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SYNTHETIC CARBONACEOUS FUELS AND FEEDSTOCKS FROM OXIDES OF CARBON AND NUCLEAR POWER

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ABSTRACT

A study has been made on converting alternative sources of oxides of carbon (CO and CO_2) to synthetic methanol with nuclear power generated electrolytic hydrogen and oxygen. The sources of CO and CO_2 include (1) the oxygen blown blast furnace which produces CO as a by-product from hot metal production, (2) the steam calcination of limestone which produces CO_2 as a by-product of the lime and cement industry, (3) fossil fuel power plant stack gas as a source of CO_2 and, (4) the atmosphere, from which CO_2 is recovered in a novel carbonate electrolytic cell. The recovered CO or CO_2 is catalytically combined with electrolytic H_2 to produce the synthetic fuels. In these systems, the conservation, efficiency, and environmental control of coal utilization is significantly improved. Blast furnace CO could supply up to 20% of the gasoline demand in the US at a 1985 cost of 56 to 60¢/gallon, breaking even with $19/Bbl imported oil. The CO_2 from steam calcination of limestone could supply about 9% of the gasoline demand and the recovered CO_2 from only 60% of the coal fired powered plants in the country could supply all of todays gasoline demand in the country (∿ 100 billion gallons/yr). The 1985 cost estimates for carbon dioxide based gasoline range from 68¢/gallon for the highly concentrated calciner CO_2 feedstream to 83¢/gallon for the very dilute

atmospheric CO_2 feed, breaking even with $21 to $28/barrel oil for conventional gasoline. The dominating cost factor is the electrical power cost from the nuclear plant. The sharing of peaking and base load costs between the power and synthetic fuels consumers offers a cost and energy effective system.

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INTRODUCTION

It is fairly well recognized by now that the principal energy deficit in the US relates to inadequate supplies and production of fluid hydrocarbon fuels, i.e., fuel oil, gasoline, and natural gas. The government policy in the near future is first, to reduce demand by conservation measures, i.e., lighter more efficient automobiles, better home insulation, and more efficient industrial operations and second, to foster the utilization of our indigenous coal reserves for power production and for conversion to gaseous and liquid synthetic hydrocarbon fuels.

There is increasing concern about the problems of strip and deep coal mining, environmental factors, water requirements, transportation facilities, and the increased cost associated with coal production conversion and utilization. Indeed, the ultimate concern may be the effects of disposal of increasing quantities of carbon dioxide to the atmosphere by increased coal utilization (Ref. 1).

Nuclear energy seems to be relegated to electrical power production only because it is thought to be inflexible as far as synthetic fuels are concerned. As has been shown in several recent studies (Ref. 2) this is a popular misconception since nuclear power can be utilized to generate synthetic fuels through the thermal and electrolytic decomposition of water in combination with the processing of carbonaceous raw materials. Nuclear energy can thus conserve fossil energy in a two-fold way; not only for production of electrical power but also for production of carbonaceous fuels.

Alternative non-fossil sources of hydrocarbon fuel include biomass and agricultural solid waste. These fundamentally rely on solar energy which is limited to large terrestrial surface areas, and the vagaries of climatic conditions, low efficiencies, and high costs.

The greatest impact on conservation of fossil fuel would be to use nuclear power to reduce oxidized forms of carbon with nuclear generated hydrogen. The major sources of carbon dioxide and carbon monoxide are:

1. Fossil fueled steam and power plants.
2. Industrial chemical plants (lime, cement, ammonia, etc.).
3. Metallurgical operations, blast furnace operations.
4. Natural gas wells.
5. Geothermal wells.
6. Limestone calcination.
7. Seawater
8. The atmosphere.

In this paper, we will concentrate on further development of three sources of carbon oxides for conversion to synthetic fuels and feedstocks, (1) blast furnace, (2) limestone calcination and, (3) coal fired power plants and, (4) a new system for fixation of atmospheric CO_2.

