventional shell-and-tube heat exchangers using pressure drops and heat transfer coefficients for analogous situations. Although production versions of the integrated fuel cell system would probably use compact plate-type heat exchangers to minimize cost and systems size, this system combines several fairly conventional technologies and, thus, represents the present state-of-the-art.

1.2.2 High Temperature Steam Reforming

1.2.2.1 Process Description and Flow Diagram

A more advanced, energy efficient, dispersed generating system was analyzed on the basis of combining a high temperature steam reformer with a molten carbonate fuel cell. The integrated system is illustrated schematically in Figure B-1-3. Except for the higher system operating temperatures and pressures and the lack of shift conversion, the system is similar to the phosphoric acid cell technology previously described. Shift conversion is not necessary with the molten carbonate cell; instead, the CO in the synthesis gas from the reformer can be utilized directly at the higher temperatures of the molten carbonate system.

As shown in the schematic, low sulfur No. 2 fuel oil is vaporized, mixed with steam in a 4:1 steam-oil mol ratio, and reformed to synthesis gas at 1650°F in a steam reforming furnace. This syngas is cooled to approximately 800°F for efficient desulfurization in a zinc oxide bed.



BASIS: MODIFIED RP 919-1 FLOW SHEET & MASS BALANCE

Figure B-1-3. MOLTEN CARBONATE CELL INTEGRATED PROCESSOR HIGH TEMPERATURE STEAM REFORMING

(Desulfurization is necessary in order to avoid contamination of the nickel electrodes used in a fuel cell.) The desulfurized synthesis gas is reheated by a fuel cell waste heat exchanger to near the cell operation temperature of 1350°F. In the cell anode, the syngas is partially oxidized, yielding electricity.

The anode vent gas is cooled to recover water that will be used in the steam reformer, then reheated, and the residual hydrogen, carbon monoxide, and methane in the vent gas are used as fuel in the steam reforming furnace. In this integrated system, supplementary oil firing in the reforming furnace is not necessary. Because carbonate ions are consumed at the anode, the reformer exhaust is recycled to supply carbon dioxide requirements at the fuel cell cathode. Additional air combines with this stream and reduces the temperature to approximately 1000°F before entering the fuel cell.

1.2.2.2 Mass Balance and Utility Summary

The mass flows for the various in-process streams of the integrated fuel processor module and the duties of the various heat exchangers are summarized in Table B-1-4. The mass balance and schematic described above were derived from a similar schematic developed by Catalytica for the EPRI RP-919 study. The Catalytica scheme was redesigned by increasing the operating pressure of the carbonate fuel cell, by simplifying the heat exchange and by including power recovery within the system. The operating voltage was also increased to 0.81 volts to reflect the increased cell performance obtained when cell pressure is increased.

The utility balance for the integrated steam reforming/molten carbonate fuel cell system is summarized in Table B-1-5. Because the molten carbonate fuel cell is operated at elevated pressure--approximately 50 psia--it would be wasteful of energy just to exhaust the cathode vent. Instead, this vent stream is expanded in a turbine to recover shaft work; our analysis indicates that sufficient shaft work could be generated in this fashion to power the reformer and cathode air compressors. Additional work could be recovered by attaching an electrical generator to the

TABLE B-1-4

HT. STEAM REFORMING/MOLTEN CARBONATE

(5 mw Net Output)

MASS BALANCE - 0.81 v
(lb mols/hr)

STREAM NO.	1 Oil Feed	2 Condensate	3 Anode Feed	4 Anode Vent	5 Reformer Fuel	6 Reformer Air	7 Reformer Exhaust	8 Cathode Air	9 Cathode Feed	10 Cathode Vent
Component										
H_2	--	--	316.94	101.29	101.36	--	--	--	--	--
H_2O	--	579.57	379.74	595.40	15.73	--	134.61	--	134.61	134.61
CO	--	--	69.55	22.23	22.21	--	--	--	--	--
CO_2	--	--	67.94	378.21	378.19	--	409.16	--	409.16	146.20
CH_4	--	--	8.76	8.76	8.76	--	--	--	--	--
O_2	--	--	--	--	--	87.28	16.74	255.00	271.74	131.49
N_2	--	--	--	--	--	328.33	328.28	959.30	1287.58	1287.58
Oil, lb/hr	2,016	--	--	--	--	--	--	--	--	--
TOTAL	--	579.57	842.93	1105.89	526.25	415.61	888.79	1214.30	2103.09	1699.88
lb/hr	2,016	10,441	12,558	28,338	17,894	11,990	29,884	35,033	64,917	49,115
MMBtu/hr	39.72	--	65.00	43.26	18.92	--	18.98	--	17.95	19.30

Exchanger:	E_1	E_2	E_3	E_4	E_5	E_6	E_7	E_8	E_9	E_{10}	E_{11}	E_{12}	E_{13}
Duty, MMBtu/hr:	1.71	0.98	2.20	2.12	7.77	12.87	2.22	1.69	0.90	6.66	1.04	2.19	2.19

TABLE B-1-5

STEAM REFORMING/MOLTEN CARBONATE

(5.0 mw Net Output)

Utility Requirements

Gross Power, kw		
Fuel Cell	5,216	
Turboexpander[*]	779	
Subtotal		5,995
Plant Power Requirements, Kw		
Pumps and Compressors	(786)	
Inverter Loss	(209)	
Subtotal		(995)
Net Output		5,000
Fuel Input, $\overline{\text{M}}$Btu/hr		39.72

ZnO, ft^3/yr	90
Catalyst, ft^3/yr	2
Labor,[†] hr/yr	750
Stack Replacement, mil/kwh	2.0
Other Maintenance, mil/kwh	0.2[§]

Source: A. D. Little Estimates

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[*] shaft work, expressed in kw

[†] for 3,000 hr/yr operation

[§] 5% of CI for remainder of plant per year

TABLE B-1-6

STEAM REFORMING/MOLTEN CARBONATE FUEL PROCESSOR MAJOR EQUIPMENT

Name	Description	Size	Comments
Exchangers			
E-1	Reformer Fuel Preheat #2	196 ft^2	
E-2	Reformer Steam Preheat	52 ft^2	
E-3	Reformer Feed Preheat	155 ft^2	
E-4	Reformer Air Preheat #2	759 ft^2	
E-5	Turbine Exhaust	413 ft^2	
E-6	Condenser	579 ft^2	
E-7	Water Preheater	657 ft^2	
E-8	Reformer Air Preheater #1	1999 ft^2	
E-9	Naphtha Vaporizer	105 ft^2	
E-10	Reformer Fuel Preheater #1	4512 ft^2	
E-11	Fuel Cell Waste Heat	121 ft^2	
E-12	Desulphurizer Feed Cooler	172 ft^2	
E-13	Anode Feed Reheat	803 ft^2	
Pumps			
P-1	Naphtha Feed Pump	0.2 HP	
P-2	Water Recycle	0.5 HP	
Vessels			
V-1	ZnO Tank	5' x 3' Diam.	
V-2	Knock-Out Drum	8' x 4' Diam.	
Reactors			
R-1	Steam Reforming Furnace	5.4 $\overline{\text{M}}$Btu/hr	
Supplies			
S-1	Reforming Catalyst	12 ft^3	
S-2	ZnO Charge	24 ft^3	

expander turbine, thus offsetting inverter losses. However, the inclusion of a small electrical generator in this system introduces additional complexity and cost which may be undesirable for a dispersed power plant.

The system as illustrated and analyzed is more efficient than the conventional acid system technology, developing a heat rate of 7945 Btu/kwh. At an operating voltage of 0.78 volts, the heat rate would be 8250 Btu/kwh for the system as shown. However, this is probably less efficient than actually achievable with the mating of these two advanced technologies. By slight redesign of the heat exchange and power recovery systems, the heat rate could probably be reduced to approximately 8000 Btu/kwh at 0.78 volts.

1.2.2.3 Major System Components

Based on the modified schematic, equipment was sized and costed on the basis of typical handbook design criteria. Only minor efforts were made to optimize the network of heat exchangers. Exchanger heat transfer coefficients and pressure drops were taken as typical for similar equipment; detailed exchanger design studies and extensive conversations with heat exchanger vendors were not part of this effort.

The major components of the fuel processor are summarized in Table B-1-6. As this table shows, and as will be shown under the discussion of costs, heat exchangers dominate the fuel processor module.

1.2.2.4 System Design Issues

For the purposes of this study, major equipment components were sized and costed on the basis of conventional technology. However, some of the equipment used in this system is beyond the state-of-the-art. The high temperature steam reformer has only operated at the pilot plant level to date, though several groups are developing this technology. One would also prefer to use more compact and efficient plate-type heat exchangers rather than shell-and-tube type. However, several heat exchanger vendors indicated that the system temperatures described here were beyond the capabilities of plate-type exchangers. Such exchangers are generally not used at temperatures in excess of 800°F; thus, further development of this equipment would be necessary before it could be used with this high temperature system.

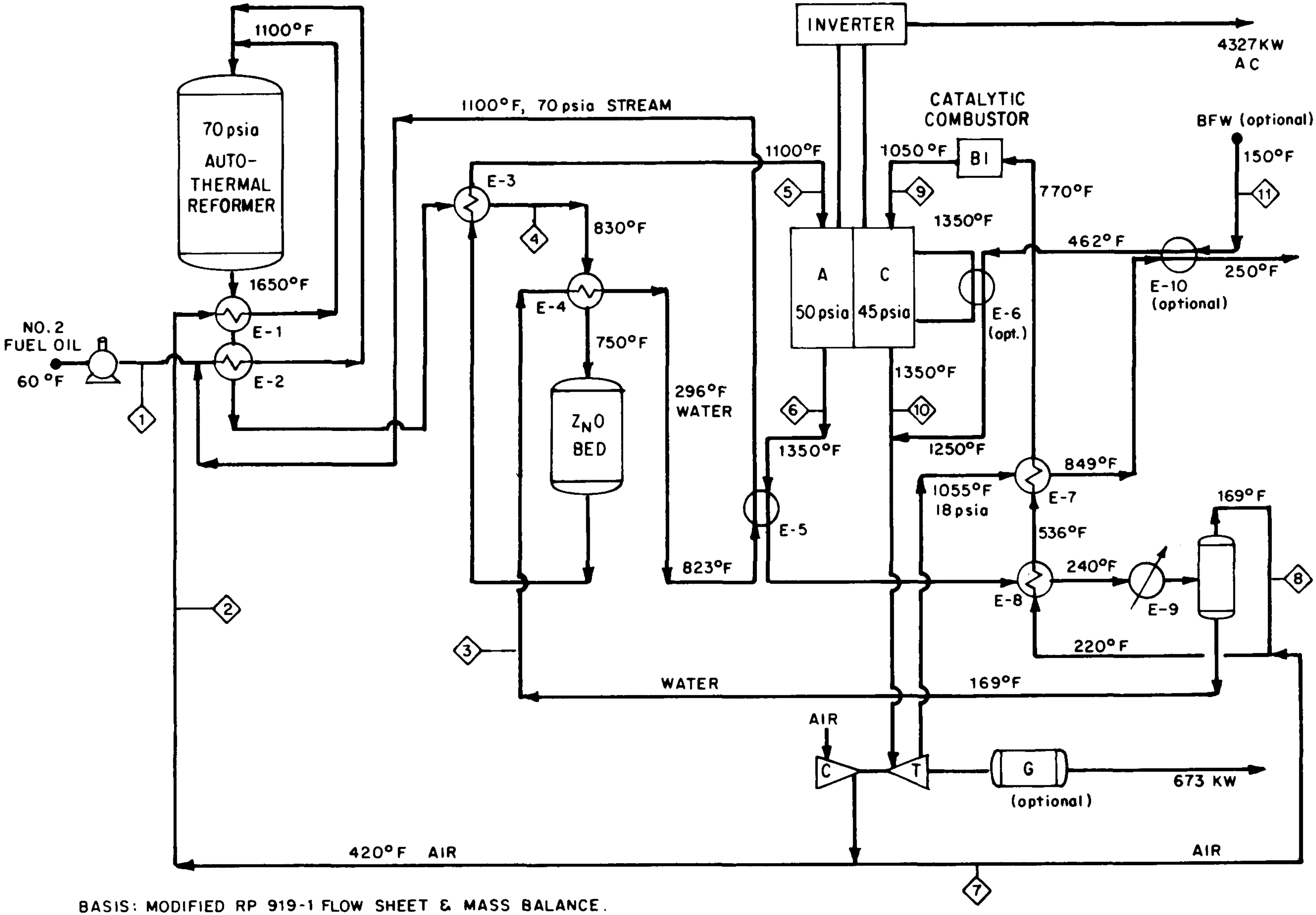
One other aspect of the high temperature steam reforming/molten carbonate system that might merit further developmental study is the desulfurization step. A zinc oxide bed should not be operated above 800°F for efficient utilization of the zinc oxide and sulfur removal. Therefore, to reduce the number of heat exchangers in the molten carbonate system, it would be interesting to search for desulfurization technologies that could operate at higher temperatures.

A design issue common to all the dispersed generator and baseload fuel cell systems concerns the excess air or oxidant provided to the fuel cell cathodes. For this study, 100% excess air was provided for all cases analyzed as directed by EPRI. A lesser quantity of excess air would reduce parasitic power requirements, reduce the sizes of certain heat exchangers, and improve (to a small extent) the systems' heat rates. The feasibility of using less than 100% excess air is well established particularly for phosphoric acid electrolyte fuel cell systems; the FCG-1 module being assembled is designed to operate with approximately 50% excess air. Obviously, design optimization must include the oxidant utilization parameter though such optimizations are beyond the scope of this study.

1.2.3 Autothermal Reforming

1.2.3.1 Process Description and Flow Diagram

The integration of this concept with a molten carbonate fuel cell is depicted in Figure B-1-4. A mixture of steam and vaporized low sulfur fuel oil is combined with a substoichiometric quantity of preheated air in a nozzle-type combustor. Partial combustion of the hydrocarbon fuel provides the heat needed to increase the temperature of the reactants and the endothermic heat of reaction required by the reforming process. The hot combustion products pass through a fixed bed of nickel reforming catalyst on which the reforming reactions take place.



BASIS: MODIFIED RP 919-1 FLOW SHEET & MASS BALANCE.

Figure B-1-4, MOLTEN CARBONATE CELL INTEGRATED PROCESSOR AUTOTHERMAL REFORMER (Catalytic Partial Oxidation).

At the temperature encountered with conventional steam reforming, the nickel reformer catalyst is rapidly deactivated by sulfur. The higher temperature achieved with autothermal reforming should allow feed sulfur levels to be increased. However, it is still questionable that virgin No. 2 fuel oil with 0.2 wt. percent sulfur can be handled without partial desulfurization, and in any event the product gases must be desulfurized prior to entering the molten carbonate fuel cell.

The gaseous effluent from the ATR reactor is cooled by exchange with combustion air, fuel and steam to below 900°F before entering a zinc oxide bed in which residual hydrogen sulfide is removed by reaction with zinc oxide (ZnO). Cooling the gas to below 900°F is necessary to achieve optimum removal efficiency with zinc oxide. The gas leaving the ZnO bed is reheated by exchange with inlet gas and fed to the fuel cell anode.

The anode vent stream is cooled in a series of exchanges to recover water which is vaporized and recycled to the ATR reactor. Air is injected into the dehumidified anode vent and the stream is reheated against itself and the cathode vent. The anode recycle next enters a catalytic combustor (afterburner) where its temperature is increased adiabatically to 1050°F before entering the fuel cell cathode.

Upon leaving the fuel cell, the cathode vent is combined with 45 psia steam which has been superheated by exchange with the cell coolant loop. The combined stream is let down through a turboexpander-compressor unit in which the resulting shaft work is used to compress the plant air requirements. A small electrical generator (optional) is also coupled to the turboexpander to generate additional power. After exchange with the inlet air, the cathode vent is discharged to atmosphere. Pressure conditions selected for the ATR reactor and the fuel cell stack are 70 psia and 50 psia, respectively, based on a conservative allowance for pressure drop through the system.

1.2.3.2 Mass Balance and Utility Summary

The mass balance for the ATR molten carbonate system is shown in Table B-1-7. The stream numbers refer to flags on the flow sheet (Figure B-1-4). The flow rates shown are based on a 5 $\overline{\mathrm{M}}$w (net) power plant with cell voltage at 0.78 volts. Raising pressure on the molten carbonate cell with the ATR processor does not increase the cell performance. Because nitrogen is present in the fuel, the hydrogen partial pressure at 50 psia is the same as for reformer systems at atmospheric pressure. Major heat exchanger duties are also presented in Table B-1-7.

Total energy input for the system is 36.7 $\overline{\mathrm{M}}$Btu/hr which results in an overall system heat rate of 7340 Btu/kwh.* Utility requirements for the system are summarized in Table B-1-8. The power plant gross output is 5910 kw including shaft work furnished by the turboexpander. The power plant parasitic load is 910 kw including fluid pumping and inverter losses. Net power plant output after satisfying plant needs is 5000 kw.

Water requirements (16.4 gpm) for the optional steam injection scheme are included for the expander power-recovery circuit. To be water-conserving with this power-recovery scheme, the exhaust stream must be cooled to 106°F. Elimination of the steam injection scheme will reduce the power plant net output to 4.7 Mw and increase the net station heat rate to 7835 Btu/kwh. The effect of this charge on capital investment is addressed later in this section.

The annual zinc oxide catalyst requirement for 200 ppm sulfur feed is 75 ft^3. This quantity would increase by a factor of ten with virgin No. 2 fuel oil. Labor for periodic inspection was estimated at 25% of annual operating time.

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*36.7 $\overline{\mathrm{M}}$Btu/hr ÷ 5000 kw = 7340 Btu/kwh.

TABLE B-1-7

ON-SITE FUEL PROCESSOR AUTOTHERMAL REFORMING/MOLTEN CARBONATE CELL

MASS BALANCE - 0.78 VOLTS

(lb mols/hr)

Stream No.	1	2	3	4	5	6	7	8	9	10	11*
Component											
H_2	--	--	--	190.68	190.68	28.60	--	28.60	--	--	--
H_2O	--	--	204.35	132.68	132.68	294.76	--	90.38	118.98	118.98	455
CO	--	--	--	87.41	87.41	13.11	--	13.11	--	--	--
CO_2	--	--	--	48.13	48.13	358.84	--	358.84	371.95	135.57	--
CH_4	--	--	--	0.67	0.67	0.67	--	0.67	--	--	--
O_2	--	55.99	--	--	--	--	258.62	--	237.77	119.53	--
N_2	--	210.66	--	210.66	210.66	210.66	922.89	210.66	1183.55	1183.55	--
H_2S	--	--	--	0.13	--	--	--	--	--	--	--
Oil, lb	1,882	--	--	--	--	--	--	--	--	--	--
TOTAL	--	266.65	204.35	670.36	670.23	906.64	1231.51	702.26	1912.25	1557.68	455
lb/hr	1,882	7,693	3,681	13,260	13,254	27,436	35,809	23,760	59,292	45,105	8,200
$\overline{M}$Btu/hr	36.7	--	--	45.33	42.15	21.83	--	9.85	15.73	17.79	--

Exchanger	E-1	E-2	E-3	E-4	E-5	E-6	E-7	E-8	E-9	E-10*
Duty, 10^6 Btu/hr	1.35	1.45	1.79	0.47	4.86	3.30	3.70	4.71	4.08	9.40

*Optional

TABLE B-1-8

ATR-MOLTEN CARBONATE FUEL CELL UTILITY REQUIREMENTS

3,000 Hr/Yr

Gross Power, kw		
Fuel Cell		4,516
Turboexpander*		1,394
Subtotal		5,910
Plant Power Requirement, kw		
Pumps and Compressors		(727)
Inverter Loss		(183)
Subtotal		(910)
Net Output		5,000
Fuel Input, MBtu/hr		36.7
Water, Kgal/yr	3,070	
Reformer Catalyst, ft^3/yr	4	
ZnO, ft^3/yr	75§	
Labor,† hr/yr	750	
Stack Replacement, mil/kwh	2.0	
Other Maintenance, mil/kwh	0.2¶	

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* shaft work expressed in kw.

