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ECONOMICS OF HYDROGEN PRODUCTION AND LIQUEFACTION UPDATED TO 1980

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FORWARD

All computations performed under this contract were in U.S. Customary engineering units.

All results are presented in the International System of Units (SI) followed, in parentheses, by the U.S. Customary equivalent from which they were converted.

All economic values are given in terms of mid-1980 dollars.

The work presented in this report was performed during the period February - May 1979 by Charles Baker, Engineering Associate, of the Industrial Gas Process Division, Linde Division, Union Carbide Corporation, Tonawanda, New York, 14150. Mr. Robert D. Witcofski of the Aeronautical Systems Division of NASA Langley Research Center was technical monitor for the contract.

1.0 SUMMARY

This work provides revised costs for generating and liquefying hydrogen in mid-1980 and represents an economic update of data from mid-1974 which was presented in three previously-issued NASA Contractor Reports.

Selected tables from these reports have been revised and are presented so that they correspond to the equivalent tables in the original reports. Plant investments have been treated as straightforward escalations resulting from inflation. Operating costs, however, have been derived in terms of the unit cost of coal, fuel gas and electrical energy to permit the determination of the influence of these parameters on the cost of liquid hydrogen.

In addition, inflationary influence has been recognized by requiring a 15% discounted rate of return on investment for Discounted Cash Flow financing analysis, up from 12% previously. Utility financing has been revised to require an 11% interest rate on debt, compared with 9% before.

The scope of operation of the hydrogen plant has also been revised from previous studies to include only the hydrogen generation and liquefaction facilities. On-site fuel gas and power generation, originally a part of the plant complex, have now been eliminated. Fuel gas and power are now treated as purchased utilities. Costs for on-site generation of fuel gas have been developed, however, and these are presented.

2.0 INTRODUCTION

Starting in 1974, and over a period of about 5 years, several NASA-sponsored study projects were conducted on the technology and economics of the production and liquefaction of hydrogen. These studies supplemented and supported other studies which were made on the application of liquid hydrogen as an alternate and replacement fuel for the petroleum-based Jet A fuel presently used in jet aircraft transportation. The economic studies in all of these projects were based on constant mid-1974 dollars. Since that time, costs have escalated at an average annual rate of approximately 7% so that today (early 1979), they are approximately 40% greater than they were in mid-1974. To permit better economic analyses to be made on liquid hydrogen applications in the market place of today and the near future, the present study was authorized in which the economics presented in the summary reports of three previous projects were updated based on mid-1980 dollars. The rate of return on capital investment and the investment rate on debt were also increased over previous levels. The reports involved are:

1. NASA CR-132631 - Survey Study of the Efficiency and Economics of Hydrogen Liquefaction; Tables 18 through 26.

2. NASA CR-145077 - Study of the Potentials for Improving the Efficiency and Economics of Liquid Hydrogen Produced from Coal; Tables 7, 8, and 21.

3. NASA CR-145282 - Study of the Potential for Improving the Economics of Hydrogen Liquefaction through the Use of Centrifugal Compressors and the Addition of a Heavy Water Plant; Tables 16 through 23, 29, and 34 through 36.

In addition to updating the economics presented in these reports, the scope of operation of the hydrogen plant was revised. Previous studies were based on complete on-site generation of power required by the plant. Power gasifiers were provided, using the Koppers-Totzek Process to provide fuel gas. The fuel gas was used partly to supply process heat and generate process steam and partly to fuel gas turbines which were coupled to electrical generators to provide electricity, to be used mostly as motive power for compression machinery. In the current study, the power gasification section of the plant was deleted and the purchase of electrical power was assumed. The plant, therefore, presently consists of only the coal gasification section for generation of feedstock plus the hydrogen liquefaction complex. Rather than establish a fixed rate for electrical power, a variable cost is assumed, permitting the effect of power rate on the unit cost of liquid hydrogen to be calculable. The cost of coal as raw material and the cost of fuel gas are treated similarly. Economic tables of operating cost and unit production cost are all expressed in linear equation form as functions of electric power rate, the cost of coal and the cost of fuel gas.

3.0 RESULTS

Tables of updated economic data are presented in the three appendices to this report.

Appendix A

Updated tables for "Survey Study of the Efficiency and Economics of Hydrogen Liquefaction", NASA CR-132631, Tables 18 through 26.

Appendix B

Updated tables for "Study of the Potentials for Improving the Efficiency and Economics of Liquid Hydrogen Produced from Coal", NASA CR-145077, Tables 7, 8, and 21.

Appendix C

Updated tables for "Study of the Potential for Improving the Economics of Hydrogen Liquefaction through the Use of Centrifugal Compressors and the Addition of a Heavy Water Plant", NASA CR-145282, Tables 16 through 23, 29, and 34 through 36.

A total of 24 tables has been revised. Each table in the three appendices of this report is numbered to correspond to the equivalent table of the report in which it originally appeared. In addition, three tables appear in the main body of this report. Table 1 is a listing of the economic assumptions used in updating the tables which appear in the Appendix. Table 2 presents the analytical equations for determination of unit cost by discounted cash flow and utility financing methods. The dis-counted cash flow equation differs from that used in previous projects because of a change in assumed rate of return. The utility financing equation was not affected although an increase in the interest rate on debt was applied. Table 3 presents unit costs for fuel gas generated via coal gasification. These were generated for optional use in evaluation of unit costs appearing in the tables of updated data.

4.0 COMMENTS ON PROCEDURES AND RESULTS

4.1 NASA CR-132631 (Ref. 1)

Tables 18 through 22 update the economics of hydrogen liquefaction complex requiring, primarily, the escalation to mid-1980, of costs for capital equipment and for operating the plant. Table 21, presenting operating costs, includes a variable term for the cost of electrical power expressed as a function of the power rate, Rp, in cents per kWh. This variable term also appears in Tables 18, 19, and 22, which present unit costs for DCF and Utility financing and for present and future technologies.

Table 22, which presents unit liquefaction costs based on technology projected to the 1985-2000 AD time period, includes only inflationary effects on the dollar. The technological developments projected for this time frame remain the same as in the original report.