II. OXYGEN BLOWN BLAST FURNACE AS A SOURCE OF CO

The steel industry presently uses a hot air blast furnace to reduce iron ore with coke. The CO produced for a total US iron capacity of 125 x 10^6 tons/yr roughly amounts to 125 x 10^6 tons/yr of CO. However, the top gas concentration is only 20% CO diluted with nitrogen and CO_2, and thus has a low BTU value (80 BTU/ft^3). If the hot air is replaced with an oxygen blast diluted with recycled CO_2, the concentration of CO in the top gas increases to 80% producing an intermediate BTU gas (250 BTU/ft^3) (Ref. 4). This gas can be readily used for conversion to synthetic carbonaceous fuels. In addition, by blowing with enriched oxygen, the hot metal capacity of the blast furnace effectively doubles and the present regenerative air preheat ovens are eliminated. With oxygen enrichment, the blast furnaces runs at a higher temperature. In the conventional approach to the oxygen blown blast furnace for synthetic fuel production, the oxygen is supplied to the furnace from an air liquefaction plant. In this conventional case, 2/3 of the top CO gas must be shifted to hydrogen and the remaining 1/3 is catalytically converted with the hydrogen to synthetic fuel. With nuclear power a decided process improvement can be achieved. Employing nuclear power generated electrolytic decomposition of water, the oxygen can be used to blast the furnace and the hydrogen can be mixed with the concentrated CO in the top gas to make up synthesis gas for the catalytic conversion to methanol and by catalytic dehydration to synthetic gasoline (Ref. 5). In this manner, utilization of the CO in the top gas is increased three-fold. The flow sheet for this scheme is shown in Figure 1. According to the stoichiometry, the system is well balanced and the overall reactions are essentially:

$$yC + mO_2 + xFe_2O_3 = nCO + zCO_2 + 2xFe$$

$$nH_2O + nCO = (CH_2)_n + nO_2$$

where $(CH_2)_n$ represents the mixed aliphatic and aromatic synthetic gasoline distillate fuel produced in the catalytic conversion processes. It should be noted that in the two-step catalytic process either methanol, gasoline or a blend of the two can be produced on demand.

A preliminary economic analysis of the cost of production of methanol and gasoline from the top gas of an oxygen blown blast furnace was made comparing a conventional air liquefaction oxygen and shift conversion process with a nuclear-electrolytic oxygen and hydrogen system. The blast furnace was assumed to have a nominal conventional capacity of 3000 T/D of hot metal which when blown with oxygen tempered with CO_2 would double the capacity to 6000 T/D and produce roughly 6000 T/D of CO top gas. Table I shows the production capacity of synthetic fuel and the estimated capital cost. With air liquefaction and shift conversion the blast furnace could supply CO to a plant producing 16,500 Bbl/D of methanol or 7,900 Bbl/D of gasoline. The nuclear-electrolytic plant would triple the capacity to 50,000 Bbl/D methanol which can produce 24,000 Bbl/D gasoline or a blend of the two. The 1985 equipment costs were assumed to escalate 80% over the 1975 costs. The production cost estimate is given in Table II. Power cost for the nuclear reactor is assumed to be charged on an incremental off-peak basis as previously estimated (Ref. 3). The economic analysis shows the following:

1. Significant synthetic fuel capacity can be achieved with the combination of the blast furnace and nuclear-electrolytic systems.

2. The unit capital investment is significantly reduced from the conventional blast furnace system because, (a) the regenerative air preheat ovens are eliminated, (b) air liquefaction is eliminated and, (c) CO shift conversion is eliminated. However, electrolyzers are added.

3. The nuclear-electrolytic production cost is much lower than the air liquefaction-shift plant because of the increased capacity of the plant and the use of incremental power cost.