§ ZnO for 200 ppm feed sulfur.

† Inspection for 3,000 hr/yr operation

¶ 5% of CI for remainder of plant per year.

1.2.3.3 Major System Components

The major system components included for the autothermal reformer are listed in Table B-1-9 along with the principal size parameter used for cost estimating.

In addition to exchangers and vessels, the system includes a catalytic oxidizer (afterburner) to combust the dilute fuel (H_2 and CO) contained in the anode recycle stream. An electrical generator is also included to recover additional expansion work as electricity. The turboexpander and compressors are not included in the list since they are included in the cost of the power section.

The equipment required in these designs was sized on the basis of the flows and temperatures previously discussed. Heat exchangers were sized on the basis of typical heat transfer coefficients for the service involved. The transfer coefficients were selected to reflect the elevated process pressures specified in the designs. Simple handbook criteria were employed in sizing all components, instead of detailed vendor designs of optimized equipment.

1.2.3.4 System Design Issues

In the ATR system the exotherm of partial oxidation supplies preheat and reaction energy requirements. Therefore, an important design consideration* for the ATR system is utilization of the heat (chemical and thermal) contained in the anode vent. The system design used for this assessment incorporates catalytic oxidation of the combustible, reheating in the fuel cell and energy recovery using a turboexpander.

The incorporation of these elements into the system results in a gross power recovery equivalent to 1400 kw. This quantity is greater than needed for air compression by about 675 kw. By not utilizing this available energy, the overall heat rate would be 8490 Btu/kwh rather than 7340 Btu/kwh. Naturally,

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*initially identified in RP-318.

TABLE B-1-9

AUTOTHERMAL REFORMING/MOLTEN CARBONATE CELL FUEL PROCESSOR MAJOR EQUIPMENT

Name	Description	Size	Comments
Exchangers			
E-1	Reformer Air Preheater	80 ft^2	
E-2	Reformer Fuel Vaporizer	60 ft^2	
E-3	Anode Feed Preheater	3575 ft^2	
E-4	Recycle Water Reheater	40 ft^2	
E-5	Recycle Water Vaporizer	495 ft^2	
E-6	Cell Stack Cooler	435 ft^2	
E-7	Anode Recycle-Cathode Exhaust	620 ft^2	
E-8	Anode Exhaust Precooler	1785 ft^2	
E-9	Anode Exhaust Cooler	470 ft^2	
E-10	Expander Exhaust Boilers	3460 ft^2	
Pumps			
P-1	Oil Feed		
P-2	Recycle Water		
Vessels			
V-1	Autothermal Reformer	2.5ϕ x 8 ft	
V-2	ZnO Bed	10ϕ x 15 ft	
V-3	K.O. Drum	3.5ϕ x 10 ft	
Heaters			
H-1	Catalytic Afterburner	13,560 scfm	Cost vs fume incinerator
Catalyst			
	Zinc Oxide	20 ft^3	
	Reformer	10 ft^3	
Generator			
G-1	Electric Generator	675 kw	

there is an economic trade-off between system capital cost and operating cost, but clearly there is a need to utilize a portion of the anode vent energy somehow to obtain an optimized system.

ATR test results with No. 2 fuel indicate that the system may be capable of using virgin distillate fuels containing 0.2 wt. percent sulfur. Unfortunately, the sulfur must eventually be removed from the syngas since the fuel cell is relatively intolerant to sulfur.

At this level of feed sulfur, the cost of sulfur removal by zinc oxide capture is well beyond the break-even cost for other acid gas removal methods. However, the standard H_2S absorption/reduction systems operate best at steady load. Consequently, the application of amine/Claus or Stretford technology to a fuel cell would have to be evaluated to determine its load-following potential. The addition of flow-smoothing concepts may be necessary to produce a workable system.

1.3 SYSTEM COSTS

Like the fuel cell, the fuel processor will be a manufactured, skid-mounted chemical plant, specifically designed to meet the fuel cell power plant performance requirements. As such, the cost of the system in constant dollars should initially decline with the number of units produced. However, the first full-scale 5 Mw fuel cell module has yet to be completed, albeit the FCG-1 demonstrator is currently under construction.

Because this study deals in part with advanced fuel processor systems which have not been fully engineered, a methodology for projecting ultimate system cost was employed. The approach taken was to estimate the capital cost of the fuel processor first using cost methods typically applied to process industry plants.* Following this procedure, base module costs for each of the major components were determined and aggregated to obtain a total fixed plant cost. Cost reductions due to mass production and continued optimization of the mechanical design were generalized on the basis

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*Guthrie, K. M., Process Plant Estimating, Evaluation, and Control, Craftsman Book Company of America, 1974.

of a learning curve mechanism. As a general rule, the unit cost of a new system or component will be reduced by 10-25% for each doubling of production. Selecting the most conservative learning curve in this range of 0.9 exp. $\text{Log}_2 N$ (a 10% reduction for each doubling), the cost of the one-hundredth module would be about 50% of the cost of the original unit.

This methodology was applied to each of the three fuel processor systems previously discussed. The resulting capital investment cost along with estimated operating costs are presented in subsequent sections.

The fuel processor costs which are developed in this report are not directly comparable to the investment costs presented in the RP-318 report. The "processor costs" in RP-318 were for a bare reformer without the thermal management subsystem, the latter being included in the power section cost. The processor costs presented herein include all the integration equipment and thermal management equipment required for the power plant, except for the turbo machinery which is included in the power section cost. Consequently, the processor costs are a greater portion of the total CIF power plant cost than in RP-318.

1.3.1 Steam Reforming

1.3.1.1 Capital Investment for the Fuel Processors

Tables B-1-10 and B-1-11 summarize the component capital investments for the two steam reforming systems, based on process plant cost estimates. Heat exchangers account for the dominant costs of the fuel processor module; 76% of the acid cell system, 85% of the higher temperature molten carbonate system. This dominance suggests that the system designs might be optimized to decrease exchanger costs, perhaps at the expense of fuel efficiency, in order to minimize net power cost. However, optimization efforts were not part of this study, so the capital and operating costs reflect the system designs already discussed. The cost of factory assembled steam reformers is $150/kw and $160/kw for the low and high temperature designs respectively at the conservative end of the range of learning curves.

TABLE B-1-10

STEAM REFORMING/ACID FUEL CELL ESTIMATED CAPITAL INVESTMENT

DISPERSED FUEL PROCESSOR

(5 Mw Net Capacity)

Equipment Items	Installed Price - \$1000 (1975)
Heat Exchangers	1063.4
Reformer	244.8
Pumps	3.3
Knock-out Drums	6.8
Sulfur Removal Bed	2.9
Shift Converter	48.4
Allowance for HDS Unit	35.0
Total Price Fixed Plant*	1404.6
Engineering and Fee, 8%	112.4
Total Price	1516.0
Unit Price, \$/kw	303.2
Unit Price of One-hundredth Manufactured Unit	149.7

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*process plant bare module basis.

Source: Arthur D. Little estimates.

TABLE B-1-11

HIGH TEMPERATURE STEAM REFORMING/MOLTEN CARBONATE ESTIMATED CAPITAL INVESTMENT

DISPERSED FUEL PROCESSOR

(5 Mw Net Capacity)

Equipment Items	Installed Price - $1000 (1975)
Heat Exchangers	1282.3
Pumps	5.6
Knock-out Drum	4.5
Sulfur Removal Bed	4.0
Steam Reformer	206.4
Total Price Fixed Plant*	1502.8
Engineering and Fee, 8%	120.2
Total Price	1623.0
Unit Price, $/kw	324.6
Unit Price of One-hundredth Manufactured Unit	160.2

---

*process plant cost estimates.

Source: Arthur D. Little estimates.

1.3.1.2 Total Capital Investments

Table B-1-12 summarizes the total costs of the two steam reforming designs. The cost assigned to the power section established in the RP-318 assessment has been reduced to reflect that the thermal control equipment has been priced as part of the fuel processor subsystem.

1.3.1.3 Operating and Maintenance Costs

The variable and semi-variable costs (excluding fuel) of operating the two steam reforming/acid cell systems are summarized in Tables B-1-13 and B-1-14. Maintenance costs, which dominate the operating costs, are based on a unit cost of 2.2 mils/kwh. This amount includes amortized stack replacement and annual repairs as itemized in Tables B-1-2 and B-1-5.

1.3.2 Autothermal Reforming

1.3.2.1 Fuel Processor Capital Investment

Using the equipment sizes previously discussed, the capital investment for the ATR system was determined using process plant estimating techniques. The breakdown of costs on this basis is shown in Table B-1-15. The resulting unit cost for the original module is $280/kw in 1975 dollars. The unit cost of a one-hundredth module is estimated to be $140/kw based on the assumed learning curve. By excluding the steam injection system, exchanger E-10 is eliminated and the supplemental generator capacity is reduced by 315 kw. The cost of the processor for the derated plant is $115/kw based on a net output of 4685 kw.

1.3.2.2 Fuel Cell and Ancillaries

The capital investment presented in section 1.3.2.1 is only for the fuel processor and thermal management portion of the fuel cell power plant. The cost of the fuel cell power section and site ancillaries must be added

TABLE B-1-12

TOTAL CAPITAL INVESTMENTS

DISPERSED STEAM-REFORMING FUEL CELL SYSTEMS

5.0 Mw Net Capacity
(1975 Dollars)

Component	Steam Reforming and Phosphoric Acid Cell $/kw	Steam Reforming and Molten Carbonate Cell $/kw
Fuel Processor	150	160
Power Section*	130	130
Site Ancillaries†	50	50
Total	330	340

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*includes: cell stack, inverter, expander-compressors, engineering and fee.

†includes: foundation, fuel storage, passivation and purge gases, and BFW treatment.

Source: Arthur D. Little estimates.

TABLE B-1-13

OPERATING AND MAINTENANCE COSTS

STEAM REFORMING/ACID FUEL CELL

Basis: 3,000 Operating Hours at 100% Load

	Unit	Annual Quantity	\$/Unit	Annual Cost (\$)
Water Makeup	Kgal	0	.40	0
Reformer Catalyst*	ft^3	2	97	190
Shift Converter Catalyst*				
Low Temperature	ft^3	9.8	121	1,190
High Temperature	ft^3	9.3	40	370
ZnO	ft^3	90	75	6,750†
Labor	man-hrs	750	850	6,375†
Maintenance	\$/kwh	--	.0022	33,000
TOTAL				47,875
Unit Cost, mil/kwh	3.19§			

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* assumed catalyst lifetime of 20,000 hrs.

† 200 ppm sulfur fuels.

§ 2.74 mils/kwh for 0.5 ppm sulfur fuels.

TABLE B-1-14

OPERATING AND MAINTENANCE COSTS

STEAM REFORMING/MOLTEN CARBONATE

Basis: 3,000 Operating Hours at 100% Load

	Unit	Annual Quantity	$/Unit	Annual Cost ($)
ZnO	cu ft	90	75.00	6,750*
Labor	hr	750	8.50	6,375
Reformer Catalyst	cu ft	2	97	194
Maintenance and Stack Replacement	$/kwh	--	.0022	33,000
TOTAL				46,319

Unit Cost, mil/kwh 3.09

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*for 200 ppm sulfur fuels.

TABLE B-1-15

AUTOTHERMAL REFORMING/MOLTEN CARBONATE ESTIMATED CAPITAL INVESTMENT

DISPERSED FUEL PROCESSOR

(5 Mw Net Capacity)

Equipment Items	Installed Price - $1000 (1975)
Heat Exchangers	1170.2
Pumps	5.4
Knock-out Drum	4.4
Sulfur Removal Bed	4.4
ATR Reactor (incl. catalyst)	52.7
Catalytic Afterburner	151.7
Electric Generator	16.8
Total Price Fixed Plant*	1405.6
Engineering and Fee, 8%	112.4
Total Price	1518.0
Unit Price, $/kw	303.6
Unit Price of One-hundredth Manufactured Unit, $/kw	149.9

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*process plant basis.

Source: Arthur D. Little estimates.

to this cost to obtain the total installed power plant system price. Using value established in RP-318, the capital cost of the molten carbonate power section is $130/kw in 1975 dollars. Included in this cost are the fuel cell stack, inverter and expander/compressor. The sum of the fuel processor, thermal management and power section costs represents the CIF cost of the preassembled power plant. In addition to this cost the owner must furnish the following site ancillaries: plant foundations, fuel storage, and miscellaneous gases for system purge and passivation. Based on estimates for the FCG-1 demonstrator, the cost of site ancillaries is about $50/kw. The total cost for the molten carbonate/ATR power plant is summarized in Table B-1-16. The total installed plant in 1975 dollars is estimated at $330/kw.

1.3.2.3 Annual O&M Cost

The estimated annual operating and maintenance (O&M) expense for the molten carbonate/ATR power plant is shown in Table B-1-17, based on 3,000 operating hours at 100% load. The catalyst cost for zinc oxide is based on 200 ppm sulfur fuel. The maintenance materials and labor expense is based on a representative figure of 2.2 mil/kwh which includes stack replacement and annual repairs. The total O&M expense is estimated at 3.11 mil/kwh. For a 2000 ppm sulfur fuel the absorbent cost would increase by a factor of 10 assuming a zinc oxide system is used. In this case the total O&M expense increases to 6.5 mil/kwh.

TABLE B-1-16

DISPERSED AUTOTHERMAL-REFORMING FUEL CELL SYSTEM ESTIMATED CAPITAL INVESTMENT

5 Mw Net Capacity

(1975 Dollars)

Component	$/kw
Fuel Processor	150
Power Section*	130
Site Ancillaries†	50
Total Installed Plant	330

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*includes: cell stack, inverter, expander-compressor, engineering and fee.

†includes: foundations, fuel storage, passivation and purge.

TABLE B-1-17

ATR-MOLTEN CARBONATE FUEL CELL OPERATING AND MAINTENANCE EXPENSE

Basis: 3,000 Operating Hours at 100% Load
200 ppm sulfur fuel

	Unit	Annual Quantity	$/Unit	Annual Cost ($)
Water	Kgal	3070	0.40	1228 (optional)
ZnO	cu ft	75	75.00	5625*
Reformer Catalyst	cu ft	4	97.00	388
Labor	hr	750	3.50	6375
Maintenance	$/kwh	--	0.0022	33,000
Total				46,616
Unit Cost, mil/kwh				3.11†

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*$56,250 for 2000 ppm sulfur fuels.

†6.48 mil/kwh for 2000 ppm sulfur fuels.

2.0 BASELOAD SYSTEMS

2.1 Fuel Conversion Technology

Since fuel cells represent a relatively new concept for baseload power plants, Arthur D. Little, Inc., with EPRI's consent, chose to examine the integration of fuel cells with present-day or soon-to-be-available fuel conversion technologies. The purpose was to determine the feasibility and promise of fuel cell development without having to depend on the concurrent development of a suitable advanced coal gasification system. Therefore, the Lurgi and Texaco gasification systems for production of gaseous fuel cell fuels were selected for this analysis.

Because of basic differences in their gasification schemes, Lurgi and Texaco are contrasting technologies in several regards:

- The upper limit on operating pressure is well above 100 atmospheres for the Texaco gasifier since it is slurry-fed. The upper pressure limit is about 30 atmospheres (determined by lock hopper losses) for the Lurgi process which is dry coal-fed.
- Thermodynamics and kinetics in the Lurgi reactor dictate a significatn methane yield. The Texaco process produces very little methane. Since methane remains virtually unconverted in a fuel cell, it could be advantageous to minimize its production.
- Texaco coal gasification is considerably less efficient when run with air; that is, because much of the energy required to heat the nitrogen (in air) must be supplied by oxidizing or burning a portion of the coal within the gasifier, hence the conversion to combustible fuel gases is lessened. Lurgi, on the other hand, is amenable to either air- or oxygen-blowing, since the feed is dry.

An additional fuel conversion technology was chosen for the gasification of heavy oil--Shell Partial Oxidation Process. This process is quite similar to the Texaco Gasification Process and has many of the same advantages (i.e., high pressure, low methane). In addition, this gasifier may be air-blown and operated efficiently if sufficient air preheat is provided.

To examine the contrasting qualities of these processes, we studied: (1) intermediate pressure, air-blown Lurgi; (2) high pressure, oxygen-blown Texaco; and (3) intermediate pressure, air-blown Shell. Significant differences between the first two could then be largely attributed to differences between the gasifier technologies which would bracket the state-of-the-art and more technically advanced options. Differences between the latter two in large part would result from the difference in the oxidant used.

The following sections briefly discuss these gasification technologies.

2.1.1 Process Description - Lurgi Gasification

In the Lurgi Gasification Process (Figure B-2-1), sized* New Mexico coal, steam and compressed air are fed to a bank of parallel moving-bed gasifiers which operate at about 300 psia and 1800°F at the bottom. As it passes down through the bed, the coal successively undergoes partial dehydration, devolatization, and hydro-gasification. Synthesis gas is generated at the bottom of the bed where the devolatized coal is contacted by steam and air on a revolving grate which supports the bed, discharges ash, and distributes the incoming steam and air.

The crude gas leaving the gasifier is then directed to a quench scrubber, where it is washed with a stream of process condensate. This washing process quenches the gas to approximately 350°F and simultaneously condenses the high boiling tar fractions. Coal and ash dust are also removed with the condensed tar, thereby leaving the quenched effluent gas essentially free of particulate matter.

The gasifier quench scrubber effluent is then cooled in a series of heat exchangers to approximately 100°F. The condensate resulting from this gas cooling is directed to a tar oil separation unit where the

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*The economic estimates include sizing the coal but no provision for disposal or sale of the fines.