Table 23 presents updated capital investment for coal gasification for both present and future technologies. This table differs from the corresponding table in the original report in that it now includes the coal gasification equipment for only feedstock generation; the power gasification equipment, which was originally included, has been deleted.

During the revision of Tables 24 and 25, it appeared desirable to present the operating cost for coal as a variable in a manner similar to the presentation of the cost of electricity. The cost of coal is, therefore, presented as a function of the unit cost of coal, R_C, in $$ per million Btu. This is included to represent the cost of fuel gas needed by the feedstock gasification section for:

1. Drying of coal
2. Generation of steam for the water gas shift
3. Generation of steam for CO_2 removal

as shown in Figure 20 of CR-132631. There was originally some inclination to provide this fuel gas by adding a scaled down power gasification unit similar to that shown in Figure 21 of CR-132631. The final decision was to present the data in terms of a variable gas cost, which would be consistent with the treatment for electrical power cost. This decision was preceded, however, by the derivation of the several relationships for the unit cost of fuel gas produced from a power gasification unit sized to generate the required amount of fuel gas. These relationships are, therefore, available and are presented in Table 3 for optional use, as desired.

The derivation of the costs of coal gasification in terms of 3 variable costs required the presentation of the annual operating cost, the startup cost and the working capital as a 4-term linear equation in 3 variables. For compactness in presentation, tabular values of the coefficients for the 4 terms of the linear equation are presented in Tables 24 and 25. Overall unit costs for feedstock gasification, however, are presented in equation form.

Table 26 represents a combination of the data presented in Tables 18, 19, 22, 24, and 25 for the total unit cost of liquid hydrogen via coal gasification for feedstock, and purchased energy.

4.2 NASA CR-145077 (Ref. 2)

Table 7 presents comparative investments for the standard and pressurized Koppers-Totzek gasifier complex. Again, only feedstock gasification equipment is included; power is assumed to be purchased instead of being produced on-site.

The purchased-power assumption is continued in Table 8 for the determination of annual operating cost comparisons. Lack of a power gasification section requires the purchase of fuel gas for coal drying, and steam generation for the water gas shift and CO_2 removal. Again, the cost of coal, electrical power, and fuel gas are presented as variable costs which are a function of the unit costs of each.

The contents of Table 21 have been expanded to include net savings, resulting from implementation of loss reduction measures, expressed as a percentage of the value of the gaseous hydrogen feedstock and also of the liquid hydrogen product. The first column in the table presents the updated capital expenditure required to implement the loss reduction measure. The second column presents the updated capitalized value of the hydrogen saved through implementation of the loss reduction measure. The capitalized value is used to permit direct comparison with the implementation cost. It represents the present value of the sum of the annual values of all the gaseous hydrogen saved over the life of the project. Numerically, it is equal to the annual savings taken over 3.25 years, a relationship derived from the DCF financing method and the financing assumptions employed in this project.

The third column shows the present value of net savings which is the capitalized value of the hydrogen minus the implementation cost. The annual value of net savings is the present value, as tabulated, divided by 3.25.

The fourth and fifth columns list the annual net savings expressed as a percentage of the annual value of the liquid hydrogen product and of the gaseous hydrogen feedstock, respectively. The values for the hydrogen gas and liquid were based on the updated unit costs determined in this project for the 2.63 kg/s (250 TPD) capacity of the single module production unit. In addition to the mid-1980 dollar, utility financing and future technology were assumed. Based upon the equations presented in Tables 25 and 26 of CR-132631 (Appendix A) , and the following unit costs:

Coal	\$0.71/Gj	(\$0.75/MM Btu)
Electricity	3¢/kWh	
Fuel Gas	\$2.84/Gj	(\$3.00/MM Btu)

the unit value for hydrogen is calculated to be as follows:

Liquid H_2 Product	\$1.100/kg	(\$0.499/lb.)
Gaseous H_2 Feedstock	\$0.586/kg	(\$0.266/lb.)

The total annual value is \$86,500,000 for the liquid hydrogen product and \$46,200,000 for the gaseous hydrogen feedstock.

4.3 NASA CR-145282 (Ref. 3)

Table 16 presents capital investment for the feedstock coal gasification complex; the power gasification unit has been deleted. Because the feedstock requirements are the same for both Process RRC and Process CRC, the feedstock gasification equipment is the same for each and no difference exists in capital investment.

Differences do exist between the annual operating costs for the two processes, Table 17. This results from different sources of motive power for the feedstock compressor of the gasifier complex. Process RRC uses a compressor driven by an electric motor while the compressor for Process CRC is steam-turbine driven. Consequently, the electrical power requirement for Process RRC is high relative to that for Process CRC in which only modest power requirements are supplied by electrical power. Alternatively, the fuel gas requirement for Process CRC is high relative to that for Process RRC because fuel gas is combusted to produce steam as motive power for the feedstock compressor. In all other respects, annual operating costs for feedstock gasification are equal for both processes.

Tables 18 and 19 present unit gasification costs for feedstock gasification based upon DCF and Utility financing methods. Unit costs are presented as linear functions in three variables; the unit cost of coal, the unit cost of electricity and the unit cost of fuel gas.

Table 20 gives the capital investment for the hydrogen liquefaction complex for both Process RRC and Process CRC. Inasmuch as steam is the motive power for the compressor used in Process CRC, the cost of the power system, including steam turbines, steam boilers, and boiler feedwater pumps, has been included as part of the liquefier cost. A cost breakdown showing the contribution of the power system to the total cost is also presented.

Annual operating costs for the liquefaction complex of the two competing processes, presented in Table 21, reflect the difference in motive power for the two systems. The principal item in the total operating cost for Process RRC is electricity, while for Process CRC it is fuel gas. Electricity is used in the latter process only for auxiliaries, lighting and other minor requirements. This electricity is purchased, rather than generated by the on-site power system as in the original report. Fuel gas is also purchased.

Table 22 presents the unit liquefaction cost for the two competing processes. This includes energy costs for liquefaction as well as investment-related costs for the power system provided with Process CRC. Total unit cost for producing liquid hydrogen, including feedstock gasification, is shown in Table 23.