4. If indeed, incremental nuclear power cost can be obtained at 6 mills/kwh(e), synthetic gasoline could be produced for 60¢/gallon or 56¢/gallon if oxygen by-product credit can be realized. These costs should break even with $19/Bbl foreign oil to produce conventional gasoline. If the incremental cost of power would double from 6 to 12 mills, the cost of the gasoline would go up about 32% to 80¢/gallon which breaks even with $26/Bbl oil, a value which some believe will be reached in 10 years.

5. As will be shown later in Table VII if all the blast furnace capacity could be converted to oxygen blown systems, there would be enough CO for conversion to gasoline to supply at least 20% of the automotive fuel market.

6. It should also be noted that if the capital cost for the electrolytic cells were increased to $200/kwh(e) as recently estimated (Ref. 13) instead of the $90 kwh(e) assumed here, the costs would increase by only 12% bringing the gasoline cost to 67¢/gallon.

III. STEAM CALCINATION OF LIMESTONE AS A SOURCE OF CO_2

At least 25 million tons of limestone are calcined to lime a year and over 80 million tons a year of cement is also produced. This amount of limestone decarbonation yields a potential of approximately 60×10^6 tons/yr of CO_2 available from the lime and cement industry in the US. Rotary kilns are fired by either natural gas, oil or coal with calcination temperatures usually reaching about 900°C. A possible conservation measure is to introduce steam into the kiln to lower the calcination temperature by as much as 200°C to 700°C (Ref. 6). This would reduce the energy requirement in the kiln and produce a more reactive lime. The steam effect has been controversial, however, in laboratory studies a 3 to 5 fold increase in rate of calcination has been observed at elevated steam concentrations (Ref. 7). With fossil fuel, the equilibrium and rate of calcination is reduced by the partial pressure of CO_2 from the combustion process. It is thus proposed to replace fossil fuel firing of the kiln with hydrogen-oxygen-steam firing, with the hydrogen and oxygen being produced by the electrolytic decomposition of water using nuclear power. The calcination rate should be markedly increased in a pure steam atmosphere where the partial pressure of CO_2 introduced is zero. Since the stoichiometric H_2/O_2 flame has an extremely high adiabatic temperature ($\sim$ 2800°C), the flame must be tempered with a quantity of steam obtained from the power generation and the heat recuperator at the back end of the kiln to reduce the temperature of the flame to about 2000°C. A schematic process flow sheet of the system is shown in Figure 2. The H_2 from the cells are then combined with the

CO_2 catalytically to form methanol, gasoline or blends. This system is fundamentally different from that proposed previously (Ref. 8) in which a high temperature gas cooled reactor (HTGR) is used to supply 900°C steam to the kilns. This temperature is still a little beyond the capability of present HTGR's which presently can reach about 600°C (1100°F). At higher temperatures, the operation of the reactor may be materials limited. A flame burning free in space, produced by burning H_2-O_2-steam is not materials limited. The H_2 and O_2 generated electrolytically required for firing the kiln can additionally be used for the synthesis of distillate fuel.

It should be pointed out that if stack gas scrubbing with lime takes hold in the US to control sulfur emissions from coal-burning power plants, then a large amount of calcium sulfate will be formed which will require regeneration to reduce limestone costs and to eliminate the limestone waste disposal problem. The steam calcination of sulfated limestone driven by the nuclear-electrolytic system could be applicable for production of by-product synthetic fuels.

IV. FOSSIL FUEL POWER PLANT STACK GAS AS A SOURCE OF CO_2

The largest source of man made CO_2 input to the atmosphere is from the combustion of fossil fuel for heat and power purposes. Central power plants where large quantities of CO_2 are generated in stack gases makes up about 1/3 of the total CO_2 emissions due to fossil fuel burning in the US. The concentration of CO_2 in stack gas varies from about 8% to about 19% in the flue gas depending upon whether gas or coal is burned. Thus both as a large centralized source of CO_2 and a concentrated source of CO_2 it appears logical that if we are to continue burning fossil fuel and especially in view of our plans to rely heavily on coal, for purposes of controlling the CO_2 effluent in the atmosphere, it would be most beneficial and economical to remove and recover the CO_2 from fossil fuel power plants for production of synthetic carbonaceous fuel and feedstock.