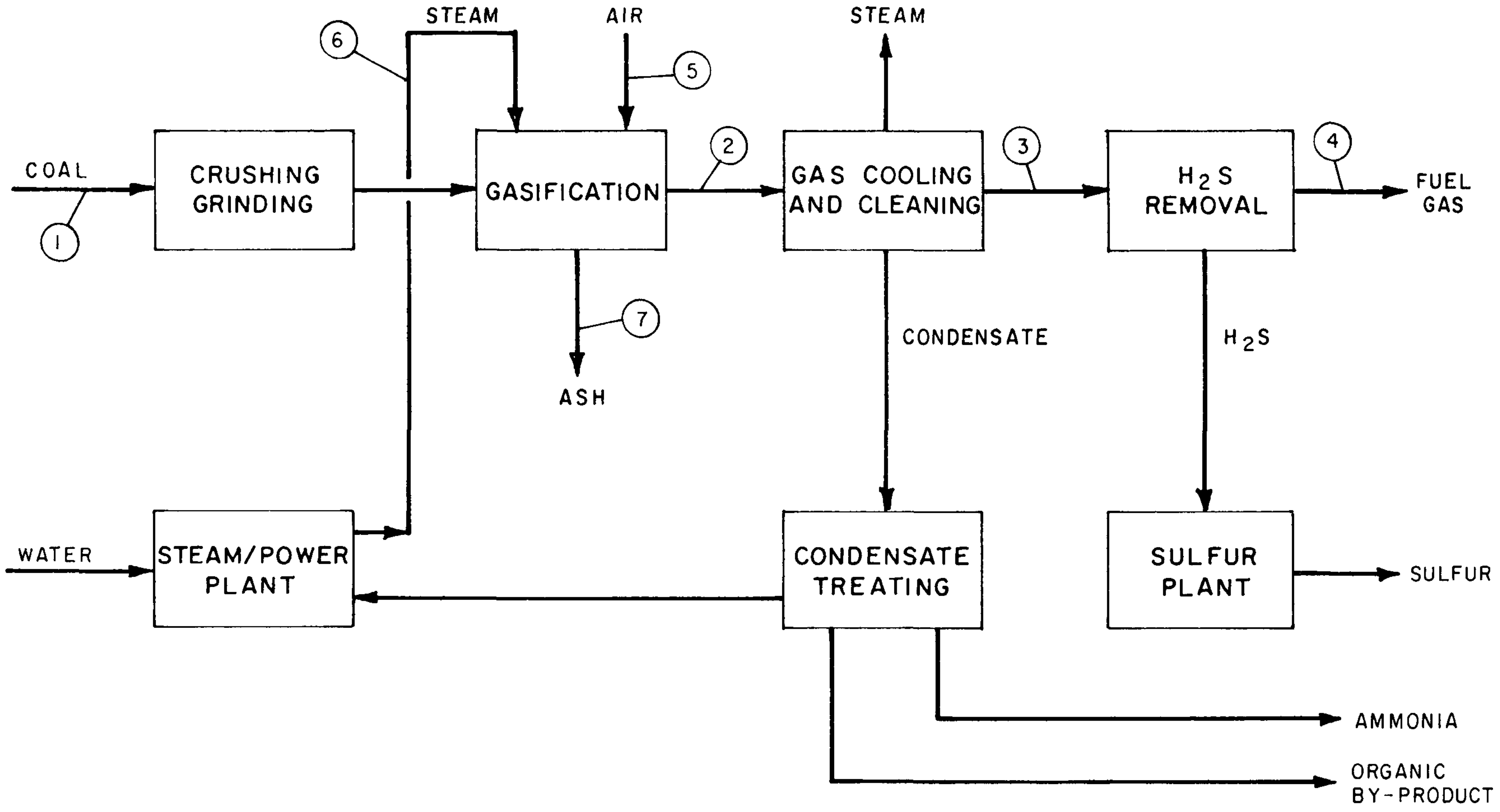

FIGURE B-2-1. LURGI COAL GASIFICATION

condensate is separated into its tar, oil and gas liquor fractions by distillation. The tar and oil fractions are pumped to liquid fuel storage tanks, while the gas liquor fraction is further processed through a phenol extraction unit. The crude phenols recovered are then also pumped to the liquid fuels storage tank and are combined with the tar and oil fractions, constituting a liquid hydrocarbon byproduct with appreciable byproduct fuel value. The dephenolized condensate is then fed to an acid gas and ammonia recovery system, which results in the production of both elemental sulfur and anhydrous ammonia and subsequently allows the remaining process condensate to be treated in a biological treating unit to be reused as cooling tower makeup water.

The cooled gas leaving the water coolers at ambient temperature then flows to an acid gas absorber, where hydrogen sulfide is absorbed in order to reduce its level in the treated gas to less than 5 ppm, as appropriate for feed to the fuel cell. The treated gas is then sent to a knock-out drum to minimize solvent losses, while the acid gas from the regenerator is further processed through a sulfur recovery unit where liquid sulfur is processed and subsequently pumped to storage tanks. The clean fuel gas leaves the acid gas absorber at near ambient temperature.

2.1.2 Process Description - Texaco Gasification

The Texaco Coal Gasification Process is a high pressure, slurry fed, partial oxidation system. In principle, it is the same process widely used for partial oxidation of hydrocarbons (Figure B-2-2).

Pulverized coal is slurried with water and pumped to the gasifier, which is essentially a burner where pulverized coal is burned to CO and CO_2. Since the reactor effluent is at more than 2000°F, the equilibrium formation of methane is not favored, and even at high pressure, very little methane (less than 1%) is present in the effluent gas. This is advantageous for use in a fuel cell, since methane remains essentially unconverted in the fuel cell. Since the water slurry fed to the Texaco gasifier contains an excess of water above that required for soot suppression, no steam is added.

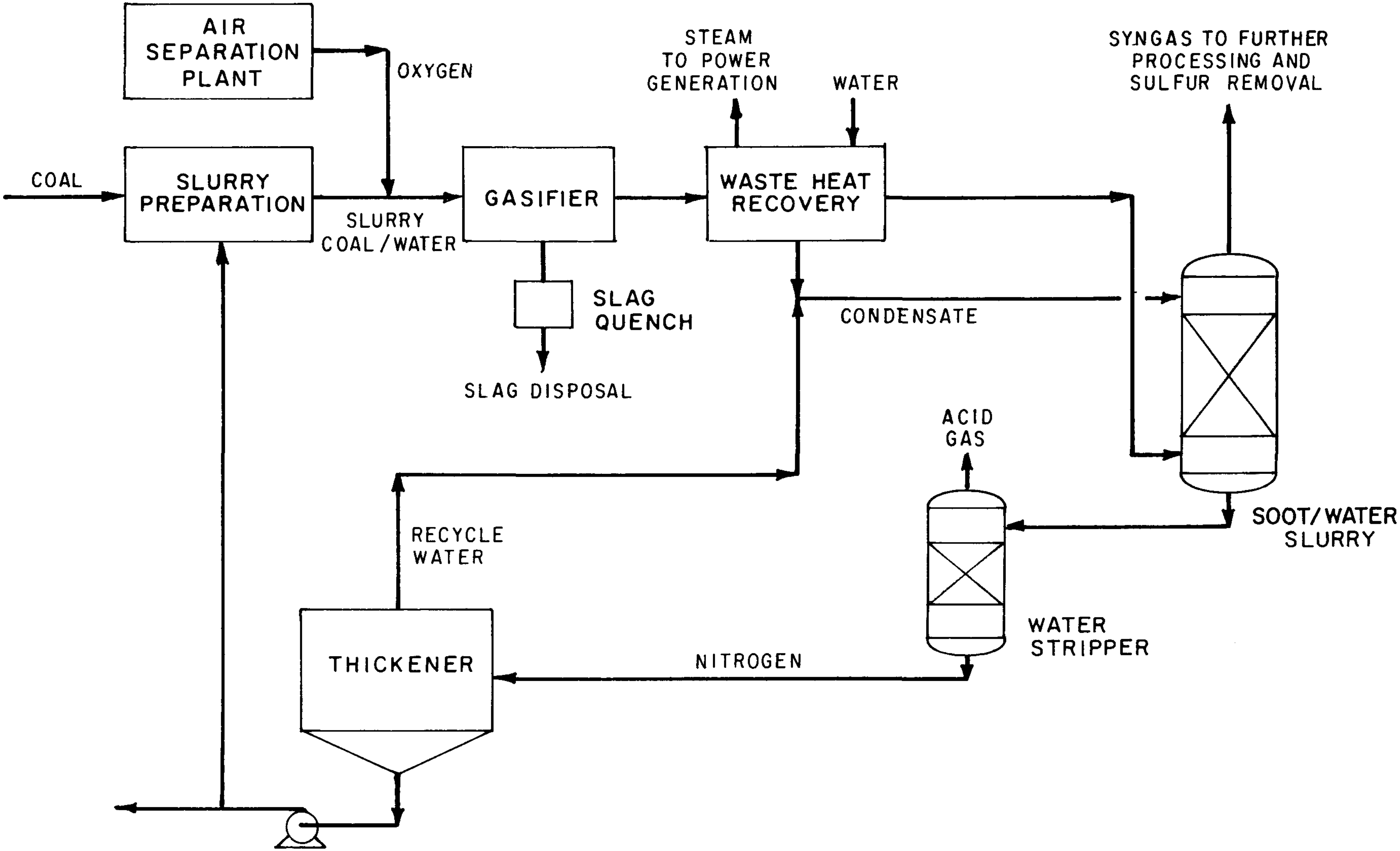

FIGURE B-2-2. TEXACO COAL GASIFICATION.

The gasifier effluent is cooled by heating a variety of process streams and raising steam for byproduct power generation in a bottoming cycle. Upon leaving the gasifier, the raw gas stream is partially quenched by evaporating water which is injected to reduce the gas temperature to a level acceptable for entry into a gas reheat exchanger. Eventually, the raw gas is cooled to a temperature below the dew point of the water. The main purpose for quenching the gas is to recover the unconverted carbon and ash from the gas and to produce a solid-free gas. Also, recycling the soot to the gasifier enhances the overall carbon conversion (usually greater than 99%). The soot/water slurry is stripped of acid gas (H_2S, COS) and sent to a thickener where solids are concentrated for recycle.

While the gas is cool, it is scrubbed in an acid gas removal unit (Rectisol) to reduce the H_2S content to below the 5 ppm tolerable limit for the molten carbonate fuel cell.

2.1.3 Process Description - Shell Gasification

In the Shell Gasification Process (Figure B-2-3), vacuum residue and air are preheated to approximately 500°F and fed along with steam into the Shell gasifier through a specially-designed combustor nozzle. The nozzle atomizes the vacuum residue and mixes the hydrocarbon feed, steam, and air as they enter the combustion zone. The gasifier is operated at a pressure of 250 psia and a temperature in the range of 2200-2800°F.

Most of the heat generated in the gasifier is contained in the hot gases (2400°F) leaving the reactor. Therefore, in order to achieve economical thermal efficiencies, this heat is recovered in a waste heat boiler by generating high pressure steam, which is used not only as a feed source to the gasifier but also is used in a steam bottoming cycle to drive the large air compression equipment and to produce byproduct electric power. The product gases leaving the waste heat boiler at a temperature moderately above the generated steam temperature (700°F) also heat incoming boiler feedwater. As a result, the gas is cooled to near its dew point.

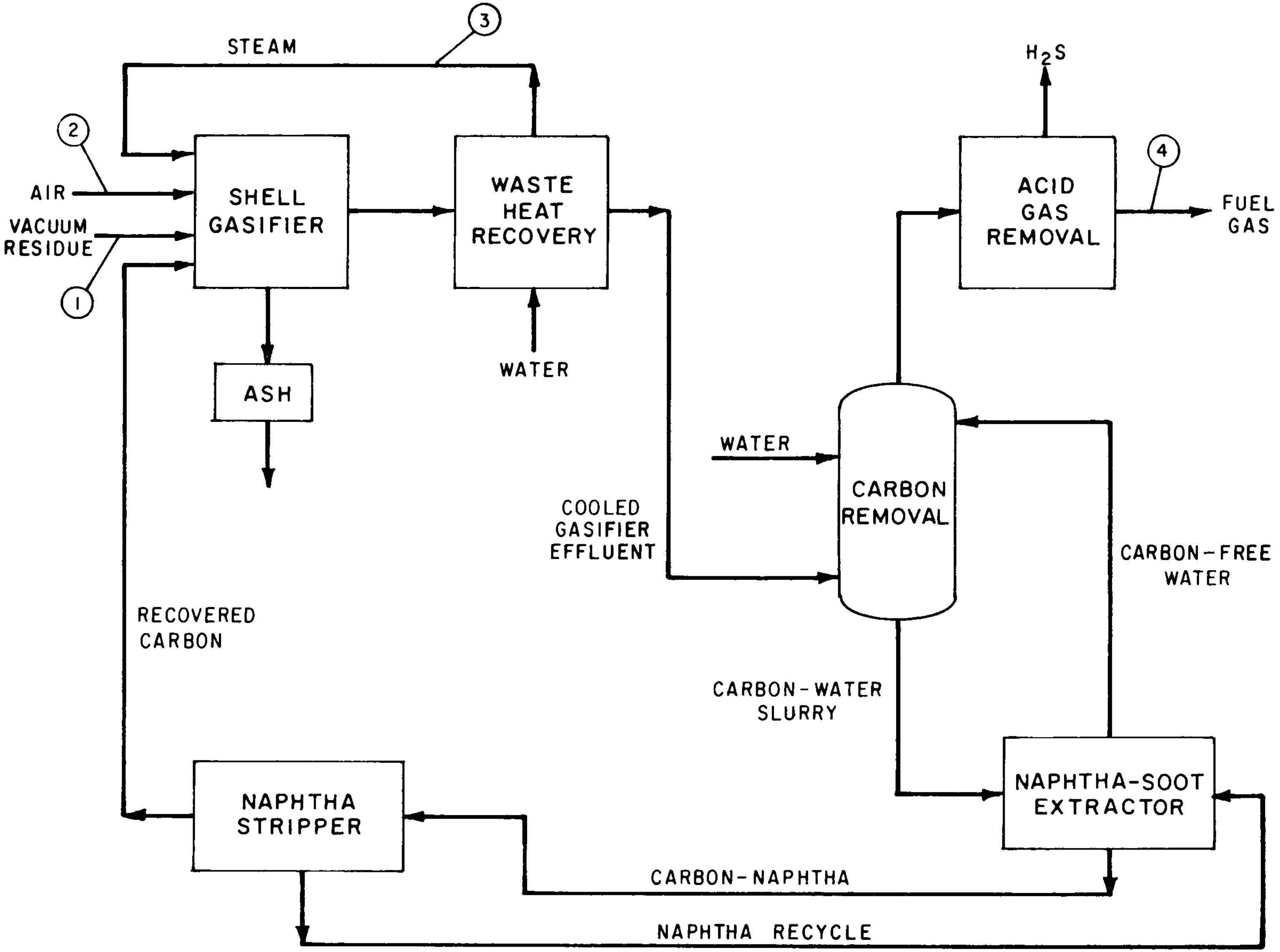

FIGURE B-2-3. SHELL GASIFICATION PROCESS

The gas leaving the boiler feedwater heaters next enters a carbon removal section to eliminate both the soot and ash formed in the gasifier. This is accomplished by scrubbing the gas of these solids, first in a quench vessel with spray nozzles and then in a scrubber column. Recovery of the carbon from the resultant carbon-water slurry is achieved by the Shell Pelletizing Process. In this process the carbon-water slurry is contacted with a low viscosity oil, where the oil preferentially wets the soot particles and forms pellets which are homogenized into the vacuum residue for return to the gasification reactor.

The fuel gas leaving the carbon recovery section at approximately 100°F then enters a Sulfinol solvent absorber. Here, sulfur removal to less than 1 ppm in the treated fuel gas is achieved. Simultaneously, the H_2S from the Sulfinol solvent regeneration step is fed to a Claus plant where elemental sulfur is produced. To minimize sulfur emissions from the Claus plant, since the tail gas from the Claus plant still contains some sulfur compounds, the tail gas is fed to a Shell Claus Off-Gas Treating (SCOT) Process, where most all of this sulfur is recovered and recycled back to the Claus plant. The clean fuel gas leaves the acid gas absorber at approximately 100°F.

2.2 Fuel Cell Integration

Table B-2-1 shows generalized energy balances for the three gasification technologies considered. Since the energy quantities tabulated show the forms in which the energy leaves the gasifier per Btu of fuel input, the table gives insight into the relative sizes of the energy conversion sections which generate electric power. For example, much of the energy leaving the Lurgi reactor is in the form of hydrocarbons (including methane) which cannot be utilized directly in a fuel cell. Therefore, these combustibles would have to be burned for power generation via a steam or combined power cycle (representing a substantial portion of output power) if electricity production is to be maximized. Alternatively, they can be chemically converted, at additional expense, to H_2 and/or CO for use in a fuel cell. Much of the energy leaving the

TABLE B-2-1

ENERGY BALANCES FOR GASIFICATION TECHNOLOGIES

	Lurgi	Texaco	Shell
Fuel	Coal	Coal	Oil
Oxidant	Air	O_2	Air
Gasifier Effluent, Btu/Btu in Fuel			
H_2 + CO	.59	.72	.66
CH_4	.18	.02	.01
Total Cold Gas	.77	.74	.67
Byproducts	.09	---	---
Sensible Heat	.11	.20	.31
Total	.97	.94	.98

Texaco gasifier is already in the form of CO + H_2 which enhances the potential for relatively high efficiency conversion of the gas to power via a fuel cell. The sensible heat of the gasifier effluent above a reference state of 77°F represents high, intermediate, and low quality heat which may be used in a bottoming steam cycle for incremental power generation. Since Lurgi is a moderate temperature process, its sensible heat content is relatively small. The major difference between the Texaco and Shell efficiencies relates to the sensible heat of the large inert diluent content (N_2) of the effluent stream of the Shell process (at above 2200°F) because it is air-blown rather than oxygen-blown.

The two fuel cell technologies considered are based on molten carbonate and phosphoric acid electrolytes. A summary of the important projected operating requirements and characteristics of these fuel cells is presented in Table B-2-2. Of particular importance are the facts that the molten carbonate cell: (1) operates at high temperature, which renders its waste heat of high quality, and (2) can utilize both hydrogen and carbon monoxide fuel, while the acid cell can utilize only hydrogen.

2.2.1 Gasifier/Fuel Cell Combinations

Integration of the gasification technologies described above importantly involves the optimal transfer of heat between the major sections of the plant in order to maximize the quantity and quality of heat which can be made available to a bottoming steam system for electric power production above that produced by the fuel cell.

Because advantages appear *a priori* to favor the molten carbonate cell for baseload systems, our analysis emphasizes this technology. However, we did investigate power systems using the phosphoric acid cell under what appear to be its most favorable conditions, namely integrating it with the Texaco Process (oxygen-blown). Design considerations contributing to this selection include low methane yield, low gas volume (no diluent N_2 from air) keeping the shift converters as small as possible, and no high pressure steam requirement for the gasifier, which leaves a large excess of low level heat (300°F) from the fuel cell for heat recovery.

TABLE B-2-2

REQUIREMENTS AND CHARACTERISTICS OF BASELOAD FUEL CELL SYSTEMS

Fuel Cell Type	Molten Carbonate	Phosphoric Acid
Fuel Utilized	H_2, CO	H_2
Fuel Utilization	85% max.	90% max.
Fuel Impurities (max.)[*]		
H_2S	5.0 ppm	200 ppm
Chlorides	0.1 ppm	0.1 ppm
Ammonia	0.1 ppm	0.1 ppm
C_2^+	100 ppm	100 ppm
CO	N.A.	4%
Oxidant Stoichiometry	2:1	2:1
Inlet Temperature	1100°F (min.)	325°F (min.)
Outlet Temperature	1300°F (max.)	375°F (max.)
Operating Pressure[†]	35-110 psia	65 psia

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[*] Maximum fuel impurity limits have not been fully established quantitatively. The limits listed are conservative based on available data.

[†] Selected for this study. The operation of fuel cells is not necessarily limited to these ranges.

Source: EPRI and Arthur D. Little, Inc.