A relationship for breakeven cost between Process RRC and Process CRC is obtained by equating the unit cost equations for the two processes. The following relations are obtained:

For DCF Financing:

$$R_p = 0.0077 + 1.052\ R_g$$

For Utility Financing:

$$R_p = 0.0615 + 1.0508\ R_g$$

The economics of Process CRC will be favored for all electrical power rates greater than given by these two relations. The interesting result is that the favored process is determined almost completely by the relative costs of electricity and fuel gas. High fuel gas cost and a low electricity rate will favor Process RRC while low fuel gas costs and a high electricity rate will favor Process CRC. Differences in plant investment are not significant and the cost of coal is not a factor at all.

Table 29 updates the capital investment for the deuterium recovery facility to mid-1980 dollars.

Tables 34 and 35 present the annual operating cost and the unit cost of heavy water product assuming purchased electrical power. Liquid hydrogen was evaluated at a fixed price of $0.50 per lb. for the loss of product item. Oxygen was evaluated at 30¢/100 cu.ft.

Because of the deletion of the power gasification section of the plant and the assumption of purchased coal and utilities, the economic impact of heavy water production had to be shown, in Table 36, in equation form as a function of the cost of coal, electricity and fuel gas. The selling price of heavy water is taken to be $100 per lb. which is the estimated price when sold in the present commercial market.

5.0 REFERENCES

1. NASA CR-132621 - Survey Study of the Efficiency and Economics of Hydrogen Liquefaction.

2. NASA CR-145077 - Study of the Potentials for Improving the Efficiency and Economics of Liquid Hydrogen Produced from Coal.

3. NASA CR-145282 - Study of the Potential for Improving the Economics of Hydrogen Liquefaction through the Use of Centrifugal Compressors and the Addition of a Heavy Water Plant.

TABLE 1

BASIS FOR ECONOMIC EVALUATIONS

1. Mid-1980 dollars.
2. Plant Capacity = 26.25 kg/s (2500 TPD) Liquid H_2.
3. Economic evaluation via DCF and Utility financing methods - Table 2.
4. Interest during construction at 11% for 1.875 years.
5. Startup costs at 20% of total annual gross operating cost for coal gasifiers.
6. Startup costs at 2.75% of total plant investment for H_2 liquefier.
7. Working cpaital at 0.9% of total plant investment plus raw materials (coal) inventory of 60 days for coal gasifiers.
8. Working cpaital at 0.9% of total plant investment plus net receivables at 1/24 of annual liquid hydrogen revenue at \$4.26/Gj (\$4.50/MM Btu) for H_2 liquefier.
9. Makeup water at 13.2¢/m^3 (50¢/M gal).
10. Operating labor at \$8.60/hr.
11. Maintenance labor (annual) at 1.5% of total plant investment.
12. Administration and Overhead at 60% of total labor.
13. Operating supplies at 30% of operating labor.
14. Maintenance supplies (annual) at 1.5% of total plant investment.
15. Local taxes and insurance at 2.7% of total plant investment.
16. Sale of by-product sulfur at 5.22¢/kg (\$53/long ton).
17. 8322 operating hours per year (95% on stream).
18. Coal, electricity and fuel gas utilites presented in terms of variable unit cost.

TABLE 1 (Continued)

19. Transmission of feedstock from gasifier to liquefaction complex not included.

20. Land acquisition not included.

21. DCF financing based on:

    25 year project life

    16 year sum-of-the-years'-digits depreciation on Total Plant Investment

    100% Equity Capital

    15% DCF return rate

    48% Federal income tax rate

    No escalation

21. Utility financing based on:

    20 year project life

    5% per year straight line depreciation on Total Capital Requirement excluding Working Capital

    48% Federal income tax rate

    3/1 Debt-equity ratio

    11% Interest rate on debt

    15% Return on equity

    No escalation

22. Heavy water selling price at $220/kg ($100/lb.)

TABLE 2

FINANCING METHODS

Definitions

I - Total Plant Investment

S - Startup Costs

W - Working Capital

C - Total Capital Requirement

N - Total Net Annual Operating Cost

G - Annual Liquid H_2 Production = 786.43 Mg (1733.8 MM lb)

a - Escalation Factor (= 1.0)

d - Fraction debt = 0.75

i - Interest rate on dept = 11.0%

r - Return on Equity = 15.0%

p - Return on rate base

Discounted Cash Flow (DCF) Financing

$$\text{Unit Cost} = \frac{aN + 0.31085I + 0.15470S + 0.28846W}{G}$$

Utility Financing

$$\text{Unit Cost} = \frac{aN + 0.05(C-W) + 0.005\,[P + 48/52\,(1 - d)r](C+W)}{G}$$

where:

$$p = d(i) + (1-d)r$$

TABLE 3

UNIT COST OF FUEL GAS VIA COAL GASIFICATION

Capacity - 125 m^3/s (17.1 MMSCFH) Fuel Gas - Present technology

- 117 m^3/s (16.0 MMSCFH) Fuel Gas - Future technology

DCF Financing

Present Technology

Unit Cost = 2.865 + 1.574 Rc + 0.232 Rp $/MM Btu

= 2.715 + 1.492 Rc + 0.220 Rp $/Gj

Future Technology

Unit Cost = 2.510 + 1.574 Rc + 0.145 Rp $/MM Btu

= 2.379 + 1.492 Rc + 0.137 Rp $/Gj

Utility Financing

Present Technology

Unit Cost = 1.686 + 1.532 Rc + 0.231 Rp $/MM Btu

= 1.598 + 1.452 Rc + 0.219 Rp $/Gj

Future Technology

Unit Cost = 1.484 + 1.532 Rc + 0.145 Rp $/MM Btu

= 1.407 + 1.452 Rc + 0.137 Rp $/Gj

Rc - Cost of Coal, $/MM Btu

Rp - Cost of Electricity, ¢/kwh

APPENDIX A

TABLES OF UPDATED ECONOMIC DATA FOR NASA CR-132631

SURVEY STUDY OF THE EFFICIENCY AND ECONOMICS OF HYDROGEN LIQUEFACTION

April, 1975

TABLE A-18
(TABLE 18, REFERENCE 1)