A number of processes can be considered for separating the CO_2 from stack gases including, (a) absorption-stripping, (b) cryogenic separation, (c) adsorption-stripping and, (d) diffusion barriers. Relying on a previous study (Ref. 9), it appears that absorption-stripping techniques are the more economical means for recovering and concentrating the CO_2. In view of the fact that environmental control processes are being applied to fossil fuel plants because of particulate and sulfur emissions, CO_2 scrubbing system can be appropriately integrated into the gas scrubbing operation. Figure 3 indicates a schematic flow diagram of the production of synthetic carbonaceous fuels and feedstocks, converting CO_2 from stack gas with nuclear-electrolytic hydrogen. The sequence of operation is electrostatic precipitation followed by lime-limestone wet scrubbers to remove the SO_2 forming a calcium sulfite/sulfate waste (which can be regenerated by desulfation), followed by CO_2 removal using a potassium carbonate absorption-stripping operation (Ref. 12). The concentrated CO_2 can then be compressed to 50 to 100 atm, or alternatively, the stripping towers may be operated at these higher pressures and combined with hydrogen from the cells which also can be operated under higher pressure, thereby eliminating the need for compression. The mixed compressed gas is catalytically converted to methanol and then dehydrated to distillate fuel or it may be converted from the synthesis gas directly to synthetic gasoline.

The conversion of CO_2 from fossil fuel fired power plants for distillate fuel production in effect makes use of the carbon in the fossil fuel twice; one time for producing electrical power and a second time for making synthetic fuel.

V. ATMOSPHERIC CO_2

The ultimate repository for CO_2 from fossil fuel combustion is the atmosphere. A vast resource is already available ($\sim 10^{13}$ tons/yr). A concern is to maintain the present CO_2 balance between the biosphere and the atmosphere. The only methods to halt the increase in CO_2 content of the atmosphere are to cease burning fossil fuel and to reduce the CO_2 back to carbon as fast as we burn it. By substuting nuclear power for fossil fuel we accomplish the first objective and by recovering and reducing atmospheric CO_2 we accomplish the second. Two systems for removal and recovery have been proposed previously (Refs. 2 and 9). One depended on a dilute aqueous carbonate absorption-stripping technique and the other on a chlor-alkali carbonate neutralization system. The former process suffers from a large heat exchange load and the second from high power requirements. A new system is proposed here which combines the best of both systems. The process depends on the electrolytic decomposition of water containing higher concentrations of carbonate/bicarbonate ion, which is essentially the electrolytic decomposition of sodium carbonate/bicarbonate thus forming caustic, H_2, O_2 and CO_2.

$$Na_2CO_3 + 4H_2O = 2NaOH + 3H_2 + 3/2\ O_2 + CO_2$$

Observations at Brookhaven (Ref. 10) indicated that water can be electrolytically decomposed to H_2 and O_2 in a sodium carbonate/bicarbonate cell releasing CO_2 in addition to O_2 at the anode and H_2 at the cathode. The theoretical cell voltage is approximately 3.0 volts per mole. of H_2. This compares to 2.4 volts for the decomposition of water. If a three compartment cell is used, the CO_2 can be separated in a center compartment. Referring to flow sheet schematic Figure 4, the regenerated caustic is then used to scrub the CO_2 from the atmosphere and the CO_2 recovered from the cell is combined with hydrogen catalytically to methanol and gasoline blends. In this case, the oxygen is vented to the atmosphere. It may also be possible to operate the carbonate cells under pressure to match the operating conditions of the catalytic convertors so as to eliminate compression. The cell may operate at elevated temperature (100-150^{o}C) to improve efficiency. The system, however, is essentially an isothermal one for removal and recovery of CO_2 compared to the absorption-stripping process which requires large heat exchange loads. Furthermore, no additional cell equipment, power requirement and corrosion problems are encountered as existed in the earlier proposed chlor-alkali flow sheet (Ref. 3).