Table B-2-3 summarizes the six cases selected for study. The Lurgi gasifier was integrated with the molten carbonate system using two coals as a basis for comparison. Similarly the Texaco and molten carbonate systems were integrated using both bituminous and subbituminous coals. In the other bituminous coal case, Texaco and phosphoric acid technologies were considered; here it was assumed that any relative difference using subbituminous coal would be similar to the differences evident in the molten carbonate cases. Finally, the Shell gasifier was used to convert heavy oil to fuel gas for use in a molten carbonate fuel cell.

2.2.1.1 Molten Carbonate Fuel Cell Integration

For each given gasification section characteristic (gas composition and temperature), the objective was to maximize the overall power output of the plant including both the fuel cell and the bottoming steam cycle. Since the gasification sections were generally under pressure, there was also opportunity for power recovery of pressure energy prior to entering the fuel cell. As a result, each molten carbonate integration analysis was largely one of heat management. Where differences among the cases existed, they were generally variations in heat transfer schemes as a result of differing energy needs and availabilities in the gasification section.

Table B-2-4 lists the considerations which dictate the flow configuration in the molten carbonate fuel cell section of the plant. First, the carbonate ion which leaves the electrolyte on the anode side of the fuel cell as CO_2 (one mole $CO_3^=$ per mole of fuel converted) must be replenished as CO_2 on the cathode side. This can be accomplished by passing the anode vent to the cathode to recover the necessary CO_2. An alternate scheme, which was not evaluated in this study, is to recover the CO_2 from the anode vent by scrubbing with a solvent such as potassium carbonate and adding the CO_2 gas directly to the cathode without the other constituents of the anode vent.

Second, the anode vent contains a considerable quantity of combustibles, namely the unutilized CO and H_2 as well as any CH_4 which cannot be converted to any significant degree in the fuel cell. If this anode vent is to be

TABLE B-2-3

BASELOAD FUEL CELL SYSTEMS STUDIES

Gasifier Technology	Fuel Cell Type	Fuel
Lurgi Air Blown	Molten Carbonate	Illinois No. 6 Coal
Lurgi Air Blown	Molten Carbonate	New Mexico Coal
Texaco O_2 Blown	Molten Carbonate	Illinois No. 6 Coal
Texaco O_2 Blown	Molten Carbonate	New Mexico Coal
Texaco O_2 Blown	Phosphoric Acid	Illinois No. 6 Coal
Shell Air Blown	Molten Carbonate	Vacuum Residue

TABLE B-2-4

SPECIAL CONSIDERATIONS (DESIGN CONSTRAINTS) IN MOLTEN CARBONATE FUEL CELL INTEGRATION

Issue	Result
Need to Replenish $CO_3^=$	Run Anode Vent to Cathode
Combustibles in Anode Vent	Burn Anode Vent (WHB)
Adiabatic Fuel Cell Operation	Recirculating Cathode Vent
Fuel Cell Pressure Limit	Power Recovery Turbine(s)

directed to the fuel cell cathode along with the required air, these combustibles must first be oxidized in a controlled fashion, such as in a waste heat boiler, in order to avoid the possibility of an explosive mixture entering the fuel cell.

Third, the molten carbonate fuel cell is designed to run adiabatically. That is, unlike the phosphoric acid cell which has an integral heat exchanger for removal of resistance heat losses, the molten carbonate stack is envisioned without such an exchanger because of the higher operating temperatures and more difficult corrosion problems associated with the molten salt. As a result, all heat must be removed from the fuel cell by heat added to the anode and cathode streams. The lower (1100°F) and upper (1300°F) temperature limits on these streams are fixed by considerations of cell materials and performance, and electrolyte loss rates. Within these limits, the heat removal requirement is not normally met on a once-through basis. Therefore, either the anode and/or the cathode vent must be recirculated in sufficient quantity to effect the proper heat removal rate. The recirculated stream acts as an intermediate heat transfer fluid as it is cooled externally such as by generating steam in a waste heat boiler.

Finally, since fuel cells are presently envisioned to run only at low to moderate pressures, coupling them with many higher pressure gasifiers may indicate a need for power recovery turbines. In many situations, high temperature turbines (e.g., 1500°F or greater) are necessary to maximize the energy output of the plant.

2.2.1.1.1 Process Description and Flow Diagrams

The process description below discusses the general plan of heat management in the fuel cell area. Specific differences among the schemes are shown in the flowsheets of Figures B-2-4 through B-2-8.

The clean gas leaving the gasification and acid gas removal sections is reheated to 1500°F by exchange with the gasifier effluent and enters a gas turbine for recovery of pressure energy by expansion to the operating pressure of the fuel cell. The gas is further reheated to 1100°F prior

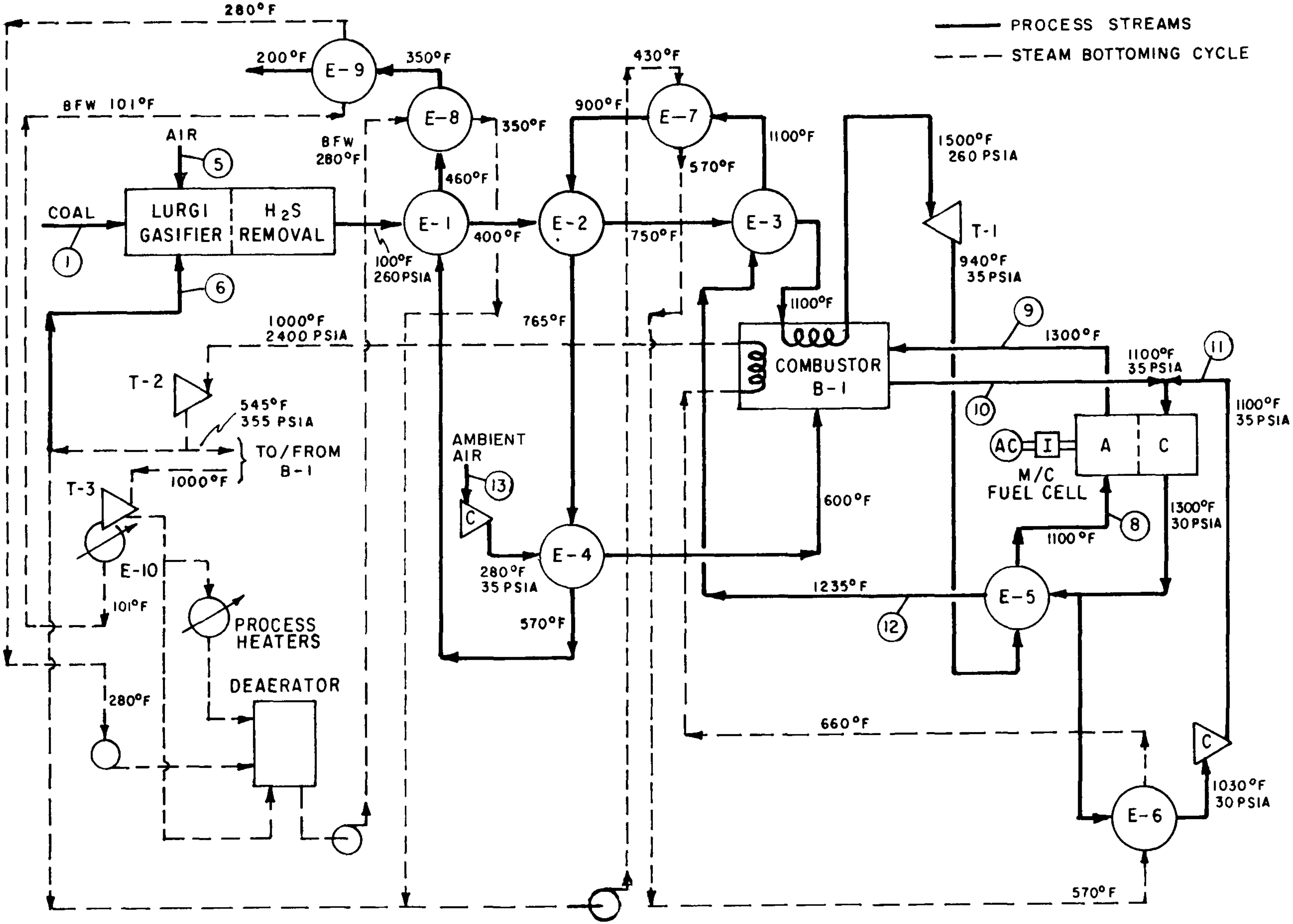

FIGURE B-2-4. BASELOAD FUEL CELL POWER PLANT LURGI GASIFIER WITH MOLTEN CARBONATE FUEL CELL, (ILLINOIS COAL).

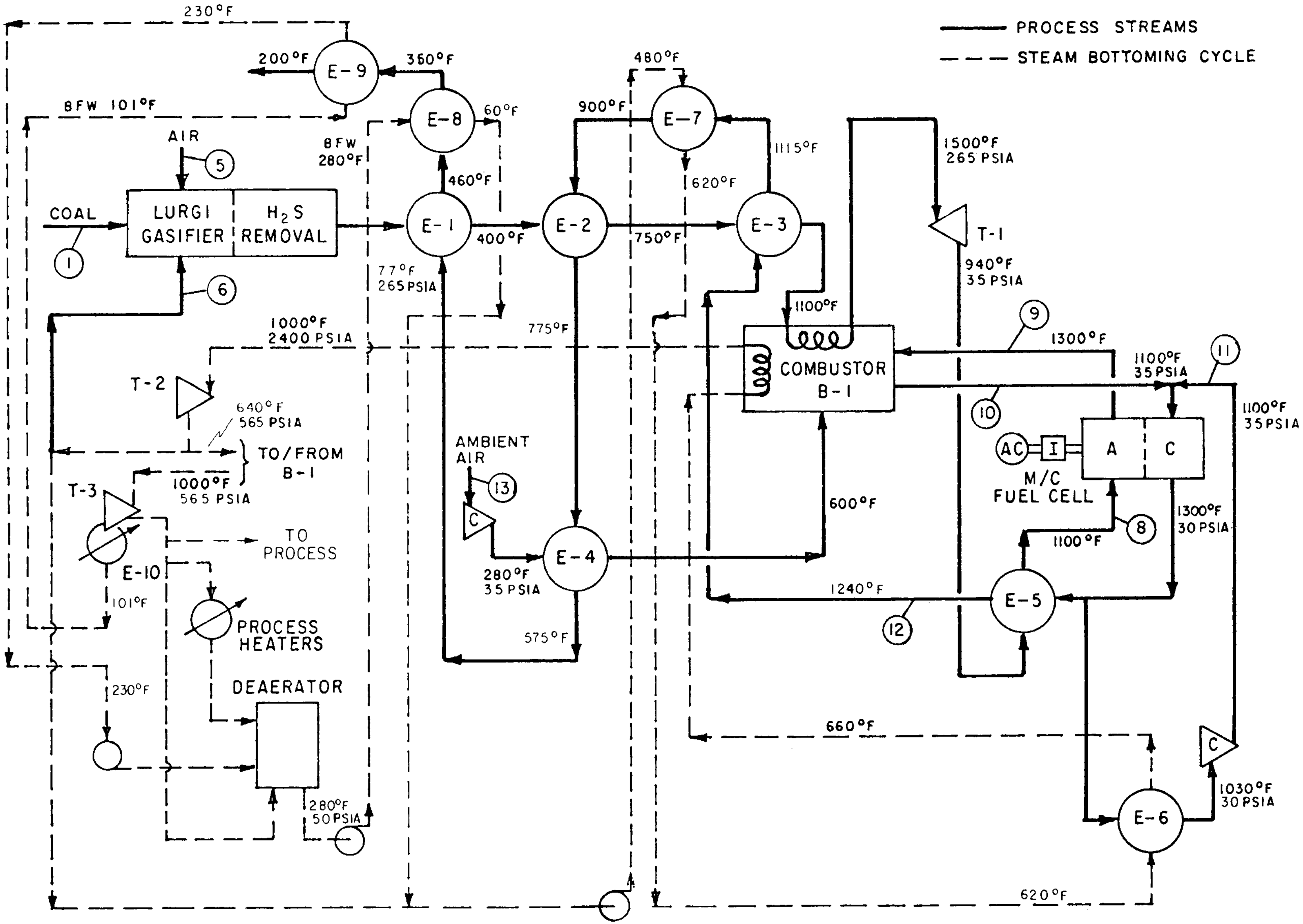

FIGURE B-2-5. BASELOAD FUEL CELL POWER PLANT LURGI GASIFIER WITH MOLTEN CARBONATE FUEL CELL, (NEW MEXICO COAL).

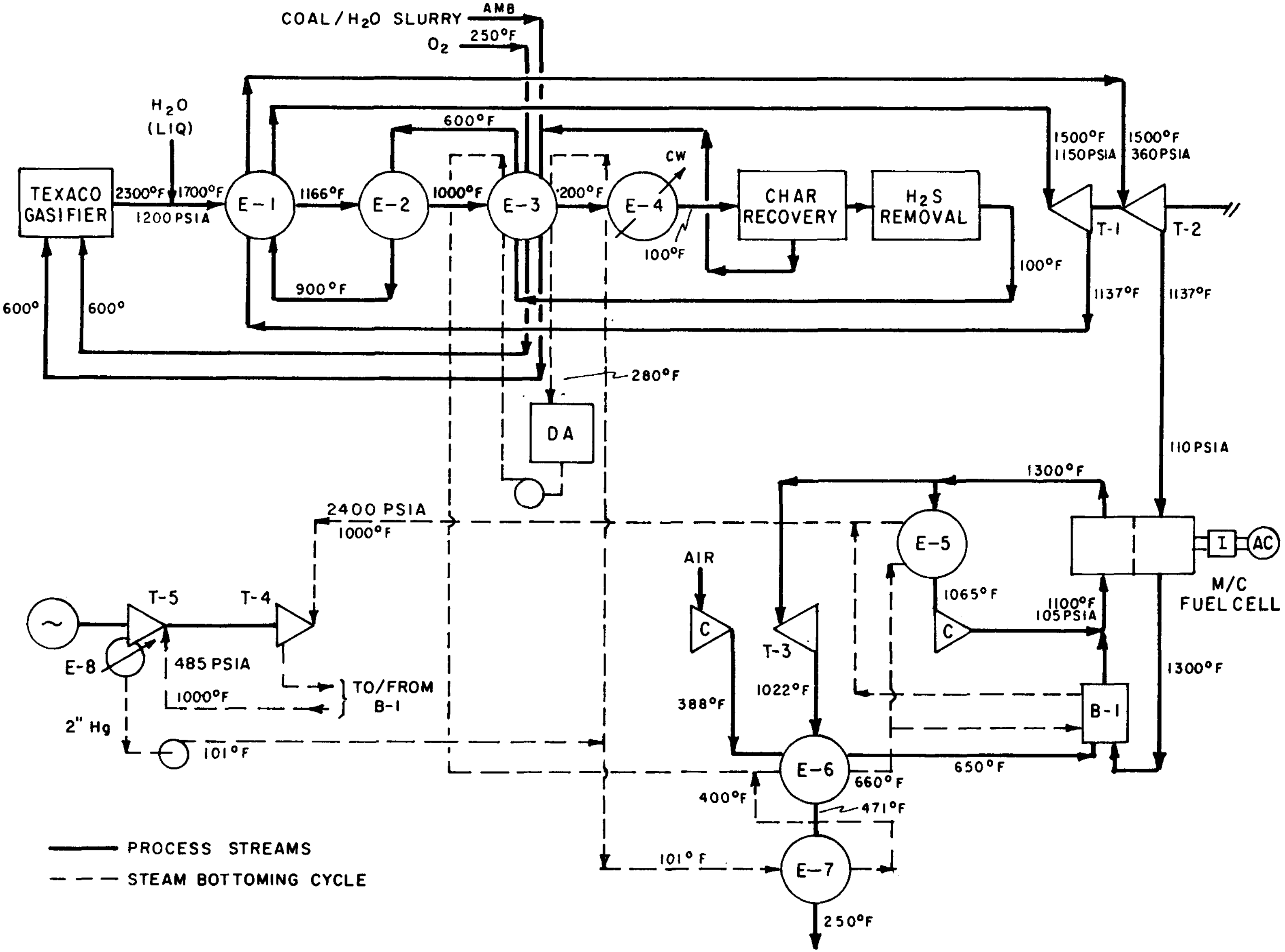

FIGURE B-2-6. BASELOAD FUEL CELL POWER PLANT. TEXACO GASIFIER WITH MOLTEN CARBONATE FUEL CELL, (ILLINOIS COAL).

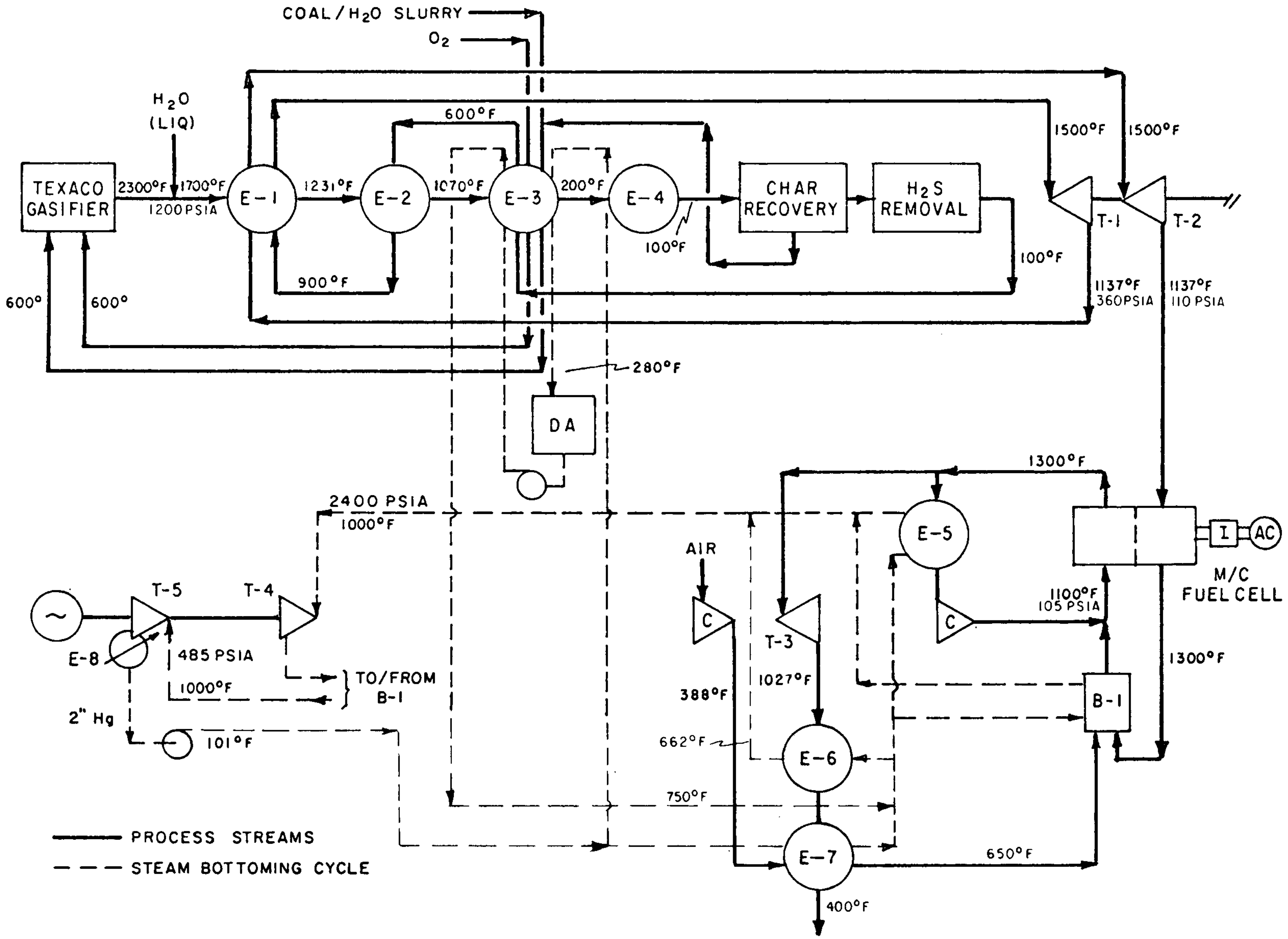

FIGURE B-2-7. BASELOAD FUEL CELL POWER PLANT. TEXACO GASIFIER WITH MOLTEN CARBONATE FUEL CELL, (NEW MEXICO COAL).