LIQUEFACTION COST
ACTUAL BASE CASE (1980)
DCF FINANCING

26.25 kg/s (2500 TPD) LIQUID H_2

BASIS

25 Years Project Life
16 Years Sum-of-the-Years'-Digits Depreciation
100% Equity Capital
15% DCF Rate of Return
48% Federal Income Tax Rate

I = Total Plant Investment	\$790,900,000
S = Startup Costs	21,750,000
W = Working Capital	26,981,000
N = Total Net Annual Operating Cost	43,998,000 + 98,381,000 Rp
G = Annual Liquid H_2 Production, lb	1733.8×10^6
a = Escalation Factor	1.0
R_p = Cost of Electricity, ¢/kWh	

$$\text{Unit Liquefaction Cost} = \frac{aN + 0.31085I + 0.1547S + 0.2885W}{G}$$

$$= \frac{1(43.998+98.381Rp)+0.31085(790.9)+0.1547(21.75)+0.2885(26.98)}{1733.8}$$

$$= (0.1736 + 0.0567\ Rp) \qquad \$/lb$$

$$= 0.3827 + 0.1251\ Rp \qquad \$/kg$$

TABLE A-19

(TABLE 19, REFERENCE 1)

LIQUEFACTION COST

ACTUAL BASE CASE (1980)

UTILITY FINANCING

26.25 kg/s (2500 TPD) LIQUID H_2

BASIS

20 Year Project Life

5% Per Year Straight Line Depreciation on Total Capital Requirement Excluding Working Capital

48% Federal Income Tax Rate

C = Total Capital Requirement	$1,002,730,000
W = Working Capital	26,980,000
N = Total Net Operating Annual Cost	43,998,000 + 98,381,000 R_p
G = Annual Liquid H_2 Production	1733.8×10^6 lb/yr
a = Escalation Factor - No Escalation	1.00
R_p = Cost of Electricity, ¢/kWh	

$$\text{Unit Liquefaction Cost} = \frac{aN + 0.05(C-W) + 0.005\left[p + \frac{48}{52}(1-d)r\right](C + W)}{G}$$

$$= \frac{1.0(43.998+98.381R_p) + .05(975.75)+0.005\left[12+\frac{48}{52}(.25)15\right](1029.71)}{1733.8}$$

$$= (0.0994 + 0.0567\ R_p) \qquad \$/\text{lb}$$

$$= 0.2192 + 0.1251\ R_p \qquad \$/\text{kg}$$

TABLE A-20
(TABLE 20, REFERENCE 1)

CAPITAL INVESTMENT
LIQUEFACTION COMPLEX
ACTUAL BASE CASE - 1980

26.25 kg/s (2500 TPD) LIQUID H_2

Total Plant Investment	$ 790,900,000
Interest During Construction	163,100,000
Startup Costs	21,750,000
Working Capital	26,980,000
Total Capital Requirement	$1,002,730,000

TABLE A-21

(TABLE 21, REFERENCE 1)

ANNUAL OPERATING COST
LIQUEFACTION COMPLEX
ACTUAL BASE CASE - 1980

26.25 kg/s (2500 TPD) LIQUID H_2

RAW MATERIALS			
Feedstock - From Coal Gasifier			
CHEMICALS AND ADSORBENTS			
H_2SO_4:	4,000 LB/HR @ $60/ton	$	999,000
Dessicants & Adsorbents:	450,000 LB/YR @ $1.00/lb.		450,000
UTILITIES			
Makeup Water:	15,000 GPM @ 50¢/MGAL		3,745,000
Electricity:	1,182,180 KW @ R_p¢/kWh		98,381,000 R_p
Labor:			
Operating Labor			2,576,000
Supervision			323,000
ADMINISTRATION AND OVERHEAD			1,914,000
SUPPLIES			
Operating			773,000
Maintenance			11,864,000
TAXES AND INSURANCE			21,354,000
	Total Operating Cost	$	43,998,000 + 98,381,000 R_p

TABLE A-22
(TABLE 22, REFERENCE 1)

LIQUEFACTION COST
PROJECTED TO 1985-2000 TIME PERIOD

26.25 kg/s (2500 TPD) LIQUID H_2

DCF FINANCING

BASIS

Same as Table 18 except:

I = Total Plant Investment	=	\$743,400,000
W = Working Capital	=	26,550,000
N = Total Net Annual Operating Cost	=	41,238,000 + 80,574,000 R_p
S = Startup Costs		21,750,000

$$\text{Unit Cost} = \frac{1.0(41.238+80.574R_p)+0.31085(743.4)+0.1547(21.75)+0.2885(26.55)}{1733.8}$$

$$= (0.1634 + 0.0465\ R_p) \quad \$/lb$$

$$= 0.3602 + 0.1025\ R_p \quad \$/kg$$

UTILITY FINANCING

BASIS

Same as Table 19 except:

C = Total Capital Requirement	\$945,026,000
W = Working Capital	26,550,000
N = Total Net Annual Operating Cost	41,238,000 + 80,574,000 R_p

$$\text{Unit Cost} = \frac{1.0(41.238+80.574\ R_p)+0.05(918.48)+0.005\left[12+\frac{48}{52}(.25)(15)\right](971.58)}{1733.8}$$

$$= (0.0936 + 0.0465\ R_p) \quad \$/lb$$

$$= 0.2063 + 0.1025\ R_p \quad \$/kg$$

TABLE A-23

(TABLE 23, REFERENCE 1)

CAPITAL INVESTMENT
COAL GASIFICATION TO H_2 FEEDSTOCK

26.25 kg/s (2500 TPD) LIQUID H_2

$ MILLIONS

SECTION	TECHNOLOGY CURRENT	FUTURE*
H_2 Gas Production, Coal Preparation and Water Gas Shift	280.4	291.3
Raw Gas Compression	86.3	--
H_2 Gas Purification Sulfur and CO_2 Removal	104.5	100.3
O_2 Plant and Compressors	181.3	159.9
Steam Generation and Power	26.7	23.7
Water Treatment and Cooling	16.6	15.9
Electrical Substation and Switchgear	35.5	34.1
General Facility, Roads, Building, Etc.	14.0	13.4
Sub-Total Plant Investment	745.3	638.6
Project Contingency @ 15 percent	111.8	95.8
Total Plant Investment	857.1	734.4