The nuclear-electrolytic carbonate system may be of value for stack gas scrubbing of CO_2, however, in this case where CO_2 gas compositions up to 15% are available compared to 0.035% in the atmosphere, adoption of a conventional absorption/stripping technique is probably more economical.

VI. ECONOMIC ANALYSIS OF CO_2 BASED FEEDSTREAMS FOR SCFF PRODUCTION

There is a certain commonality among the three CO_2 based processes for evaluating the economics of the production of synthetic fuels and feedstocks. Using a CO_2 feedstream the common process chemistry is as follows:

1. Nuclear-electrolytic hydrogen

$$3H_2O = 3H_2 + 3/2\ O_2$$

2. Catalytic methanol synthesis

$$CO_2 + 3H_2 = CH_3OH + H_2O$$

3. Catalytic gasoline synthesis

$$nCH_3OH = (CH_2)_n + nH_2O$$

It is noted in producing gasoline or light distillates from CO_2, for every mole of carbon reduced from the oxide to the distillate $(CH_2)_n$ 3 moles of hydrogen must be utilized of which 2 moles are converted to water and are recycled, while only one mole appears in the final product. Thus, a-priori, the system is inherently an energy intensive one.

The basic process data used in performing the preliminary economic evaluations are given in Table III. The capital investment data for the economic analysis is shown in Table IV. All data are given in escalated 1985 dollars based on the references given in the footnotes of the table. One notable difference is the unit cost of electrolytic cells which is assumed to be \$200/kw(e) (Ref. 12) or about double previous estimates. This increased estimate was assumed in the event cost reductions due to large scale development of cells do not materialize. The production cost of methanol and gasoline are shown in Tables V and VI. The main differences in the processes which affects the economics and the capacity, relate to the concentration of the source CO_2 and the volume of feedstream. Since power supply and cost are controlling factors, the production capacity of the synthetic plant is fixed by the nuclear power plant capacity. This was assumed to be an optimized 1000 MW(e) base-loaded nuclear plant operating in an off-peak mode. Thus, about 725 MW(e) of power can be available at an incremental cost of 6 mills/kwh(e) (Ref. 3). Tables V and VI also include the production cost based on the total cost of nuclear power from a dedicated plant which is estimated to be at 30 mills/kwh(e).

The salient conclusion of the cost estimates is that with a reasonable off-peak power cost of 6 mills/kwh, synthetic distillate fuels can be produced at costs ranging from 68.2 to 82.9¢/gallon of gasoline depending on the concentration of CO_2 in the feedstream, going from a pure CO_2 stream to a CO_2 content found in the atmosphere. These costs can break-even with distillate fuel from crude oil estimated at costs ranging from $21.00 to $27.50 per barrel. Thermal cost values in this range are equivalent to from $5.50 to $6.50/MMBTU. It is interesting to note that current cost estimates for coal gasification and liquefaction plants for producing synthetic fuels are estimated to break-even with crude oil at costs ranging from $25 to as high as $40 per barrel (Ref. 13). In view of these costs, the above do not seem unreasonable. The estimates also indicate that the cost of synthetic fuel would increase by two and a half times for a dedicated nuclear plant at the full power cost recovery of 30 mills/kwh(e) which would probably not be competitive in the near term future.