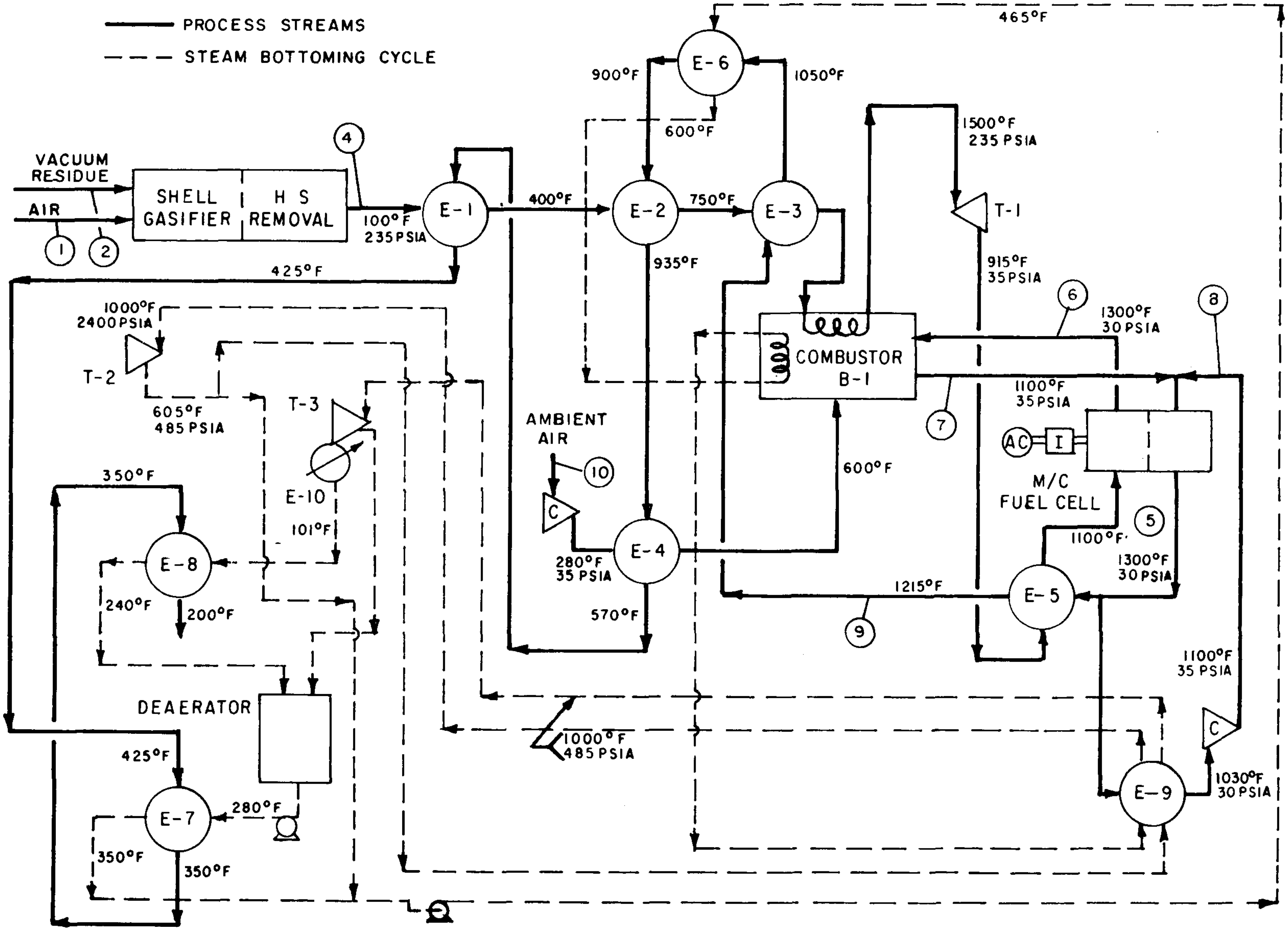

FIGURE B-2-8. BASELOAD FUEL CELL POWER PLANT SHELL GASIFIER WITH MOLTEN CARBONATE FUEL CELL VACUUM RESIDUE.

to entering the fuel cell, where the chemical energy of its major components, H_2 and CO, is converted directly into electrical energy via electrochemical oxidation.

The fuel cell anode vent at 1300°F combines with compressed air heated to 600°F and enters a burner, where its unconverted combustible stream components, H_2, CO and CH_4, are completely oxidized in a waste heat boiler to form carbon dioxide and water vapor. The burner effluent, cooled to near 1100°F, then enters the fuel cell's cathode side, along with additional air as necessary to supply O_2 and CO_2 for the formation of carbonate ion ($CO_3^=$) to restore the electrolyte.

Most of the fuel cell's cathode effluent serves as the fuel cell cooling medium as most of it is cooled from 1300°F and recycled back to the cathode side at 1100°F. The cooling of this cathode recycle stream provides the mechanism for converting waste heat from the fuel cell to high pressure steam, which may be used in a steam bottoming cycle. The unrecycled portion of the cathode vent is bled through several heat exchangers to preheat the fuel cell feed gas and to generate additional steam for the bottoming cycle.

2.2.1.1.2 Mass Balances and Utility Summary

The mass balances for the molten carbonate baseload systems are summarized in Tables B-2-5 to B-2-9. The coal-based power plant capacities are based on 10,000 tons per day of fuel and the oil-based plant on 10,000 bbl/day. Mass balance quantities are expressed in mols per hour. Two material balances are included for the Lurgi and Texaco cases; one each for Illinois and New Mexico coals.

Overall energy balances for each system are provided in Tables B-2-10 to B-2-14. The systems incorporating the Texaco gasifier are the more efficient of the coal-based designs. The heat rate obtained with Illinois No. 6 coal is greater than that for New Mexico coal. The resulting heat rates are as follows:

TABLE B-2-5

MATERIAL BALANCE
LURGI/MOLTEN CARBONATE/ILLINOIS NO. 6 COAL

Stream No.	1 Coal Feed #/hr	2 Gasifier Effluent #mols/hr	3 Outlet Gas Cooling #mols/hr	4 Acid Gas Removal #mols/hr	5 Gasifier Air #mols/hr	6 Gasifier Steam #mols/hr	7 Ash #mols/hr
Component Flows							
CH_4		4,837	4,837	4,800			
CO		20,271	20,271	20,239			
CO_2		16,479	16,479	12,944			
H_2		29,141	29,141	29,130			
H_2O	35,000	47,089	441	275	943	65,976	
N_2		50,295	50,295	50,262	50,264		
H_2S		927	927	148			
O_2					13,362		
Ash	80,000						80,000
DAF Coal	718,333						3,333
Hydrocarbons							
NH_3		606					
C_2^+		120	120	112			
TOTAL	833,333	169,765	122,511	117,910	64,569	65,976	83,333

Stream No.	8 Anode Feed #mols/hr	9 Anode Vent #mols/hr	10 Combustor Effluent #mols/hr	11 Cathode Recycle #mols/hr	12 Cathode Vent #mols/hr	13 Air Stream #mols/hr
Component Flows						
CH_4	4,748	4,748				
CO	20,132	3,020				
CO_2	12,578	71,495	79,484	91,158	128,837	
H_2	29,049	4,357				
H_2O	269	24,961	39,101	94,597	133,697	
N_2	50,095	50,095	202,691	490,379	693,071	152,596
H_2S						
O_2			27,015	14,789	20,902	40,564
Ash						
DAF Coal						
Hydrocarbons						
NH_3						
C_2^+	110	110				
TOTAL	116,981	158,786	348,291	690,923	976,507	193,160

Heat Exchanger Duties, $\overline{\text{M}}$Btu/hr

E-1	E-2	E-3	E-4	E-5	E-6	E-7	E-8	E-9	E-10	B-1
258	320	335	442	158	1595	480	256	322	2227	2220

TABLE B-2-6

MATERIAL BALANCE
LURGI/MOLTEN CARBONATE/NEW MEXICO COAL

Stream No.	1	2	3	4	5	6	7
	Coal Feed #/hr	Gasifier Effluent #mols/hr	Outlet Gas Cooling #mols/hr	Acid Gas Removal #mols/hr	Gasifier Air #mols/hr	Gasifier Steam #mols/hr	Ash #/hr
Component Flows							
CH_4		3,786	3,786	3,786			
CO		13,006	13,006	13,006			
CO_2		11,053	11,053	8,676			
H_2		17,337	17,337	17,337			
H_2O	135,395	26,501	186	120	288	28,777	
N_2		28,708	28,708	28,708	29,106		
H_2S		173	173				
O_2					7,714		
Ash	160,440						160,440
DAF Coal	537,498						8,440
Hydrocarbons							
NH_3		445					
C_2^+		469	469	469			
TOTAL	833,333	101,478	74,718	72,102	37,108	28,777	168,880

Stream No.	8	9	10	11	12	13
	Anode Feed #mols/hr	Anode Vent #mols/hr	Combustor Effluent #mols/hr	Cathode Recycle #mols/hr	Cathode Vent #mols/hr	Air Stream #mols/hr
Component Flows						
CH_4	3,786	3,786				
CO	13,006	1,951				
CO_2	8,676	45,522	52,197	56,681	83,087	
H_2	17,337	2,601				
H_2O	120	14,854	26,248	56,340	82,588	
N_2	28,708	28,708	135,513	290,872	426,385	106,805
H_2S						
O_2			16,994	8,797	12,895	28,392
Ash						
DAF Coal						
Hydrocarbons						
NH_3						
C_2^+	469	469				
TOTAL	72,102	97,891	230,952	412,690	604,955	135,197

Heat Exchanger Duties, $\overline{M}$Btu/hr

E-1	E-2	E-3	E-4	E-5	E-6	E-7	E-8	E-9	E-10	B-1
173	199	209	309	101	952	347	166	219	1862	1544

TABLE B-2-7

MATERIAL BALANCE
TEXACO/MOLTEN CARBONATE/ILLINOIS NO. 6 COAL

Stream No.	1 Coal/Water Slurry #/hr	2 Oxidant Feed #mols/hr	3 Gasifier Effluent #mols/hr	4 Quench Water #mols/hr	5 Cleaned Gas #mols/hr	6 Anode Vent #mols/hr
Component Flows						
CH_4			569		569	569
C_2^+						
CO			32,083		32,083	4,813
CO_2			13,134		12,215	90,771
H_2			28,253		28,253	4,238
H_2O	681,667		28,765	18,440		24,015
N_2		111	468		468	468
O_2		22,123				
H_2S			963			
MAF Coal, Oil	718,333					
Ash	80,000					
TOTAL	1,480,000	22,234	104,235	18,440	73,588	124,874

Stream No.	7 Fuel Cell Air #mols/hr	8 Cathode Vent #mols/hr	9 Fuel Cell Recycle #mols/hr	10 Vent Gas #mols/hr	11 Ash #/hr
Component Flows					
CH_4					
C_2^+					
CO					
CO_2		270,247	225,380	44,867	
H_2					
H_2O		177,028	147,637	29,391	
N_2	134,432	809,661	674,761	134,900	
O_2	35,564	25,641	21,384	4,257	
H_2S					
MAF Coal, Oil					5,550
Ash					80,000
TOTAL	169,996	1,282,577	1,069,162	213,415	85,550

Heat Exchanger Duties, $\overline{M}$Btu/hr

E-1	E-2	E-3	E-4	E-5	E-6	E-7	E-8	B-1
600	177	1603	120	2499	1045	376	3300	1057

TABLE B-2-8

MATERIAL BALANCE
TEXACO/MOLTEN CARBONATE/NEW MEXICO COAL

Stream No.	1 Coal/Water Slurry #/hr	2 Oxidant Feed #mols/hr	3 Gasifier Effluent #mols/hr	4 Quench Water #mols/hr	5 Cleaned Gas #mols/hr	6 Anode Vent #mols/hr
Component Flows						
CH_4			203		203	203
C_2^+						
CO			21,158		21,158	3,174
CO_2			12,458		11,586	64,497
H_2			19,933		19,933	2,990
H_2O	625,000		29,204	15,209		16,943
N_2		88	342		342	342
O_2		17,634				
H_2S			180			
MAF Coal, Oil	537,500					
Ash	160,000					
TOTAL	1,322,500	17,722	83,478	15,209	53,222	88,149

Stream No.	7 Fuel Cell Air #mols/hr	8 Cathode Vent #mols/hr	9 Fuel Cell Recycle #mols/hr	10 Vent Gas #mols/hr	11 Ash #/hr
Component Flows					
CH_4					
C_2^+					
CO					
CO_2		191,084	158,137	32,947	
H_2					
H_2O		117,961	97,662	20,339	
N_2	90,580	527,405	436,483	90,922	
O_2	23,963	17,463	14,452	3,011	
H_2S					
MAF Coal, Oil					4,100
Ash					160,000
TOTAL	114,543	853,913	706,734	147,219	164,100

Heat Exchanger Duties, $\overline{M}$Btu/hr

E-1	E-2	E-3	E-4	E-5	E-6	E-7	E-8	B-1
397	131	1734	107	1662	424	387	2390	595

TABLE B-2-9

MATERIAL BALANCE
SHELL/MOLTEN CARBONATE/VACUUM RESIDUE

Stream No.	1	2	3	4	5	6	7
	Vacuum Residue bbl/hr	Gasifier Air #mols/hr	Gasifier Steam #mols/hr	Acid Gas Removal #mols/hr	Anode Feed #mols/hr	Anode Vent #mols/hr	Combustor Effluent #mols/hr
Component Flows							
CH_4				80	80	80	
CO				9,469	9,469	1,420	
CO_2				401	401	21,444	22,944
H_2				5,818	5,818	873	
H_2O			46,296			4,945	5,978
N_2		24,290		24,354	24,354	24,354	59,318
H_2S							
O_2		6,457					7,988
Ash							
DAF Coal							
Hydrocarbons							
NH_3							
C_2^+							
TOTAL	417	30,747	46,296	40,122	40,122	53,116	120,066

Stream No.	8	9	10	11	12	13
	Cathode Recycle #mols/hr	Cathode Vent #mols/hr	Air Stream #mols/hr			
Component Flows						
CH_4						
CO						
CO_2	33,407	43,357				
H_2						
H_2O	20,071	26,049				
N_2	199,162	258,480	34,964			
H_2S						
O_2	5,006	6,497	9,295			
Ash						
DAF Coal						
Hydrocarbons						
NH_3						
C_2^+						
TOTAL	257,646	334,383	44,259			

Heat Exchanger Duties, $\overline{M}$Btu/hr

E-1	E-2	E-3	E-4	E-5	E-6	E-7	E-8	E-9	E-10	B-1
85	102	105	101	56	96	45	86	585	609	108

TABLE B-2-10

OVERALL ENERGY BALANCE ($\overline{M}$ Btu/hr)
LURGI/MOLTEN CARBONATE/ILLINOIS NO. 6

Inputs		
Coal Feed		10,195
Outputs		
Electric Power	3,390	
Cathode Vent		
Sensible Heat	263	
Water Vapor	739	
Ash	65	
Hydrocarbon Liquids	914	
Cooling Water	4,419	
Miscellaneous Losses	405	
TOTALS	10,195	10,195

TABLE B-2-11

OVERALL ENERGY BALANCE ($\overline{M}$ Btu/hr)
LURGI/MOLTEN CARBONATE/NEW MEXICO

Inputs		
Coal Feed		7,220
Outputs		
Electric Power	2,167	
Cathode Vent		
Sensible Heat	177	
Water Vapor	496	
Ash	128	
Hydrocarbon Liquids	915	
Cooling Water	2,932	
Miscellaneous Losses	405	
TOTALS	7,220	7,220

TABLE B-2-12

OVERALL ENERGY BALANCE ($\overline{M}$ Btu/hr)
TEXACO/MOLTEN CARBONATE/ILLINOIS NO. 6 COAL

Inputs			
Coal Feed			10,195
Outputs			
Electric Power		4,497	
Cathode Vent			
Sensible Heat	284		
Water Vapor	555	839	
Ash, Char		126	
Acid Gas Removal		108	
Cooling Water		4,310	
Miscellaneous Losses		315	
TOTALS		10,195	10,195

TABLE B-2-13

OVERALL ENERGY BALANCE ($\overline{M}$ Btu/hr)
TEXACO/MOLTEN CARBONATE/NEW MEXICO

Inputs			
Coal Feed			7,220
Outputs			
Electric Power		2,955	
Cathode Vent			
Sensible Heat	367		
Water Vapor	384	751	
Ash, Char		148	
Acid Gas Removal		63	
Cooling Water		3,081	
Miscellaneous Losses		222	
TOTALS		7,220	7,220

TABLE B-2-14

OVERALL ENERGY BALANCE ($\overline{M}$ Btu/hr)
SHELL/MOLTEN CARBONATE/VACUUM RESIDUE

Inputs		
Vacuum Residue		2,818
Naphtha		19
Outputs		
Electric Power	1,112	
Cathode Vent		
Sensible Heat	70	
Water Vapor	113	
Cooling Water	1,466	
Miscellaneous Losses	76	
TOTALS	2,837	2,837

Gasification Technology/Fuel	Heat Rate, Btu/kwh		
	Illinois No. 6	New Mexico	Vacuum Residue
Lurgi	10,011	11,238	--
Texaco	7,738	8,339	--
Shell	--	--	8,569

Plant utility summaries are presented in Tables B-2-15 through B-2-19. Utilities' quantities consumed and produced by each major process unit are provided.

2.2.1.1.3 Major System Components

Tables B-2-20 to B-2-24 present abbreviated summaries of the major equipment components of each plant section to provide insight into the magnitude of this equipment. These major components were the basis for subsequent capital cost estimates performed for each system. Other factors also entered into the cost analysis, including data on operating conditions and estimated number of units required.

2.2.1.1.4 System Design Aspects

While the requirements of the major system components could be readily determined from the hypothetical power plant designs set forth in the flowsheets, material and utility balances presented above, we encountered several situations in our discussions with equipment suppliers where equipment of the necessary description was not readily available in today's market for several reasons.

Generally, it was found that multiple units are required for virtually all applications. This can present an especially difficult problem in controlling gas flows through heat exchangers, gas expanders, etc., where it is important to minimize pressure losses. Even if the plant capacity were to have been limited to the 400 Mw range, multiple units would have been required in most applications.