* 1985-2000 AD

TABLE A-24

(TABLE 24, REFERENCE 1)

GASIFICATION COST

DCF FINANCING

26.25 kg/s (2500 TPD) LIQUID H_2

I = Total Plant Investment, \$ Million

S = Startup Costs, \$ Million

W = Working Capital, \$ Million

N = Total Net Annual Operating Cost, \$ Million

G = Annual Liquid H_2 Production, 1733.8 MM lb

a = Escalation Factor = 1.0

R_p = Electrical Power Rate, ¢/KWh

R_c = Cost of Coal, \$/MM Btu

R_g = Cost of Fuel Gas, \$/MM Btu

$$\text{Unit Gasification Cost} = \frac{aN + 0.31085\ I + 0.1547\ S + 0.2885\ W}{G}$$

Let: $\text{Cost}(n) = b + cR_c + dR_p + eR_g$ for n = I, S, W, N

TECHNOLOGY	CURRENT				FUTURE*			
	b	c	d	e	b	c	d	e
I	857.10	0	0	0	734.40	0	0	0
S	13.63	32.27	5.14	7.83	11.97	31.08	4.03	7.32
W	7.71	28.79	0	0	6.61	26.89	0	0
N	56.65	166.37	25.72	39.13	49.10	155.40	20.15	36.62

UNIT COST

	CURRENT	FUTURE*
\$ per kg	$0.4163+0.2285R_c+0.0337R_p+0.0513R_g$	$0.3575+0.2136R_c+0.0265R_p+0.0480R_g$
\$ per lb	$(0.1888+0.1036R_c+0.0153R_p+0.0233R_g)$	$(0.1622+0.0969R_c+0.0120R_p+0.0218R_g)$

* 1985-2000 AD

\$1/MMBtu = \$0.9478/Gj

TABLE A-25

TABLE 25, REFERENCE 1)

GASIFICATION COST

UTILITY FINANCING

26.25 kg/s (2500 TPD) LIQUID H_2

C = Total Capital Requirement, $ Million

W = Working Capital, $ Million

N = Total Net Annual Operating Cost, $ Million

G = Annual Liquid Hydrogen Production = 1733.8 MM lb/yr

d = Fraction Debt — r = Return on Equity = 15%

i = Interest Rate on Debt = 11% — p = Return on Rate Base = 12.0%

a = Escalation Factor = 1.0

$$\text{Unit Gasification Cost} = \frac{aN + 0.05(C-W) + 0.005\left(p + \frac{48}{52}(1-d)r(C+W)\right)}{G}$$

Let: $\text{Cost}(n) = b + cR_c + dR_p + eR_g$ for n = I, S, W, Int, C, N

TECHNOLOGY	CURRENT				FUTURE*			
	b	c	d	e	b	c	d	e
I	857.10	0	0	0	734.40	0	0	0
S	13.63	32.27	5.14	7.83	11.97	31.08	4.03	7.32
W	7.71	28.79	0	0	6.61	26.89	0	0
Interest	176.78	0	0	0	151.47	0	0	0
C	1055.20	61.06	5.14	7.83	904.45	57.97	4.03	7.32
N	56.65	166.37	25.72	39.13	49.10	155.40	20.15	36.62

UNIT COST:

\$/kg $0.2432+0.2225R_c+0.0335R_p+0.0510R_g$ $0.2091+0.2079R_c+0.0263R_p+0.0478R_g$

\$/lb $(0.1103+0.1009R_c+0.0152R_p+0.0232R_g)$ $(0.0948+0.0943R_c+0.0119R_p+0.0217R_g)$

* 1985-2000 AD

$1/MM Btu = $0.9478/Gj

TABLE A-26

(TABLE 26, REFERENCE 1)

TOTAL UNIT COST OF LIQUID H_2

VIA COAL GASIFICATION

26.25 kg/s (2500 TPD) LIQUID H_2

U_m = Unit Cost of Liquid H_2, \$/kg

U_c = Unit Cost of Liquid H_2, \$/lb

R_c = Cost of Coal, \$/MM Btu

R_p = Cost of Electricity, ¢/kWh

R_g = Cost of Fuel Gas, \$/MM Btu

DCF FINANCING

Technology

Current: $U_m = 0.7990 \quad 0.2285R_c + 0.1587R_p + 0.0513R_g$, \$/kg

$U_c = (0.3624 + 0.1036R_c + 0.0720R_p + 0.0233R_g)$, \$/lb

Future*: $U_m = 0.7178 + 0.2136R_c + 0.1290R_p + 0.0480R_g$, \$/kg

$U_c = (0.3256 + 0.0969R_c + 0.0585R_p + 0.0218R_g)$, \$/lb

UTILITY FINANCING

Technology

Current : $U_m = 0.4623 + 0.2224R_c + 0.1585R_p + 0.0511R_g$, \$/kg

$U_c = (0.2097 + 0.1009R_c + 0.0719R_p + 0.0232R_g)$, \$/lb

Future*: $U_m = 0.4154 + 0.2079R_c + 0.1288R_p + 0.0478R_g$, \$/kg

$U_c = (0.1884 + 0.0943R_c + 0.0584R_p + 0.0217R_g)$, \$/lb

* 1985-2000 AD

\$1/MM Btu = \$0.9478/Gj

APPENDIX B

TABLES OF UPDATED ECONOMIC DATA
FOR NASA CR-145077

STUDY OF THE POTENTIALS FOR IMPROVING THE EFFICIENCY AND ECONOMICS OF LIQUID HYDROGEN PRODUCED FROM COAL

July 1976

TABLE B-7

(TABLE 7, REFERENCE 2)

TOTAL PLANT INVESTMENT
FEEDSTOCK GASIFICATION COMPLEX
26.25 kg/s (2500 TPD) LIQUID H_2

	MILLIONS OF DOLLARS	
	STANDARD K-T GASIFIERS	PRESSURIZED K-T GASIFIERS
Hydrogen Production, Coal Preparation, and WGS	280.4	318.5
Raw Gas Compression	86.3	---
H_2 Gas Purification Sulfur and CO_2 Removal	104.5	101.8
O_2 Plant and Compressors	181.3	203.2
Power and Steam Generation	26.7	24.7
Electrical Substation and Switchgear	35.5	36.1
Water Treatment and Cooling	16.6	16.6
General Facility, Roads, Building, Etc.	14.0	13.4
Sub-Total Plant Investment	745.3	714.3
Project Contingency at 15%	111.8	107.1
Total Plant Investment	857.1	821.4