The general observation concerning the cost of reduction of recovered CO_2 from various sources with nuclear-electrolytic hydrogen is that even at present high estimates of capital cost for electrolytic cells (∿ $200/kwh), the fraction of the total production cost due to depreciation of this equipment is not necessarily a controlling factor for the production cost (∿ 28%). As mentioned previously, this cost item might be readily reduced to less than half of this value due to improved cell design and mass production methods. The largest part of the cost, however, is attributed to the consumption of electrical power for the production of electrolytic hydrogen (∿ 40%) and the cost of electrical power from the nuclear plant. As previously mentioned, it takes 3 moles of H_2 to reduce one mole of CO_2 to gasoline (CH_2). Only 1 mole appears in the product while 2 moles are recycled in the form of water. Thus, any reduced form of carbon, such as CO from a blast furnace helps significantly in the conversion of that carbon to synthetic fuels and feedstocks. Along these lines, it might very well be valuable to apply nuclear electrolytic hydrogen and oxygen for underground gasification of coal and conversion to methanol and gasoline. Methods of reducing costs include improving the efficiency of electrolytic decomposition of water by operating at higher temperatures. Another way, as assumed above, is utilizing the power plant in an off-peak mode thus sharing and reducing costs with the fuel producer. This is not necessarily unreasonable because it improves the total energy utilization efficiency in the overall production process, i.e., power and fuel can be coproduced in a base-loaded plant.

Finally in Table VII a summary of the available CO_2 and CO resource capacity from the various industries are given together with the gasoline production capability and the number of 1000 MW(e) nuclear reactors required to convert the carbon oxides to fuel. It appears that the lime and cement industry might be able to supply about 9% of the 100 billion gallons of gasoline consumed in the country today. The steel industry could supply 20% of the gasoline consumption and would require about 125 one-thousand megawatt reactors. If one assumes all the CO_2 recovered from the stacks of coal burning power plants (500 million tons coal per year), is converted to synthetic fuel, this supply would exceed the rate of consumption by 65%. Thus only 60% of the CO_2 from coal burning power plants would be needed to supply the total gasoline demand in the country. This would require 1000 nuclear reactors of 1000 MW(e) size supplying peaking load to the electric power grid and off-peak power to the synthetic fuel plants. Depending on the rate of growth in the next several decades, these numbers can be escalated accordingly.

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Table I

SYNTHETIC CARBONACEOUS FUEL FROM OXYGEN BLOWN BLAST FURNACE TOP GAS

Capital Investment Estimates

Process	Air Liquefaction Oxygen and CO Shift Conversion for Blast Furnace	Nuclear-Electrolytic Oxygen and Hydrogen Supply for Blast Furnace
Furnace Capacity, Hot Metal T/D	6,000	6,000
CO Production T/D	6,000	6,000
O_2 Required T/D	3,000	3,000
H_2 Required T/D	-0-	857
Power Required, MW(e)	31.3 MW(e)(1	1,336 MW(e)(2
Excess O_2 Produced, T/D	-0-	3,860
Methanol Production T/D	2,300	6,900
Bbl/Day	16,500	50,000
Gasoline Production Bbl/Day	7,900	24,000

Process Module:	$ Millions 1975	$ Millions 1985	$ Millions 1975	$ Millions 1985
1. Air Liquefier - 3,000x$12,000/Daily Ton	36	65	--	--
2. Shift Reactor - 16,500x890/Daily Bbl	15	27	--	--
3. Methanol Convertor - 2,300x$16,500/Daily Ton	38	68	115	209
4. Electrolyzers(3	--	--	67	121
5. Contingency	15	27	32	56
Total	104	187	214	386
6. Add on for Methanol Conversion to Gasoline(4	24	43	70	130
Total	128	230	286	516
$/Bbl/Day Gasoline	16,200	29,100	12,000	21,500

Notes:
1) Based on 250 kwh(e)/ton of O_2 for air liquefaction plant.
2) Based on 18.7 kwh(e)/lb H_2 - 30 atm cells.
3) Based on $50/kwh(e) of electrolyzer capacity escalated to $90/kwh(e) in 1985.
4) Derived from Ref. 5.
5) 1985 costs escalated 80% over 10 years from 1975 figures.