The molten carbonate fuel cell designs included recycle compressors (gas boosters) which provide sufficient mass flow to remove fuel cell waste heat in the 1100-1300°F range. Arthur D. Little was unsuccessful

TABLE B-2-15

UTILITY SUMMARY
LURGI GASIFICATION/MOLTEN CARBONATE/ILLINOIS NO. 6

	Steam Balance, lb/hr								Motive Power Summary, kw		Electric Power Summary, kw		Cooling Water gpm	Raw Water gpm
	2400 psia		355 psia		115 psia		65 psia							
	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed		
Coal Preparation												13,100		
Oxidant Feed System										121,100				
Gasification and Ash Handling			344,800	1,188,600								26,600	316,585	2,400
Gas Cooling and Cleaning							129,200							
Acid Gas Removal			53,500											10
Process Condensate Treating						146,000		301,000						
Power Recovery									117,500		44,900			
Fuel Cell System										117,500	754,500*			
Fuel Cell Oxidant System										75,750				
Bottoming Cycle	3,526,100	3,526,100	3,526,100	2,735,800	2,409,000	2,263,000	2,263,000	2,091,200	196,850		272,350	10,200	272,350	70
Utilities and Offsites												28,370	(588,935)	20,200
TOTAL	3,526,100	3,526,100	3,924,400	3,924,400	2,409,000	2,409,000	2,392,200	2,392,200	314,350	314,350	1,071,750	78,270	0	22,680
NET	-	-	-	-	-	-	-	-	-	-	993,480	-	-	22,680

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*Net of inverter loss.

TABLE B-2-16

UTILITY SUMMARY
LURGI GASIFICATION/MOLTEN CARBONATE/NEW MEXICO COAL

	Steam Balance, lb/hr								Motive Power Summary, kw		Electric Power Summary, kw		Cooling Water gpm	Raw Water gpm
	2400 psia		565 psia		115 psia		75 psia							
	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed		
Coal Preparation												3,100		
Oxidant Feed System										76,650				
Gasification and Ash Handling			118,145	517,980								19,200	200,510	1,040
Gas Cooling and Cleaning														
Acid Gas Removal						42,720								10
Process Condensate Treating								113,760						
Power Recovery									70,250		31,350			
Fuel Cell System										70,250	465,460*			
Fuel Cell Oxidant System										53,000				
Bottoming Cycle	2,478,810	2,478,810	2,478,810	2,078,975	1,730,045	1,687,325	1,687,325	1,573,565	129,650		186,450	7,250	190,420	50
Utilities and Offsites												18,830	(390,930)	13,410
TOTAL	2,478,810	2,478,810	2,596,955	2,596,955	1,730,045	1,730,045	1,687,325	1,687,325	199,900	199,900	683,260	48,380	0	14,510
NET	-	-	-	-	-	-	-	-	-	-	634,880	-	-	14,510

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*Net of inverter loss.

TABLE B-2-17

UTILITY SUMMARY
TEXACO GASIFICATION/MOLTEN CARBONATE/ILLINOIS NO. 6

	Steam Balance, lb/hr						Motive Power Summary, kw		Electric Power Summary, kw		Cooling Water gpm	Raw Water gpm
	2400 psia		485 psia		90 psia							
	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed		
Coal Preparation										8,000		
Oxidant Feed								178,700			81,300	
Gasification and Ash Handling								13,700				
Gas Cooling											16,000	
Acid Gas Removal						120,300		4,100		5,900	10,700	
Power Recovery							287,300					
Fuel Cell System								82,300	925,400*			
Fuel Cell Oxidant System								174,600			19,200	
Bottoming Cycle	3,478,300	3,478,300	3,358,300	3,358,300	120,300		174,900	8,000	435,800		440,000	335
Utilities and Offsites										29,700	(567,200)	19,440
TOTAL	3,478,300	3,478,300	3,358,300	3,358,300	120,300	120,300	462,200	462,200	1,361,200	43,600	0	19,775
NET	-	-	-	-	-	-	-	-	1,317,600	-	-	19,775

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*Net of inverter loss.

TABLE B-2-18

UTILITY SUMMARY
TEXACO GASIFICATION/NEW MEXICO COAL/MOLTEN CARBONATE FUEL CELL

	Steam Balance, lb/hr						Motive Power Summary, kw		Electric Power Summary, kw		Cooling Water gpm	Raw Water gpm
	2400 psia		485 psia		90 psia							
	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed		
Coal Preparation										7,000		
Oxidant Feed								142,400			64,800	
Gasification and Ash Handling								12,200				
Gas Cooling											14,300	
Acid Gas Removal						70,000		2,400		4,300	7,800	
Power Recovery							205,700					
Fuel Cell System								55,400	630,300*			
Fuel Cell Oxidant System								158,000			13,000	
Bottoming Cycle	2,501,800	2,501,800	2,431,800	2,431,800	70,000		171,700	7,000	268,800		318,700	250
Utilities and Offsites										22,000	(418,600)	14,350
TOTAL	2,501,800	2,501,800	2,431,800	2,431,800	70,000	70,000	377,400	377,400	899.100	33,300	0	14,600
NET	-	-	-	-	-	-	-	-	865,800	-	-	14,600

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*Net of inverter loss.

TABLE B-2-19

UTILITY SUMMARY
SHELL GASIFICATION/MOLTEN CARBONATE/VACUUM RESIDUE

	2400 psia		1275 psia		485 psia		65 psia		Motive Power Summary, kw		Electric Power Summary, kw		Cooling Water gpm	Raw Water gpm
	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed		
Oxidant Feed										55,160				
Gasification and Ash Handling			397,200	397,200			4,430	4,430	55,160			3,500	114,495	410
Gas Cooling and Cleaning														
Acid Gas Removal														
Power Recovery									43,800		8,700			
Fuel Cell System										43,800	234,500*			
Fuel Cell Oxidant System										17,600 17,600				
Bottoming Cycle	730,000	730,000			730,000	730,000	634,800	634,800	17,600		97,700	2,100	80,975	20
Utilities and Offsites												9,415	(195,470)	6,710
TOTAL	730,000	730,000	397,200	397,200	730,000	730,000	639,230	639,230	116,560	116,560	340,900	15,015	0	7,140
NET	-	-	-	-	-	-	-	-	-	-	325,885	-	-	7,140

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*Net of inverter loss.

TABLE B-2-20

MAJOR SYSTEM COMPONENTS - HEAT EXCHANGERS
LURGI GASIFICATION/MOLTEN CARBONATE/ILLINOIS NO. 6

Exchanger Number	Plant Section	Q MBtu/hr	U Btu/hr-Ft^2-°F	A Ft^2
E-1	Power Recovery	258	20	51,100
E-2	Power Recovery	320	20	66,300
E-3	Power Recovery	335	20	74,100
E-4	Fuel Cell Oxidant System	442	10	199,000
E-5	Power Recovery	158	10	64,100
E-6	Bottoming Cycle	1595	15	230,400
E-7	Bottoming Cycle	480	10	96,400
E-8	Bottoming Cycle	236	10	277,700
E-9	Bottoming Cycle	322	10	830,800
E-10	Bottoming Cycle	2227	290	640,000

Continued

TABLE B-2-20 (Concluded)

MAJOR SYSTEM COMPONENTS - MISCELLANEOUS EQUIPMENT

Plant Section	Component Description	Size or Capacity	Comments
Oxidant Feed	Air Compressor	162,395 HP	EPRI RP-239 (Driven by T-2)
Gasification	Coal Handling, Preparation	10,000 TPD	EPRI RP-239 (Driven by T-2)
	Lurgi Gasification	10,000 TPD	"
	Gas Cooling	10,000 TPD	"
Acid Gas Removal	Selexol	26,500 lb H_2S/hr	"
	Sulfur Reduction	26,500 lb H_2S/hr	"
Process Condensate Treating	Hydrocarbon Liquid Recovery	58,000 lb/hr	"
Power Recovery	Gas Expander (T-1)	162,400 kw	1500°F Inlet
Fuel Cell System	Fuel Cell Stack	754,400 kw	1100°F Outlet (Driven by T-1)
	Recycle Compressor	157,570 HP	1100°F Outlet (Driven by T-1)
Fuel Cell Oxidant System	Air Compressor	101,580 HP	Driven by T-3
Bottoming Cycle	Waste Heat Boiler	3,526,100 lb/hr	
	Pumps	13,680 HP	Including drivers
	Turbines (T-2, T-3)	469,200 kw	2400 psia, 1000°F Inlet
	Generators	272,350 kw	
Utilities, Offsites	Cooling Water System	588,935 gpm	
	BFW Treatment	11,340,000 lb/hr	

TABLE B-2-21

MAJOR SYSTEM COMPONENTS - HEAT EXCHANGERS
LURGI GASIFICATION/MOLTEN CARBONATE/NEW MEXICO

Exchanger Number	Plant Section	Q $\overline{M}$Btu/hr	U Btu/hr-Ft^2-°F	A Ft^2
E-1	Power Recovery	173	20	32,500
E-2	Power Recovery	199	20	40,400
E-3	Power Recovery	209	20	44,700
E-4	Fuel Cell Oxidant System	309	10	134,500
E-5	Power Recovery	101	10	41,000
E-6	Bottoming Cycle	952	15	133,900
E-7	Bottoming Cycle	347	10	76,100
E-8	Bottoming Cycle	166	10	195,300
E-9	Bottoming Cycle	219	10	204,700
E-10	Bottoming Cycle	1862	290	535,000

Continued

TABLE B-2-21 (Concluded)

MAJOR SYSTEM COMPONENTS - MISCELLANEOUS EQUIPMENT

Plant Section	Component Description	Size or Capacity	Comments
Oxidant Feed	Air Compressor	102,920 HP	El Paso Natural Gas Company (Driven by T-3)
Gasification	Coal Handling, Preparation	10,000 TPD	"
	Lurgi Gasification	10,000 TPD	"
	Gas Cooling	10,000 TPD	"
Acid Gas Removal	Stretford	5,880 lb H_2S/hr	"
	Sulfur Reduction	5,880 lb H_2S/hr	"
Process Condensate Treating	Hydrocarbon Liquid Recovery	58,000 lb/hr	"
Power Recovery	Gas Expander (T-1)	101,600 kw	1500°F Inlet
Fuel Cell System	Fuel Cell Stack	465,460 kw	1100°F Outlet (Driven by T-1)
	Recycle Compressor	94,200 HP	1100°F Outlet (Driven by T-1)
Fuel Cell Oxidant System	Air Compressor	71,100 HP	(Driven by T-2)
Bottoming Cycle	Waste Heat Boiler	2,478,810 lb/hr	
	Pumps	9,720 HP	Including drivers
	Turbines (T-2, T-3)	316,100 kw	2400 psia, 1000°F Inlet
	Generators	186,450 kw	
Utilities, Offsites	Cooling Water System	390,930 gpm	
	BFW Treatment	7,255,000 lb/hr	

TABLE B-2-22

MAJOR SYSTEM COMPONENTS - HEAT EXCHANGERS
TEXACO/MOLTEN CARBONATE/ILLINOIS NO. 6 COAL

Exchanger Number	Plant Section	Q MBtu/hr	LMTD °F	U Btu/hr-Ft^2-°F	A Ft^2
E-1a	Power Recovery	228	86	25	106,000
E-1b	Power Recovery	372	231	30	53,700
E-2	Power Recovery	177	328	30	18,000
E-3a	Gasification	82	160	30	17,100
E-3b	Gasification	448	160	105	26,700
E-3c	Bottoming	794	67	105	112,900
E-3d	Power Recovery	279	160	65	26,800
E-3e	Bottoming	372	67	105	52,900
E-4	Gasification	120	51	170	13,800
E-5	Bottoming	2,499	532	20	234,900
E-6a	Fuel Cell Oxidant System	329	53	15	413,800
E-6b	Bottoming	716	184	30	129,700
E-7	Bottoming	376	143	20	131,500
E-8	Bottoming	3,300	12	290	948,300

Continued

TABLE B-2-22 (Concluded)

MAJOR SYSTEM COMPONENTS - MISCELLANEOUS EQUIPMENT

Plant Section	Component Description	Size or Capacity	Comments
Oxidant Feed	Oxygen Plant	8,540 TPD 99.5% O_2	A.D. Little Estimates
	Oxygen Compressor	68,000 HP	Driven by T-1, T-2, T-4a
Gasification	Coal Handling, Preparation	10,000 TPD	A.D. Little Estimates
	Texaco Gasification	10,000 TPD	"
	Char Recovery	10,000 TPD	"
	Gas Cooling	10,000 TPD	"
Acid Gas Removal	Rectisol	15,000 lb H_2S/hr	Selective H_2S
	Sulfur Reduction	15,000 lb H_2S/hr	A.D. Little Estimates
Power Recovery	Gas Expanders (T-1)	66,800 kw	1500°F Inlet
	(T-2)	66,800 kw	1500°F Inlet
	(T-3)	153,700 kw	1300°F Inlet
Fuel Cell System	Fuel Cell Stack	925,400 kw	EPRI
	Recycle Compressor	82,300 kw	1100°F Outlet; Driven by T-5a
Fuel Cell Oxidant	Compressor	234,000 HP	Driven by T-3, T-4b
Bottoming System	Waste Heat Boiler	4,940,000 lb steam/hr	
	Pumps	11,800 HP	Including drivers
	Turbines (T-4a, T-4b)	167,800 kw	2400 psia, 1000°F Inlet
	(T-5a, T-5b)	442,900 kw	485 psia, 1000°F Inlet
	Generators	435,800 kw	Driven by T-4c, T-5b
Utilities, Offsites	Cooling Water System	567,200 gpm	
	BFW Treatment	168,000 lb/hr	

TABLE B-2-23

MAJOR SYSTEM COMPONENTS - HEAT EXCHANGERS
TEXACO/MOLTEN CARBONATE/NEW MEXICO COAL

Exchanger Number	Plant Section	Q MBtu/hr	LMTD °F	U Btu/hr-Ft^2-°F	A Ft^2
E-1a	Power Recovery	170	134	25	50,700
E-1b		227	256	30	29,600
E-2		131	391	30	11,200
E-3a	Gasification	64	160	30	13,300
E-3b		325	160	105	19,300
E-3c	Bottoming	1,139	60	105	180,800
E-3d	Power Recovery	206	160	65	19,800
E-4	Gasification	107	51	170	12,300
E-5	Bottoming	1,662	532	20	156,200
E-6		424	193	40	54,900
E-7a	Fuel Cell Oxidant System	212	67	15	210,900
E-7b	Bottoming	175	173	30	33,700
E-8		2,390	12	290	686,800

Continued

TABLE B-2-23 (Concluded)

MAJOR SYSTEM COMPONENTS - MISCELLANEOUS EQUIPMENT

Plant Section	Component Description	Size or Capacity	Comments
Oxidant Feed	Oxygen Plant	6,810 TPD 99.5% O_2	A.D. Little Estimates
	Oxygen Compressor	54,000 HP	Driven by T-1, T-2, T-4a
Gasification	Coal Handling, Preparation	10,000 TPD	A.D. Little Estimates
	Texaco Gasification	10,000 TPD	"
	Char Recovery	10,000 TPD	"
	Gas Cooling	10,000 TPD	"
Acid Gas Removal	Rectisol	6,120 lb H_2S/hr	Selective H_2S
	Sulfur Reduction	6,120 lb H_2S/hr	A.D. Little Estimates
Power Recovery	Gas Expanders (T-1)	49,700 kw	1500°F Inlet
	(T-2)	49,700 kw	
	(T-3)	106,300 kw	
Fuel Cell System	Fuel Cell Stack	656,600 kw	EPRI
	Recycle Compressor	55,400 kw	1100°F Outlet; Driven by T-5a
Fuel Cell Oxidant	Compressor	158,000 HP	Driven by T-3, T-4b
Bottoming System	Waste Heat Boiler	3,130,000 lb/hr	
	Pumps	9,400 HP	Including drivers
	Turbines (T-4a, T-4b)	119,000 kw	2400 psia, 1000°F Inlet
	(T-5a, T-5b)	321,500 kw	
	Generators	268,800 kw	Driven by T-5b
Utilities, Offsites	Cooling Water System	418,600 gpm	
	BFW Treatment	125,000 lb/hr	

TABLE B-2-24

MAJOR SYSTEM COMPONENTS - HEAT EXCHANGERS
SHELL/MOLTEN CARBONATE/VACUUM RESIDUE

Exchanger Number	Plant Section	Q $\overline{\text{M}}$Btu/hr	U Btu/hr-Ft^2-°F	A Ft^2
E-1	Power Recovery	85	20	17,800
E-2	Power Recovery	102	20	22,100
E-3	Power Recovery	105	20	27,200
E-4	Fuel Cell Oxidant System	101	10	49,900
E-5	Power Recovery	56	10	22,700
E-6	Bottoming Cycle	96	10	21,500
E-7	Bottoming Cycle	45	10	62,500
E-8	Bottoming Cycle	86	10	114,700
E-9	Bottoming Cycle	588	20	82,710
E-10	Bottoming Cycle	609	290	175,000

Continued

TABLE B-2-24 (Concluded)

MAJOR SYSTEM COMPONENTS - MISCELLANEOUS EQUIPMENT

Plant Section	Component Description	Size or Capacity	Comments
Oxidant Feed	Air Compressor	73,970 HP	Shell Development Co. (Driven by T-3)
Gasification	Shell Gasification	10,000 bbl/d	"
	Gas Cooling	10,000 bbl/d	"
Acid Gas Removal	Sulfinol	7,850 lb H_2S/hr	"
	Sulfur Reduction	7,850 lb H_2S/hr	"
Power Recovery	Gas Expander (T-1)	52,500 kw	1500°F Inlet
Fuel Cell System	Fuel Cell Stack	234,500 kw	
	Recycle Compressor	58,740 HP	1100°F Outlet (Driven by T-1)
Fuel Cell Oxidant System	Air Compressor	23,600 HP	Driven T-2
Bottoming Cycle	B-1	730,000 lb/hr	
	Pumps	2,820 HP	Including drivers
	Turbines (T-2, T-3)	115,300 kw	
	Generators	97,700 kw	
Utilities, Offsites	Cooling Water System	195,470 gpm	
	BFW Treatment	3,570,000 lb/hr	

in finding large volume, high temperature machines immediately available to do this job. There is apparently only a limited market for such compressors and they are available presently only in relatively small sizes. The technology, however, is well within the range of the state-of-the-art combustion turbine equipment, but some development may be required to adapt for this application.

It was further found that gas expanders are generally unavailable in the configuration required for application above 1200°F. The power recovery turbines for the molten carbonate systems are required for the temperature range of 1130-1500°F for most efficient use of high level waste heat.

2.2.1.2 Phosphoric Acid Fuel Cell Integration

Table B-2-25 lists the considerations which dictate the flow configuration of the phosphoric acid system. First, carbon monoxide cannot be utilized in the phosphoric acid fuel cell. In fact, in high concentrations, greater than 4% at the inlet, it can act as an electrode poison. Therefore, nearly any gasifier effluent would have to be treated to convert any excess CO to useful H_2 by the water gas shift reaction

$$CO + H_2O \rightarrow CO_2 + H_2$$

Second, since the fuel cell operates at relatively low temperature (375°F) the waste heat produced from the system is of fairly low quality, in the form of low pressure steam. A major use for this steam is in the shift conversion section if it is also operated at low pressure. Since the fuel cell is downstream of the shift converters, it too must operate at a modest pressure no greater than about 50 psig.