TABLE B-8

(TABLE 8, REFERENCE 2)

ANNUAL OPERATING COST

FEEDSTOCK GASIFICATION AND LIQUEFACTION COMPLEX

26 kg/s (2500 TPD) LIQUID H_2

	MILLIONS OF DOLLARS	
	STANDARD K-T GASIFIERS	PRESSURIZED K-T GASIFIERS
RAW MATERIALS		
Coal @ R_c \$/MM Btu	166.37 R_c	166.37 R_c
CHEMICALS, CATALYSTS AND ADSORBENTS	3.04	3.04
UTILITIES		
Water	5.26	5.26
Electricity @ R_p ¢/kWh	124.10 R_p	113.79 R_p
Fuel Gas @ R_g \$/MM Btu	39.67 R_g	39.67 R_g
LABOR - Operating, Maintenance and Supervision	20.31	20.31
ADMINISTRATION AND OVERHEAD	12.19	12.19
SUPPLIES		
Operating	1.85	1.85
Maintenance	24.72	24.19
TAXES AND INSURANCE	44.50	43.53

Total Gross Operating Cost

Standard K-T Gasifiers	\$111.87 + $166.37R_c + 124.10R_p + 39.67R_g$
Pressurized K-T Gasifiers	\$110.37 + $166.37R_c + 113.79R_p + 39.67R_g$
By-Product Credit - Sulfur	(\$11.49)

Total Net Operating Cost

Standard K-T Gasifiers	\$110.38 + $166.37R_c + 124.10R_p + 39.67R_g$
Pressurized K-T Gasifiers	\$ 98.88 + $166.37R_c + 113.79R_p + 39.67R_g$

\$1/MM Btu = \$0.9478/Gj

TABLE B-21

(TABLE 21, REFERENCE 2)

COST EFFECTIVENESS OF LEAKAGE REDUCTION MEASURES

FOR 2.63 kg/s (250 TPD) LH_2

$1,000's

	LEAKAGE SOURCE	IMPLEMENTATION COST	PRESENT VALUE OF H_2 SAVED	NET SAVING(1) $1,000,s	% SAVING (2)	% SAVING (3)
A.	Recycle Compressor	0	5,500	5,500	2.0	3.7
B.	Feed-Booster Compressor	0	886	886	0.3	0.5
C.	Storage and Distribution	300	1,925	1,625	0.6	1.1
D.	Liquefier Cold Box					
	1. Valve Seats	324	937	613	0.2	0.4
	2. Casing Purge	50	1,336	1,286	0.5	0.9
	3. Casing Floor	9	538	529	0.2	0.4
	4. Valve Packing	37	538	501	0.2	0.3
	5. Relief Valves	9	139	130	0.05	0.09
	6. Flanged & Screwed Joints	202	139	(63)*	--	--
	7. Welded & Brazed Joints	46	13	(33)*	--	--
E.	Purifier Cold Box					
	1. Valve Seats	206	436	230	0.08	0.2
	2. Casing Purge	---	---	---	--	--
	3. Casing Floor	---	---	---	--	--
	4. Valve Packing	37	244	207	0.07	0.1
	5. Relief Valves	6	70	64	0.02	0.04
	6. Flanged & Screwed Joints	129	70	(59)*	--	--
	7. Welded & Brazed Joints	29	4	(25)*	--	--
	Total (4)	$978	$12,549	$11,571	4.12	7.72

* These measures would not be implemented

1. Present value of net savings
2. Percentage of value of liquid hydrogen produced
3. Percentage of value of gaseous hydrogen feedstock
4. Includes implemented measures only

APPENDIX C

TABLES OF UPDATED ECONOMIC DATA
FOR NASA CR-145282

STUDY OF THE POTENTIAL FOR IMPROVING THE ECONOMICS OF HYDROGEN LIQUEFACTION THROUGH THE USE OF CENTRIFUGAL COMPRESSORS AND THE ADDITION OF A HEAVY WATER PLANT.

December, 1978

TABLE C-16

(TABLE 16, REFERENCE 3)

CAPITAL INVESTMENT

COAL GASIFICATION FOR FEEDSTOCK PRODUCTION

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000's

	PROCESSES RRC & CRC
H_2 Feed Gas Production, Coal Preparation and Water Gas Shift	$280,400
Raw Gas Compression	86,300
H_2 Feed Gas Purification, Sulfur and CO_2 Removal	104,500
O_2 Plant and Compression	181,300
Power and Steam Generation	26,700
Electrical Substation and Switchgear	35,500
Water Treatment and Cooling	16,600
General Facility, Roads, Buildings, Etc.	14,000
Sub-Total Plant Investment	$745,300
Project Contingency, 15%	111,800
TOTAL PLANT INVESTMENT	$857,100

TABLE C-17
(TABLE 17, REFERENCE 3)

ANNUAL OPERATING COST
COAL GASIFICATION FOR FEEDSTOCK PRODUCTION

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000's

	PROCESS	
	RRC	CRC
COAL CONSUMPTION kg/s (TPH)	193.8(768.9)	193.8(768.9)
COAL @ R_c \$/MM Btu	\$166,368 R_c	\$166,368 R_c
FUEL GAS @ R_g \$/MM Btu	39,134 R_g	58,704 R_g
ELECTRICITY @ R_p ¢/kWh	25,715 R_p	6,142 R_p
CATALYSTS AND CHEMICALS	1,759	1,759
PROCESS WATER	1,515	1,515
LABOR - Process	3,578	3,578
Maintenance	12,857	12,857
Supervision	980	980
ADMINISTRATION AND OVERHEAD	10,449	10,449
SUPPLIES - Operating	1,073	1,073
Maintenance	12,857	12,857
LOCAL TAXES AND INSURANCE	23,142	23,142