Table II

SYNTHETIC CARBONACEOUS FUEL FROM OXYGEN BLOWN BLAST FURNACE TOP GAS

Production Cost Estimates

Process	Air Liquefaction Oxygen Supply to Blast Furnace	Nuclear-Electrolytic Oxygen and Hydrogen Supply to Blast Furnace
Furnace Capacity, Hot Metal T/D	6,000	6,000
Methanol Production, Bbl/Day	16,500	50,000
Gasoline Production, Bbl/Day	7,900	24,000
Oxygen Production, T/D	-0-	3,860

	$/Bbl		$/Bbl	
Production Cost	1975	1985	1975	1985
1. Top Gas Value CO(1	3.17	5.70	1.06	1.90
2. Electricity(2	1.40	2.30	0.20	3.85
3. Operating Labor	0.60	1.00	0.60	1.00
4. Fixed Charges(3	3.25	5.85	2.20	4.00
5. For Equivalent Gasoline	(4.00)	(7.20)	(2.94)	(5.35)
Total Production Cost - Methanol	8.42	14.85	4.06	10.75
For Equivalent Gasoline	(9.17)	(16.20)	(4.74)	(12.10)
Total Gasoline Cost $/Bbl	19.10	33.75	9.85	25.20
¢/Gallon	45.5	80.4	23.5	60.0
Oxygen By-Product ¢/Gallon(4	--	--	2.2	3.9
Net Gasoline Cost ¢/Gallon	45.5	80.4	21.3	56.1
Net Gasoline Cost $/MMBTU	3.79	6.70	1.78	4.68
Break Even Cost of Oil for Gasoline, $/Bbl	14.30	26.50	6.50-7.20	17.80-19.00

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1) CO top gas value at $1/MMBTU in 1975 and $1.80/MMBTU in 1985.
2) Based on incremental off-peak power cost (from peak load nuclear reactor operating as 1,000 MW(e) based load plant) of 0.3 mills/kwh(e) in 1975 and 6 mills/kwh(e) in 1985 (see Ref. 3). Power cost for air liquefaction was charged at a conventional rate of 30 mills/kwh(e) in 1975 and 50 mills/kwh(e) in 1985.
3) Fixed charge taken as 15% straight line on investment and 80% plant factor.
4) Oxygen by-product sold at $5/ton in 1975 escalated 80% to $9/ton.

Table III

PROCESS DATA FOR SYNTHETIC CARBONACEOUS FUELS USING A CO_2 FEEDSTREAM AND NUCLEAR POWER

Item	Consumed per Bbl Methanol(1	Consumed per Bbl Gasoline(2
CO_2, Tons	0.19	0.40
H_2, Lbs	51.9	108.1
Net H_2O, Bbls	0.89	0.93
Electrical Power for H_2 kwh(e)(3	970	2,020
Electrical Power for CO_2 Separation in Atmosphere Process, kwh(e)	100	200
Energy Conversion Efficiency, % (Electricity to Fuel)	75.3	65.8

1) Methanol - Density = 6.6 lbs/gal; sp. gr. = 0.79, HHV = 10,000 BTU/Lb = 66,000 BTU/Gal.

2) Gasoline - Density = 6.0 lbs/gal; sp. gr. = 0.72, (C_8H_{18}, Octane), HHV = 20,000 BTU/Lb = 120,000 BTU/Gal. 1 Bbl CH_3OH produces 0.48 Bbl (CH_2).

3) High pressure cells (30 atm Lurgi), 80% efficiency requires 18.7 kwh(e)/lb H_2.

Table IV

CAPITAL INVESTMENT COST DATA FOR ECONOMIC ANALYSIS OF SYNTHETIC CARBONACEOUS FUELS USING CO_2 FEEDSTREAM AND NUCLEAR POWER

1985

	Limestone Calcination	Fossil Fuel Power Plant	Atmosphere CO_2
CO_2 Conc. in Feedstream - % by Volume	∿ 100%	9-15%	0.035%
Capital Investment			
CO_2 Recovery Unit	---(1	\$200/KW(e)(2	3,750 Bbl/Day(3
Electrolysis Cells	\$200/KW(e)(4	\$200/KW(e)(4	\$200/KW(e)(4
Methanol Convertor	\$4,500/Bbl/Day(3	\$4,500/Bbl/Day(3	\$4,500/Bbl/Day(3
Gasoline Convertor	\$4,500/Bbl/Day(3	\$4,500/Bbl/Day(3	\$4,500/Bbl/Day(3

1) Assumes heat recovery benefit from condenser accrues to limestone calcination process which is equipped to separate a relatively pure CO_2 stream made available to SCFF at no charge.