Third, the anode vent contains combustibles, CO, H_2, and CH_4, as is true in the molten carbonate case. Since these combustibles represent a significant portion of the energy produced in the plant they must be burned to produce steam for use in a bottoming steam cycle for production of additional electric power.

TABLE B-2-25

SPECIAL CONSIDERATIONS (CONSTRAINTS) IN PHOSPHORIC ACID FUEL CELL INTEGRATION

Issue	Result
No CO Utilization	Shift Conversion
Low Temperature Waste Heat (LP Steam)	LP Shift Conversion; thus, LP Fuel Cell
Combustibles in Anode Vent	Burn Anode Vent (WHB)
Low Temperature F/C	Low Temperature Expander

Finally, since the fuel cell operates at low temperature any gas expansion for recovery of pressure energy from the gasifier effluent is probably best carried out at low temperature. While this lowers the recoverable energy by expansion, there are offsetting advantages:

- The incremental unrecovered energy remains available at a higher level to a steam bottoming cycle.
- The need for heat exchangers for gas reheat and subsequent cooling to the fuel cell temperature is minimized.
- The power recovery turbines are less expensive because of both the lower temperature (materials of construction) and the lower capacity (gas volume).

2.2.1.2.1 Process Description and Flowsheet

In the phosphoric acid fuel cell power plant (Figure B-2-9), fuel gas is produced in a high pressure, oxygen-blown Texaco gasifier and cleaned of char, ash and acid gas, as in the molten carbonate integration. The cleaned gas is reheated from ambient temperature to 350°F prior to entering each stage of a two-stage power recovery turbine. The philosophy was not to reheat to a very high temperature before power recovery as in the molten carbonate system because: (1) the fuel gas is not ultimately required at high temperature, and (2) the high level heat could instead be used to produce steam for a bottoming cycle for which the large quantity of low level acid cell waste heat would provide much of the boiler feedwater preheat.

After final adjustment of the fuel gas temperature to 325°F, it passes into a low pressure fuel cell for power generation. The anode vent still contains some combustible species including the unreacted hydrogen, carbon monoxide, and methane. These species are directed to a waste heat boiler where their fuel value is realized by producing high pressure steam for the bottoming steam cycle.

On the cathode side, air is delivered via an air compressor to the required system pressure. Preheat is provided to 325°F by cooling the cathode vent before its release to atmosphere.

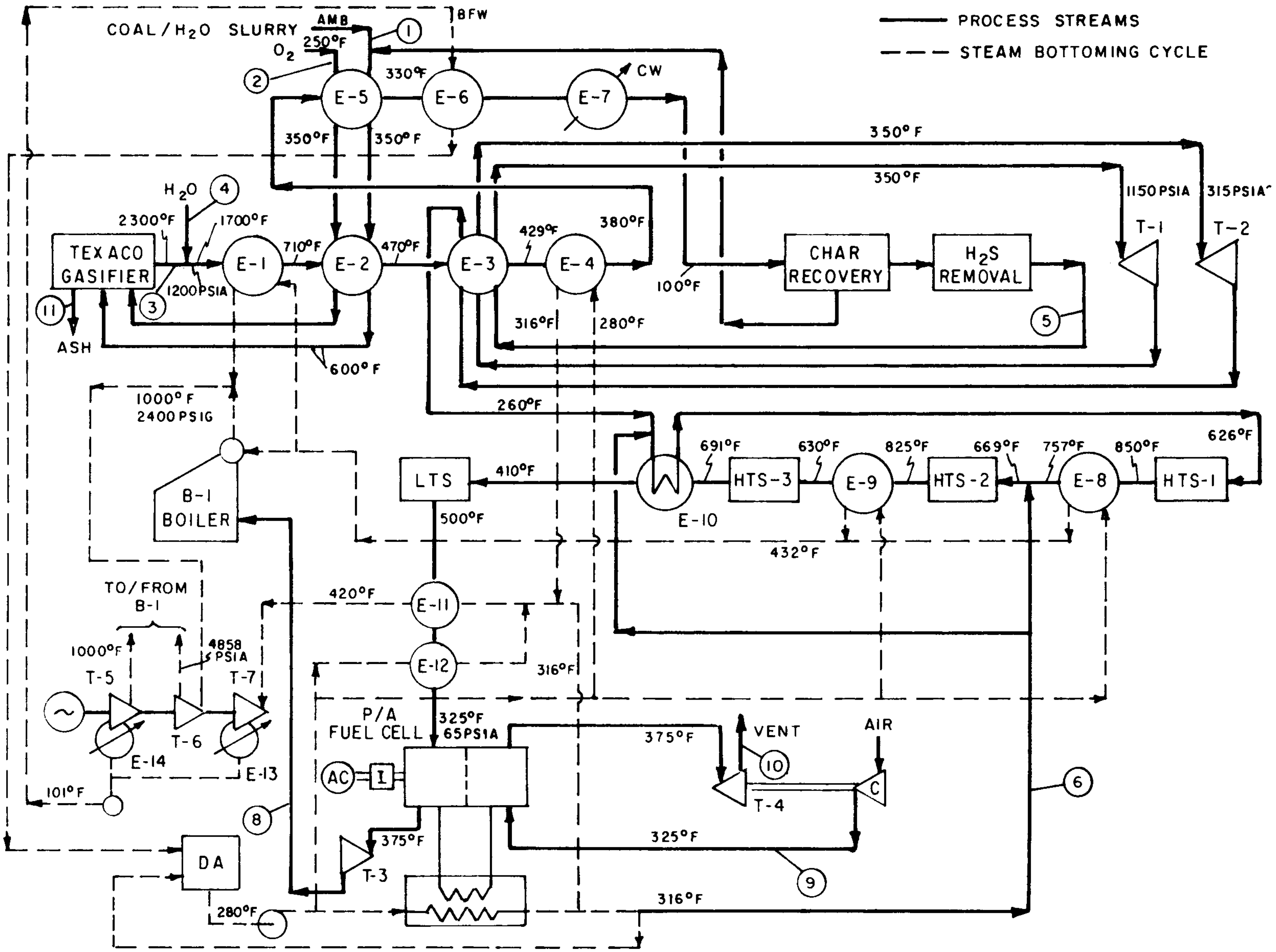

FIGURE B-2-9. BASELOAD FUEL CELL POWER PLANT. TEXACO GASIFIER WITH PHOSPHORIC ACID FUEL CELL.

2.2.1.2.2 Mass Balance and Utility Summary

The mass flow rates for the principal streams of the Texaco/phosphoric acid power plant are summarized in Table B-2-26 for Illinois coal. As before, the system size is based on 10,000 tons per day of coal fuel.

The overall energy balance for this power plant is shown in Table B-2-27. The resultant heat rate is 10,415 Btu/kwh. A summary of utilities consumed by the power plant is provided in Table B-2-28.

2.2.1.2.3 Major System Components

The major subsystems that comprise the power plant are listed in Table B-2-29. Also listed is the principal system parameter used in estimating investment costs. For several major subsystems, the source of cost information is indicated.

2.2.1.2.4 System Design Aspects

The discussion of section 2.2.1.1.4 indicating the general requirements for multiple units in most applications is true also of the phosphoric acid system. However, since a cathode recycle compressor is not required for the acid cell, the present commercial unavailability of such equipment is not a problem. Similarly, the high temperature gas expander was not a part of the acid system and presents no problem in this application. Therefore, it appears that the equipment technology required to commercialize the phosphoric acid fuel cell at an early date is generally available today.

2.3 System Costs

The general approach used in developing the capital cost estimates for each integrated system was to utilize available information for the gasification processes with appropriate modification to fit Arthur D. Little's specific designs. The remainder of the system was costed using capital cost estimates obtained from equipment suppliers. For the Lurgi gasification systems, with Illinois No. 6 coal and New Mexico coal, the

TABLE B-2-26

MATERIAL BALANCE
TEXACO/PHOSPHORIC ACID/ILLINOIS NO. 6

Stream No.	1 Coal/Water Slurry #/hr	2 Oxidant Feed #mols/hr	3 Gasifier Effluent #mols/hr	4 Quench Water #mols/hr	5 Cleaned Gas #mols/hr	6 Steam to Shift #mols/hr
Component Flows						
CH_4			569		569	
C_2^+						
CO			32,083		32,083	
CO_2			13,134		13,134	
H_2			28,253		28,253	
H_2O	681,667		28,765	18,440		37,993
N_2		111	468		468	
O_2		22,123				
H_2S			963			
MAF Coal, Oil	718,333					
Ash	80,000					
TOTAL	1,480,000	22,234	104,235	18,440	74,507	37,993

Stream No.	7 Fuel Cell Feed #mols/hr	8 Anode Vent #mols/hr	9 Fuel Cell Air #mols/hr	10 Cathode Vent #mols/hr	11 Ash #/hr
Component Flows					
CH_4	569	569			
C_2^+					
CO	4,500	4,500			
CO_2	40,718	40,718			
H_2	55,836	5,584			
H_2O	10,409	10,409		50,252	
N_2	468	468	190,189	190,189	
O_2			50,252	25,176	
H_2S					
MAF Coal, Oil					5,550
Ash					80,000
TOTAL	112,500	62,248	240,441	265,617	85,550

Heat Exchanger Duties, $\overline{M}$Btu/hr

E-1	E-2	E-3	E-4	E-5	E-6	E-7	E-8	E-9	E-10	E-11	E-12	E-13	E-14	B-1
1084	258	29	492	238	289	409	73	191	269	144	21	2210	1742	1250

TABLE B-2-27

OVERALL ENERGY BALANCE SUMMARY ($\overline{M}$ Btu/hr)
TEXACO/PHOSPHORIC ACID/ILLINOIS NO. 6

Inputs			
Coal Feed			10,195
Outputs			
Electric Power		3,371	
Cathode Vent			
Sensible Heat	599		
Water Vapor	950	1,549	
Boiler Flue			
Sensible Heat	144		
Water Vapor	324	468	
Ash, Char		126	
Acid Gas Removal		108	
Cooling Water		4,361	
Miscellaneous Losses		212	
TOTALS		10,195	10,195

TABLE B-2-28

UTILITY SUMMARY
TEXACO GASIFICATION/ILLINOIS NO. 6 COAL/PHOSPHORIC ACID FUEL CELL

	Steam Balance, lb/hr						Motive Power Summary, kw		Electric Power Summary, kw		Cooling Water	Raw Water
	2400 psia		485 psia		85 psia							
	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	Produced	Consumed	gpm	gpm
Coal Preparation										8,000		
Oxidant Feed								178,700			81,300	
Gasification and Ash Handling								13,700				
Gas Cooling											55,400	
Acid Gas Removal								4,100		5,900	10,700	
Power Recovery							186,400					
Shift Conversion						683,900						
Fuel Cell System									755,600*			
Fuel Cell Oxidant System								184,500			31,600	
Bottoming Cycle	1,740,600	1,740,600	1,620,300	1,620,300		2,525,700	199,800	5,200	274,300		527,500	2,310
Utilities and Offsites										37,100	(706,500)	24,200
TOTAL	1,740,600	1,740,600	1,620,300	1,620,300	3,329,900	3,329,900	386,200	386,200	1,029,900	51,000	0	26,510
NET												
NET	-	-	-	-	-	-	-	-	978,900	-	-	26,510

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*Net of inverter loss.

TABLE B-2-29

MAJOR SYSTEM COMPONENTS - HEAT EXCHANGER
TEXACO GASIFICATION/PHOSPHORIC ACID/ILLINOIS NO. 6

Exchanger Number	Plant Section	Q MBtu/hr	LMTD °F	U Btu/hr-Ft^2-°F	A Ft^2
E-1	Bottoming	1,084	322	35	96,200
E-2a	Gasification	224	115	25	77,900
E-2b	Gasification	34	115	30	9,900
E-3a	Power Recovery	139	207	35	19,200
E-3b	Power Recovery	84	184	35	13,000
E-3c	Shift Conversion	70	238	35	8,400
E-4	Bottoming	492	86	105	54,500
E-5a	Gasification	224	98	105	22,200
E-5b	Gasification	14	51	65	4,200
E-6	Bottoming	289	132	190	11,500
E-7	Gasification	409	85	170	28,300
E-8	Bottoming	73	447	25	6,500
E-9	Bottoming	191	371	25	20,600
E-10	Shift Conversion	269	96	15	186,800
E-11	Bottoming	144	52	20	138,500
E-12	Bottoming	21	18	25	46,700
E-13	Bottoming	2,210	12	290	635,100
E-14	Bottoming	1,742	12	290	500,600

Continued

TABLE B-2-29 (Concluded)

MAJOR SYSTEM COMPONENTS - MISCELLANEOUS EQUIPMENT

Plant Section	Component Description	Size or Capacity	Comments
Oxidant Feed	Oxygen Plant	8,540 TPD, 99.5%/O_2	Driven by T-1, T-2, T-3, T-6d
	Oxygen Compressor	68,000 HP	Driven by T-1, T-2, T-3, T-6d
Gasification	Coal Preparation	10,000 TPD	
	Gasifier	10,000 TPD	
	Char Recovery	10,000 TPD	
Acid Gas Removal	Rectisol	15,000 lb H_2S/hr	Selective to H_2S
	Sulfur Reduction	15,000 lb H_2S/hr	
Power Recovery	Expanders (T-1)	30,900 kw	350°F Inlet
	(T-2)	30,900 kw	350°F Inlet
	(T-3)	30,200 kw	375°F Inlet
	(T-4)	94,400 kw	375°F Inlet
Shift Conversion	High Temperature Shift	20,500 cu ft	
	Low Temperature Shift	23,400 cu ft	
Fuel Cell System	Fuel Cell Stack	87,080 kw	
Fuel Cell Oxidant System	Fuel Cell Air Compressor	184,500 kw	Driven by T-4, T-5
Bottoming System	B-1	1,250 $\overline{\text{M}}$Btu/hr (Net)	
	Pumps	7,000 HP	With drivers
	Turbines (T-5)	90,100 kw	2400 psia, 1000°F Inlet
	(T-6a, T-6b)	219,100 kw	485 psia, 1000°F Inlet
	(T-7)	164,900 kw	65 psia, 420°F Inlet
	Generators	274,300 kw	

EPRI RP-239 Report (1) and El Paso Natural Gas Company Report (7) were utilized, respectively, for costing all process units directly involved with the coal gasification plant section itself. All other process units related specifically to power recovery were costed based on various equipment manufacturers' budget estimates with the exception of the fuel cell stack itself whose cost was determined from EPRI RP-318 (2).

The capital cost estimates for the three systems involving the Texaco gasification system with both Illinois No. 6 and New Mexico coal were developed with the aid of EPRI RP-318 and Arthur D. Little, Inc. data. As with the Lurgi systems, all capital investment costs associated with power recovery process units were based on budget estimates supplied by equipment manufacturers. A summary of such estimates is included in Table B-2-30.

In determining the capital cost for the process units directly involved with the gasification of the vacuum residue using the Shell system, literature available from the Shell Development Company (6) was utilized. As for the power recovery process units, their capital cost estimates were obtained from equipment manufacturers.

As mentioned earlier the gasification technologies were chosen for study because they are state-of-the-art and it would be misleading to imply that fuel cell development must rely on concurrent development of suitable gasification technology. For similar reasons we generally chose available technologies and major equipment types rather than assume that major breakthroughs in these areas would occur on a timely schedule. In some cases, modest advances where they are likely were assumed.

As the flowsheets indicate, the power recovery turbines used for the molten carbonate system would operate at about 1500°F. While these limits are not commercially available today for expander application, the technology is well within the capability of present day combustion turbine know-how.

Another application which would require development of improved or advanced technology is the high temperature heat exchangers required for sensible energy recovery in the Texaco gasifier effluent. Being the product of an entrained gasifier, this gas will be at high temperature (quenched to 1700°F) and heavily loaded with particulate.

TABLE B-2-30

SUMMARY OF BUDGET ESTIMATES OF MAJOR SYSTEM COMPONENTS

Equipment	Characteristics	Size	Cost, $	Source
Heat Exchangers	Surface Condensers	260,000 sq ft	2,000,000	Graham
	Shell and Tube	150,000 sq ft	2,250,000	A.D. Little Estimates
	Shell and Fin Tube	100,000 sq ft	130,000	Perfex
Steam Turbines (with generators)	Condensing	478,000 kw	20,100,000	Westinghouse
	Non-condensing	176,300 kw	11,200,000	Westinghouse
Gas Expander	High Temperature	33,900 HP	2,000,000	Turbodyne
Fuel Cell (with inverter)	Molten Carbonate	1.0 kw	140	EPRI
	Phosphoric Acid	1.0 kw	135	EPRI
Boiler	Fired	2,600,000 lb/hr	30,000,000	Babcock & Wilcox
Compressor	Air	22,400 HP	1,350,000	Elliott
Boiler	Unfired	96,200 sq ft	1,924,000	A.D. Little Estimates
Pumps	BFW	4,000 HP	240,000	Ingersoll-Rand
	Cooling Water	1,250 HP	100,000	Ingersoll-Rand
Compressor	Fuel Cell Recycle	1,000 HP	275,000	Spencer
Acid Gas Removal	Rectisol	28,200,000 Scfh (Gas in)	22,664,000	Lotépro

2.3.1 Baseload System Capital Investment Costs

Tables B-2-31 to B-2-36 present the capital cost summaries in 1975 dollars for each of the six integrated baseload systems, with a breakdown of each system into its major process units and their respective, individual installed costs. Total fixed capital cost was obtained from the total direct plant cost by the addition of a contingency factor equivalent to 15% of the total direct plant cost. This contingency factor is included to cover anticipated costs for providing: (1) spare or redundant capacity where necessary for overall plant reliability, and (2) additional equipment required for integrated process operation which was inevitably omitted from consideration in a preliminary scoping design of this type. This contingency factor does not include any component for business risk or error in cost estimation. Any such components must be in addition. Other costs are estimated as specified percentages of both the total fixed capital and total depreciable capital and added to the fixed capital cost to obtain the total capital investment.

These total capital investments per kw of net output capacity including IDC, start-up costs and working capital are:

Lurgi - M/C - Ill.	$959/kw
Lurgi - M/C - N.M.	$960/kw
Texaco - M/C - Ill.	$762/kw
Texaco - M/C - N.M.	$856/kw
Shell - M/C - Vac. Res.	$851/kw
Texaco - Acid - Ill.	$834/kw

For a given coal, the Texaco-M/C is the lowest cost system and the Lurgi-M/C system the most expensive. The Texaco-Acid system is more expensive than the Texaco-M/C based on gasification of Illinois coal. The unit costs are higher for the New Mexico coal cases because the output capacity is less for a constant coal input of 10,000 tons/day.