Total Gross Operating Cost

Process RRC $68,210 + 25,715$R_p$ + 166,368R_c + 39,134R_g

Process CRC $68,210 + 6,142$R_p$ + 166,368R_c + 58,700R_g

Sale of By-Product Sulfur ($11,490)

Total Net Operating Cost

Process RRC $56,720 + 25,715$R_p$ + 166,368R_c + 39,134R_g

Process CRC $56,720 + 6,142$R_p$ + 166,368R_c + 58,700R_g

$1/MM Btu = $0.9478/Gj

TABLE C-18

(TABLE 18, REFERENCE 3)

FEEDSTOCK GASIFICATION COST

DCF FINANCING

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000,000's

PROCESS RRC

I	-	\$857.1	
S	-	\$13.642 + 5.143 R_p + 33.274 R_c + 7.827 R_g	
W	-	\$7.714 + 28.788 R_c	
N	-	\$56.720 + 25.715 R_p + 166.368 R_c + 39.134 R_g	
a	-	1.0	
G	-	786.43 Gg (1733.8 MM lb)	
Unit Cost	=	0.4165 + 0.0337 R_p + 0.2286 R_c + 0.0513 R_g	$/kg
		(0.1889 + 0.0153 R_p + 0.1037 R_c + 0.0233 R_g)	$/lb

PROCESS CRC

I	-	\$857.1	
S	-	\$13.642 + 1.228 R_p + 33.274 R_c + 11.741 R_g	
W	-	\$7.714 + 28.788 R_c	
N	-	\$56.720 + 6.142 R_p + 166.368 R_c + 58.700 R_g	
a	-	1.0	
G	-	786.43 Gg (1733.8 MM lb)	
Unit Cost	=	0.4164 + 0.0081 R_p + 0.2287 R_c + 0.0770 R_g	$/kg
		(0.1889 + 0.0037 R_p + 0.1037 R_c + 0.0349 R_g)	$/lb

TABLE C-19

(TABLE 19, REFERENCE 3)

FEEDSTOCK GASIFICATION COST
UTILITY FINANCING

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000,000's

PROCESS RRC

I	-	\$857.1
S	-	$\$13.642 + 5.143\ R_p + 33.274\ R_c + 7.827\ R_g$
W	-	$\$7.714 + 28.788\ R_c$
N	-	$\$56.720 + 25.715\ R_p + 166.368\ R_c + 39.134\ R_g$
IDC	-	\$176.78
C	-	$\$1055.24 + 5.143\ R_p + 62.062\ R_c + 7.827\ R_g$
a	-	1.0
G	-	786.43 Gg (1733.8 MM lb)
Unit Cost =		$0.2432 + 0.0335\ R_p + 0.2227\ R_c + 0.0511\ R_g$ \$/kg
		$(0.1103 + 0.0152\ R_p + 0.1010\ R_c + 0.0232\ R_g)$ \$/lb

PROCESS CRC

I	-	\$857.1
S	-	$\$13.642 + 1.228\ R_p + 33.274\ R_c + 11.74\ R_g$
W	-	$\$7.714 + 28.788\ R_c$
N	-	$\$56.720 + 6.142\ R_p + 166.368\ R_c + 58.7\ R_g$
IDC	-	\$176.78
C	-	$\$1055.24 + 1.228\ R_p + 62.062\ R_c + 11.74\ R_g$
a	-	1.0
G	-	786.43 Gg (1733.8 MM lb)
Unit Cost =		$0.2432 + 0.0079\ R_p + 0.2227\ R_c + 0.0765\ R_g$ \$/kg
		$(0.1103 + 0.0036\ R_p + 0.1010\ R_c + 0.0347\ R_g)$ \$/lb

TABLE C-20

(TABLE 20, REFERENCE 3)

CAPITAL INVESTMENT

LIQUEFACTION COMPLEX

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000's

	PROCESS			
	RRC	CRC		
		LIQUEFIER	POWER SYSTEM*	TOTAL
Total Plant Investment	$ 859,700	$744,000	$ 98,800	$ 842,800
Interest During Construction	117,300	153,500	20,400	173,900
Start-Up Costs	23,600	20,500	2,700	23,200
Working Capital	27,600	26,600	900	27,500
TOTAL Capital Requirement	$1,088,200	$944,600	$122,800	$1,067,400

* Includes - Steam Turbines, Steam Boilers and Boiler Feedwater Pumps.

TABLE C-21

(TABLE 21, REFERENCE 3)

ANNUAL OPERATING COST

LIQUEFACTION COMPLEX

26.25 kg/s (2500 TPD) LIQUID H_2

	PROCESS	
	RRC	CRC
RAW MATERIALS		
Feedstock - From Coal Gasifier	---	---
CHEMICALS AND ADSORBENTS		
Sulfuric Acid	$ 999,000	$ 999,000
Dessicants and Adsorbents	450,000	450,000
UTILITIES		
Make-Up Water	3,745,000	3,745,000
Electricity @ R_p ¢/kWh	102,655,000 Rp	10,025,500 Rp
Fuel Gas to Steam Boilers @ R_g $/MM Btu	---	98,432,000 Rg
LABOR		
Operating	2,576,000	2,576,000
Supervision	323,000	323,000
ADMINISTRATION AND OVERHEAD	1,914,000	1,914,000
SUPPLIES		
Operating	773,000	773,000
Maintenance	12,895,000	12,642,000
LOCAL TAXES AND INSURANCE	23,212,000	22,756,000
TOTAL OPERATING COST	$46,887,000+	$46,178,000+
	$102,655,000 Rp	$10,025,500 Rp+
		$98,432,000 Rg

$1/MM Btu = $0.9478/Gj

TABLE C-22

(TABLE 22, REFERENCE 3)

UNIT LIQUEFACTION COST

26.25 kg/s (2500 TPD) LIQUID H_2

$1,000's

	PROCESS	
	RRC	CRC
I - TOTAL Plant Investment	$ 859,700	$ 842,800
S - Start-Up Costs	23,600	23,200
W = Working Capital	27,600	27,500
C - TOTAL Capital Requirement	1,088,200	1,067,400
N - TOTAL Net Annual Operating Cost	46,887+	46,178+
	102,655 Rp	10,025.5 Rp+
		98,432 Rg
G - Annual LH_2 Production, Mg/Yr (MM LB/YR)	786.43 (1,733.8)	786.43 (1,733.8)
A - Escalation Factor	1.0	1.0

DCF FINANCING

Unit Cost

Process RRC: 0.4143 + 0.1305 Rp $/kg

(0.1879 + 0.0592 Rp) $/lb

Process CRC: 0.4065 + 0.0128 Rp + 0.1252 Rg $/kg

(0.1844 + 0.00578 Rp + 0.05677 Rg) $/lb

UTILITY FINANCING

Unit Cost

Process RRC: 0.2368 + 0.1305 Rp $/kg

(0.1074 + 0.0592 Rp) $/lb

Process CRC: 0.2324 + 0.01275 Rp + 0.1252 Rg $/kg

(0.1054 + 0.00578 Rp + 0.05677 Rg) $/lb

TABLE C-23

(TABLE 23, REFERENCE 3)

TOTAL UNIT COST OF LIQUID H_2

Rp = Electrical Power Rate, ¢/KWh

Rc = Cost of Coal, \$/MM BTU

Rg = Cost of Fuel Gas, \$/MM BTU

\$1/MM Btu = \$0.9478/Gj

DCF FINANCING

Process RRC

Unit Cost = 0.8307 + 0.1642 Rp + 0.2286 Rc + 0.0514 Rg, \$/kg

(0.3768 + 0.0745 Rp + 0.1037 Rc + 0.0233 Rg), \$/lb

Process CRC

Unit Cost = 0.8318 + 0.0209 Rp + 0.2286 Rc + 0.2021 Rg, \$/kg

(0.3773 + 0.0095 Rp + 0.1037 Rc + 0.0917 Rg), \$/lb

UTILITY FINANCING

Process RRC

Unit Cost = 0.4799 + 0.1640 Rp + 0.2227 Rc + 0.0511 Rg, \$/kg

(0.2177 + 0.0744 Rp + 0.1010 Rc + 0.0232 Rg), \$/lb

Process CRC

Unit Cost = 0.4755 + 0.0207 Rp + 0.2227 Rc + 0.2017 Rg, \$/kg

(0.2157 + 0.0094 Rp + 0.1010 Rc + 0.0915 Rg), \$/lb

TABLE C-29

(TABLE 29, REFERENCE 3)

INVESTMENT SUMMARY

DEUTERIUM RECOVERY FACILITY

FOR 10-MODULE 26.25 kg/s (2500 TPD) H_2 LIQUEFIER

PRODUCING 126.5 kg/h (6695 lb/day) HEAVY WATER

TOTAL PLANT INVESTMENT	$95,455,000
INTEREST DURING CONSTRUCTION	19,688,000
STARTUP COSTS	2,625,000
WORKING CAPITAL	10,533,000
TOTAL CAPITAL REQUIREMENT	$128,301,000

TABLE C-34

(TABLE 34, REFERENCE 3)

UNIT COST OF HEAVY WATER

DCF FINANCING

$1,000's

ANNUAL OPERATING COST			
Electricity	$9195.8\ R_p$		
Loss of LH_2 Product	116.8		
Oxygen for D_2 Combustion	67.0		
Labor & Supervision	430.0		
Administration & Overhead	258.0		
Supplies, Operating	113.0		
Maintenance	1432.0		
Local Taxes and Insurance	2577.0		
TOTAL	4993.8	+	$9195.8\ R_p$
TOTAL PLANT INVESTMENT	95,455		
STARTUP COST	2,625		
WORKING CAPITAL	10,533		

Unit Cost of D_2O, = $36.18 + 8.730\ R_p$ \$/kg

$(16.41 + 3.960\ R_p)$ \$/lb

Where

R_p = Cost of Electricity, ¢/kWh

G = Annual Production of Heavy Water

= 1.053 Gg (1161 tons)

TABLE C-35

(TABLE 35, REFERENCE 3)

UNIT COST OF HEAVY WATER

UTILITY FINANCING

$1,000's

ANNUAL OPERATING COST

	Electricity	9,195.8 Rp
	Loss of LH_2 Product	116.8
	Oxygen for D_2 Combustion	67.0
	Labor & Supervision	430.0
	Administration & Overhead	258.0
	Supplies, Operating	113.0
	Maintenance	1,432.0
	Local Taxes and Insurance	2,577.0
		4,993.8 + 9,195.8 Rp
TOTAL Capital Requirement		$128,301
Working Capital		10,533

Unit Cost of D_2O, $/kg   20.52 + 8.730 Rp

($/lb)   (9.309 + 3.960 Rp)

Rp = Cost of Electricity, ¢/KWh

G = Annual Production of Heavy Water

= 1.053 Gg (1161 Tons)

TABLE C-36
(TABLE 36, REFERENCE 3)

ECONOMIC SUMMARY

IMPACT OF HEAVY WATER PRODUCTION

Basis: 1.053 Gg (1161 T) D_2O Annually - From 26.25 kg/s (2500 TPD) LH_2

	COST $ MILLIONS
TOTAL Plant Investment,	
Deuterium Recovery & H_2 Liquefier Modification	95.455
Income At $220.46/kg ($100/lb) D_2O	232.20
Annual Cost of D_2O	
DCF Financing	38.10 + 9.195 Rp
Utility Financing	21.62 + 9.195 Rp
Annual Cost of LH_2	
DCF Financing	628.33 + 179.62 Rc + 124.83 Rp + 40.40 Rg
Utility Financing	363.58 + 174.94 Rc + 124.66 Rp + 40.22 Rg
Net Income	
DCF Financing	194.10 - 9.195 Rp
Utility Financing	210.58 - 9.195 Rp
Net Income as % of LH_2 Cost	
DCF Financing	$\frac{100(194.10 - 9.195\ Rp)}{628.33 + 179.62\ Rc + 124.83\ Rp + 40.40\ Rg}$
Utility Financing	$\frac{100(210.58 - 9.195\ Rp)}{363.58 + 174.94\ Rc + 124.66\ Rp + 40.22\ Rg}$