2.3.2 Baseload System Annual Operating Costs

Tables B-2-37 to B-2-43 present the operating cost summaries for all six integrated systems as a function of geographic location for the

TABLE B-2-31

CAPITAL COST SUMMARY

LURGI GASIFICATION WITH MOLTEN CARBONATE FUEL CELL USING ILLINOIS NO. 6 COAL

Capacity - 993,480 kw

Basis Year - 1975

CATEGORY	$\$10^3$ (mid-1975)
INSTALLED PROCESS UNITS	
Oxidant Feed System	21,020
Gasification	109,630
Acid Gas Removal	40,740
Process Condensate Treating	43,190
Power Recovery	43,620
Fuel Cell System	143,900
Fuel Cell Oxidant System	29,010
Bottoming Cycle	98,050
Utilities and Offsites	103,860
Total Direct Plant	633,020
Contingency @ 15%	94,950
Total Fixed Capital	727,970
Interest During Construction @ 17%[a]	123,750
Owner's Cost @ 3%	21,840
Total Depreciable Capital	873,560
Start up Cost[b]	35,610
Working Capital[c]	43,680
TOTAL CAPITAL INVESTMENT	952,850
Dollars/kw	959

---

[a] 5-year construction period, 8% interest.

[b] 20% of Annual Gross Operating Cost.

[c] 5% of Depreciable Capital.

TABLE B-2-32

CAPITAL COST SUMMARY

LURGI GASIFICATION WITH MOLTEN CARBONATE FUEL CELL USING NEW MEXICO COAL

Capacity - 634,880 kw

Basis Year - 1975

CATEGORY	$\$10^3$ (mid-1975)
INSTALLED PROCESS UNITS	
Oxidant Feed System	12,450
Gasification	69,210
Acid Gas Removal	9,390
Process Condensate Treating	34,550
Power Recovery	29,840
Fuel Cell System	87,100
Fuel Cell Oxidant System	21,350
Bottoming Cycle	70,410
Utilities and Offsites	69,480
Total Direct Plant	403,780
Contingency @ 15%	60,570
Total Fixed Capital	464,350
Interest During Construction @ 17%[a]	78,940
Owner's Cost @ 3%	13,930
Total Depreciable Capital	557,220
Start up Cost[b]	24,600
Working Capital[c]	27,860
TOTAL CAPITAL INVESTMENT	609,680
Dollars/kw	960

---

[a] 5-year construction period, 8% interest.

[b] 20% of Annual Gross Operating Cost.

[c] 5% of Depreciable Capital.

TABLE B-2-33

CAPITAL COST SUMMARY

TEXACO GASIFICATION WITH MOLTEN CARBONATE FUEL CELL USING ILLINOIS NO. 6 COAL

Capacity - 1,317,600 kw

Basis Year - 1975

CATEGORY	$\$10^3$ (mid-1975)
INSTALLED PROCESS UNITS	
Oxidant Feed System	91,670
Gasification	87,480
Acid Gas Removal	25,240
Power Recovery	71,080
Fuel Cell System	157,020
Fuel Cell Oxidant System	45,830
Bottoming Cycle	101,980
Utilities and Offsites	87,090
Total Direct Plant	667,390
Contingency @ 15%	100,110
Total Fixed Capital	767,500
Interest During Construction @ 17%[a]	130,480
Owner's Cost @ 3%	23,020
Total Depreciable Capital	921,000
Start up Cost[b]	36,540
Working Capital[c]	46,050
TOTAL CAPITAL INVESTMENT	1,003,590
Dollars/kw	762

---

[a] 5-year construction period, 8% interest.

[b] 20% of Annual Gross Operating Cost.

[c] 5% of Depreciable Capital.

TABLE B-2-34

CAPITAL COST SUMMARY

TEXACO GASIFICATION WITH MOLTEN CARBONATE FUEL CELL USING NEW MEXICO COAL

Capacity - 865,800 kw

Basis Year - 1975

CATEGORY	$\$10^3$ (mid-1975)
INSTALLED PROCESS UNITS	
Oxidant Feed System	74,710
Gasification	71,730
Acid Gas Removal	18,310
Power Recovery	49,170
Fuel Cell System	107,170
Fuel Cell Oxidant System	32,200
Bottoming Cycle	75,210
Utilities and Offsites	64,640
Total Direct Plant	493,140
Contingency @ 15%	73,970
Total Fixed Capital	567,110
Interest During Construction @ 17%[a]	96,410
Owner's Cost @ 3%	17,010
Total Depreciable Capital	680,530
Start up Cost[b]	26,730
Working Capital[c]	34,030
TOTAL CAPITAL INVESTMENT	741,290
Dollars/kw	856

---

[a] 5-year construction period, 8% interest.

[b] 20% of Annual Gross Operating Cost.

[c] 5% of Depreciable Capital.

TABLE B-2-35

CAPITAL COST SUMMARY

SHELL GASIFICATION WITH MOLTEN CARBONATE FUEL CELL USING VACUUM RESIDUE

Capacity - 325,885 kw

Basis Year - 1975

CATEGORY	$\$10^3$ (mid-1975)
INSTALLED PROCESS UNITS	
Oxidant Feed System	10,510
Gasification	20,750
Acid Gas Removal	6,240
Power Recovery	16,320
Fuel Cell System	49,600
Fuel Cell Oxidant System	8,000
Bottoming Cycle	35,100
Utilities and Offsites	35,040
Total Direct Plant	181,560
Contingency @ 15%	27,230
Total Fixed Capital	208,790
Interest During Construction @ 17%[a]	35,500
Owner's Cost @ 3%	6,260
Total Depreciable Capital	250,550
Start up Cost[b]	14,200
Working Capital[c]	12,530
TOTAL CAPITAL INVESTMENT	277,280
Dollars/kw	851

---

[a] 5-year construction period, 8% interest.

[b] 20% of Annual Gross Operating Cost.

[c] 5% of Depreciable Capital.

TABLE B-2-36

CAPITAL COST SUMMARY

TEXACO GASIFICATION WITH PHOSPHORIC ACID FUEL CELL USING ILLINOIS NO. 6 COAL

Capacity - 978,900 kw

Basis Year - 1975

CATEGORY	$\$10^3$ (mid-1975)
INSTALLED PROCESS UNITS	
Oxidant Feed System	88,120
Gasification	88,320
Acid Gas Removal	25,200
Power Recovery	21,850
Shift Conversion	16,210
Fuel Cell System	106,260
Fuel Cell Oxidant System	28,570
Bottoming Cycle	97,330
Utilities and Offsites	82,950
Total Direct Plant	554,810
Contingency @ 15%	83,220
Total Fixed Capital	638,030
Interest During Construction @ 17%[a]	108,470
Owner's Cost @ 3%	19,140
Total Depreciable Capital	765,640
Start up Cost[b]	34,390
Working Capital[c]	38,280
TOTAL CAPITAL INVESTMENT	838,310
Dollars/kw	834

---

[a] 5-year construction period, 8% interest.

[b] 20% of Annual Gross Operating Cost.

[c] 5% of Depreciable Capital.

TABLE B-2-37

TECHNOLOGY TEXACO/MC/ILLINOIS COAL
LOCATION COLUMBUS
YEAR 1985
LOAD FACTOR 7884. HRS/YEAR
FIXED PLANT INVESTMENT $ 767500000.
TOTAL PLANT INVESTMENT $ 1003590000.
NET POWER RATE 1314600. KW

CATEGORY	UNITS/YR	$/UNIT	$/YR(000)	COMP INDEX
VARIABLE COSTS				
COAL/OIL MBTU	80378960.	1.02	81986.	7.9
PURCHASED WATER KGAL	9359995.	0.40	3744.	0.4
CATALYSTS & CHEMICALS			15390.	1.5
ASH DISPOSAL TONS	336600.	1.00	337.	0.0
TOTAL VARIABLE			101457.	9.8
SEMI-VARIABLE COSTS				
OPERATING LABOR HRS	525600.	8.50	4468.	0.4
MAINTENANCE LABOR HRS	1860604.	9.90	18420.	1.8
SUPERVISION LABOR YRS	188.	24300.00	4578.	0.4
TOTAL LABOR			27465.	2.6
G & A			16479.	1.6
MAINTENANCE MATERIAL			12280.	1.2
TOTAL SEMI-VARIABLE COST			56224.	5.4
FIXED COST				
TAXES & INSURANCE			15350.	1.5
TOTAL GROSS OPERATING COST			173031.	16.7
BYPRODUCT CREDITS				
SULFUR L.T.	52920.	45.00	2381.	0.2
TOTAL BYPRODUCT CREDIT			2381.	0.2
TOTAL NET OPERATING COST			170650.	16.5
CAPITAL CHARGES			180646.	17.4
TOTAL ANNUAL COST			351296.	33.9

TABLE B-2-38

TECHNOLOGY TEXACO/PA/ILLINOIS COAL
LOCATION COLUMBUS
YEAR 1985
LOAD FACTOR 7884. HRS/YEAR
FIXED PLANT INVESTMENT $ 638030000.
TOTAL PLANT INVESTMENT $ 838310000.
NET POWER RATE 978900. KW

CATEGORY	UNITS/YR	$/UNIT	$/YR(000)	COMP INDEX
VARIABLE COSTS				
COAL/OIL MBTU	80378960.	1.02	81986.	10.6
PURCHASED WATER KGAL	12509994.	0.40	5004.	0.6
CATALYSTS & CHEMICALS			14049.	1.8
ASH DISPOSAL TONS	336600.	1.00	337.	0.0
TOTAL VARIABLE			101376.	13.1
SEMI-VARIABLE COSTS				
OPERATING LABOR HRS	525600.	8.50	4468.	0.6
MAINTENANCE LABOR HRS	1546739.	9.90	15313.	2.0
SUPERVISION LABOR YRS	163.	24300.00	3956.	0.5
TOTAL LABOR			23736.	3.1
G & A			14242.	1.8
MAINTENANCE MATERIAL			10208.	1.3
TOTAL SEMI-VARIABLE COST			48187.	6.2
FIXED COST				
TAXES & INSURANCE			12761.	1.7
TOTAL GROSS OPERATING COST			162323.	21.0
BYPRODUCT CREDITS				
SULFUR L.T.	52920.	45.00	2381.	0.3
TOTAL BYPRODUCT CREDIT			2381.	0.3
TOTAL NET OPERATING COST			159942.	20.7
CAPITAL CHARGES			150896.	19.6
TOTAL ANNUAL COST			310837.	40.3

TABLE B-2-39

TECHNOLOGY LURGI/MC/ILLINOIS COAL
LOCATION COLUMBUS
YEAR 1985
LOAD FACTOR 7884. HRS/YEAR
FIXED PLANT INVESTMENT $ 727970000.
TOTAL PLANT INVESTMENT $ 952850000.
NET POWER RATE 993480. KW

CATEGORY	UNITS/YR	$/UNIT	$/YR(000)	COMP INDEX
VARIABLE COSTS				
COAL/OIL MBTU	80387960.	1.02	81996.	10.5
PURCHASED WATER KGAL	10187995.	0.40	4075.	0.5
CATALYSTS & CHEMICALS			12797.	1.6
ASH DISPOSAL TONS	328500.	1.00	328.	0.0
TOTAL VARIABLE			99196.	12.7
SEMI-VARIABLE COSTS				
OPERATING LABOR HRS	578160.	8.50	4914.	0.6
MAINTENANCE LABOR HRS	1764774.	9.90	17471.	2.2
SUPERVISION LABOR YRS	184.	24300.00	4477.	0.6
TOTAL LABOR			26863.	3.4
G & A			16118.	2.1
MAINTENANCE MATERIAL			11648.	1.5
TOTAL SEMI-VARIABLE COST			54628.	7.0
FIXED COST				
TAXES & INSURANCE			14559.	1.9
TOTAL GROSS OPERATING COST			168384.	21.5
BYPRODUCT CREDITS				
SULFUR L.T.	87696.	45.00	3946.	0.5
AMMONIA TONS	40405.	100.00	4041.	0.5
HC LIQUIDS MBTU	8036996.	1.02	8198.	1.0
TOTAL BYPRODUCT CREDIT			16185.	2.1
TOTAL NET OPERATING COST			152199.	19.4
CAPITAL CHARGES			171513.	21.9
TOTAL ANNUAL COST				41.3

TABLE B-2-40

TECHNOLOGY SHELL/MC/VACUUM RESIDUE
LOCATION COLUMBUS
YEAR 1985
LOAD FACTOR 7884. HRS/YEAR
FIXED PLANT INVESTMENT $ 208790000.
TOTAL PLANT INVESTMENT $ 277280000.
NET POWER RATE 325885. KW

CATEGORY	UNITS/YR	$/UNIT	$/YR(000)	COMP INDEX
VARIABLE COSTS				
COAL/OIL MBTU	22211989.	2.42	53753.	20.9
PURCHASED WATER KGAL	3374998.	0.40	1350.	0.5
CATALYSTS & CHEMICALS			33.	0.0
ASH DISPOSAL TONS	0.	1.00	0.	0.0
TOTAL VARIABLE			55136.	21.5
SEMI-VARIABLE COSTS				
OPERATING LABOR HRS	166440.	8.50	1415.	0.6
MAINTENANCE LABOR HRS	506157.	9.90	5011.	2.0
SUPERVISION LABOR YRS	53.	24300.00	1285.	0.5
TOTAL LABOR			7711.	3.0
G & A			4626.	1.8
MAINTENANCE MATERIAL			3341.	1.3
TOTAL SEMI-VARIABLE COST			15678.	6.1
FIXED COST				
TAXES & INSURANCE			4176.	1.6
TOTAL GROSS OPERATING COST			74990.	29.2
BYPRODUCT CREDITS				
SULFUR L.T.	27625.	45.00	1243.	0.5
TOTAL BYPRODUCT CREDIT			1243.	0.5
TOTAL NET OPERATING COST			73747.	28.7
CAPITAL CHARGES			49910.	19.4
TOTAL ANNUAL COST			123657.	48.1

TABLE-B-2-41

TECHNOLOGY TEXACO//MC/NEWMEXICO COAL
LOCATION LOS ANGELES
YEAR 1985
LOAD FACTOR 7884. HRS/YEAR
FIXED PLANT INVESTMENT $ 567110000.
TOTAL PLANT INVESTMENT $ 741290000.
NET POWER RATE 865800. KW

CATEGORY	UNITS/YR	$/UNIT	$/YR(000)	COMP INDEX
VARIABLE COSTS				
COAL/OIL MBTU	56897972.	1.17	66571.	9.8
PURCHASED WATER KGAL	6929997.	0.40	2772.	0.4
CATALYSTS & CHEMICALS			10389.	1.5
ASH DISPOSAL TONS	647999.	1.00	648.	0.1
TOTAL VARIABLE			80379.	11.8
SEMI-VARIABLE COSTS				
OPERATING LABOR HRS	516840.	8.50	4393.	0.6
MAINTENANCE LABOR HRS	1374811.	9.90	13611.	2.0
SUPERVISION LABOR YRS	148.	24300.00	3601.	0.5
TOTAL LABOR			21605.	3.2
G & A			12963.	1.9
MAINTENANCE MATERIAL			9074.	1.3
TOTAL SEMI-VARIABLE COST			43641.	6.4
FIXED COST				
TAXES & INSURANCE			11342.	1.7
TOTAL GROSS OPERATING COST			135362.	19.8
BYPRODUCT CREDITS				
SULFUR L.T.	9900.	45.00	445.	0.1
TOTAL BYPRODUCT CREDIT			445.	0.1
TOTAL NET OPERATING COST			134917.	19.8
CAPITAL CHARGES			133432.	19.5
TOTAL ANNUAL COST			268349.	39.3

TABLE B-2-42

TECHNOLOGY LURGI/MC/NEWMEXICO COAL
LOCATION LOS ANGELES
YEAR 1985
LOAD FACTOR 7884. HRS/YEAR
FIXED PLANT INVESTMENT $ 464350000.
TOTAL PLANT INVESTMENT $ 609680000.
NET POWER RATE 634880. KW

CATEGORY	UNITS/YR	$/UNIT	$/YR(000)	COMP INDEX
VARIABLE COSTS				
COAL/OIL MBTU	56924972.	1.17	66602.	13.3
PURCHASED WATER KGAL	6857997.	0.40	2743.	0.5
CATALYSTS & CHEMICALS			8005.	1.6
ASH DISPOSAL TONS	665730.	1.00	666.	0.1
TOTAL VARIABLE			78016.	15.6
SEMI-VARIABLE COSTS				
OPERATING LABOR HRS	525600.	8.50	4468.	0.9
MAINTENANCE LABOR HRS	1125696.	9.90	11144.	2.2
SUPERVISION LABOR YRS	128.	24300.00	3122.	0.6
TOTAL LABOR			18734.	3.7
G & A			11241.	2.2
MAINTENANCE MATERIAL			7430.	1.5
TOTAL SEMI-VARIABLE COST			37405.	7.5
FIXED COST				
TAXES & INSURANCE			9287.	1.9
TOTAL GROSS OPERATING COST			124708.	24.9
BYPRODUCT CREDITS				
SULFUR L.T.	20695.	45.00	931.	0.2
AMMONIA TONS	29893.	100.00	2989.	0.6
HC LIQUIDS MBTU	8090996.	1.17	9466.	1.9
TOTAL BYPRODUCT CREDIT			13387.	2.7
TOTAL NET OPERATING COST			111321.	22.2
CAPITAL CHARGES			109742.	21.9
TOTAL ANNUAL COST				44.1

TABLE B-2-43

TECHNOLOGY SHELL/MC/VACUUM RESIDUE
LOCATION LOS ANGELES
YEAR 1985
LOAD FACTOR 7884. HRS/YEAR
FIXED PLANT INVESTMENT $ 208790000.
TOTAL PLANT INVESTMENT $ 277280000.
NET POWER RATE 325885. KW

CATEGORY	UNITS/YR	$/UNIT	$/YR(000)	COMP INDEX
VARIABLE COSTS				
COAL/OIL MBTU	22211989.	1.97	43758.	17.0
PURCHASED WATER KGAL	3374998.	0.40	1350.	0.5
CATALYSTS & CHEMICALS			33.	0.0
ASH DISPOSAL TONS	0.	1.00	0.	0.0
TOTAL VARIABLE			45141.	17.6
SEMI-VARIABLE COSTS				
OPERATING LABOR HRS	166440.	8.50	1415.	0.6
MAINTENANCE LABOR HRS	506157.	9.90	5011.	2.0
SUPERVISION LABOR YRS	53.	24300.00	1285.	0.5
TOTAL LABOR			7711.	3.0
G & A			4626.	1.8
MAINTENANCE MATERIAL			3341.	1.3
TOTAL SEMI-VARIABLE COST			15678.	6.1
FIXED COST				
TAXES & INSURANCE			4176.	1.6
TOTAL GROSS OPERATING COST			64995.	25.3
BYPRODUCT CREDITS				
SULFUR L.T.	27625.	45.00	1243.	0.5
TOTAL BYPRODUCT CREDIT			1243.	0.5
TOTAL NET OPERATING COST			63752.	24.8
CAPITAL CHARGES			49910.	19.4
TOTAL ANNUAL COST			113662.	44.2

1985 forecast year. The main variable costs affected by these three variables are the annual cost of coal and purchased water. Semi-variable costs such as operating labor, maintenance, and supervision are only affected by the baseload system operating factor variable. (Note: Tables 2-52 through 2-58 in Volume I of this report reflect in summary form the effects of various operating load factors for each forecast year, 1980, 1985, and 1990.)

Similarly, byproduct credits for such items as sulfur in the three Texaco gasification systems, and sulfur, hydrocarbon liquids and ammonia in the two Lurgi gasification systems are included and are affected by the baseload system operating factor.

Capital charges equivalent to 18% of the total investment are added to the total net operating cost to obtain the total annual operating cost for the system.