2) CO_2 recovery assumes installation cost equivalent to stack gas SO_2 scrubbing, in terms of dollars per KW of power plant escalated 80% to 1985.

3) Based on Refs. 2, 3, and 9.

4) Based on recent estimates in Ref. 12 which is already assumed escalated because of anticipated improvements in cell design and construction by 1985.

Table V

PRODUCTION COST OF METHANOL FROM VARIOUS CO_2 FEEDSTREAMS AND NUCLEAR POWER

Basis: 1985 Escalated Cost in \$/Bbl
1000 MW(e) Nuclear Power Plant
725 MW(e) Available Off-Peak Power

Source of CO_2	Limestone Calcination	Fossil Fuel Power Plant	Atmospheric CO_2
Source Plant Capacity	$1.2\text{x}10^6$ T/Yr Lime Plant	800 MW(e) Coal Power Plant	$80\text{x}10^6$ SCFM Air Plant
CO_2 Conc. in Feedstock - Vol. %	100%	8-15%	0.035%
Production Capacity - Bbl/Day	16,300	16,300	16,300
Production Cost - \$/Bbl			
Depreciation on Capital Investment - 15%/Yr			
CO_2	--	0.72	1.93
Electrolysis	4.15	4.15	4.15
Methanol Convertor	2.31	2.31	2.31
Subtotal - Depreciation	6.46	7.18	8.39
Labor and Maintenance	0.36	0.54	0.81
Electrical Energy			
@ 30 mills/kwh(e) (1	29.10	30.00	32.00
@ 6 mills/kwh(e)	3.82	6.00	6.40
Total Production Cost for Methanol			
30 mills/kwh(e) - \$/Bbl	35.92	37.72	41.20
6 mills/kwh(e) - \$/Bbl	12.64	13.72	15.60
" " - ¢/Gallon	30.1	32.7	37.1
" " - \$/MMBTU	4.56	4.95	5.62

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1) Power and production cost for a dedicated plant would produce \$22,500 Bbl/Day methanol.

Table VI

PRODUCTION COST OF GASOLINE FUEL FROM VARIOUS CO_2 FEEDSTREAMS AND NUCLEAR POWER

Basis: 1985 Escalated Cost in $/Bbl
1000 MW(e) Nuclear Power Plant
725 MW(e) Available Off-Peak Power

CO_2 Feedstock Source	Limestone Calcination	Fossil Fuel Power Plant	Atmospheric CO_2
Source Plant Capacity	1.2×10^6 T/Yr Lime Plant	800 MW(e) Coal Power Plant	8.0×10^6 SCFM Air Plant
CO_2 Conc. in Feedstream - % by Vol.	100%	8-15%	0.035%
Gasoline Production Capacity - Bbl/Day	8,000	8,000	8,000
Production Cost - $/Bbl			
Depreciation on Capital Investment - 15%/Yr			
CO_2 Recovery	--	1.50	4.02
Electrolytic H_2	8.65	8.65	8.65
Methanol Convertor	4.81	4.81	4.81
Gasoline Convertor	2.31	2.31	2.31
Subtotal - Depreciation	15.77	17.27	19.79
Labor and Maintenance	0.75	1.13	1.69
Electrical Energy			
@ 30 mills/kwh(e)(1	60.60	62.50	66.66
@ 6 mills/kwh(e) (Off-Peak)	12.12	12.50	13.33
Total Production Cost			
@ 30 mills/kwh(e) (Dedicated)(1	77.12	80.90	88.14
@ 6 mills/kwh(e) (Off-Peak)	28.64	30.90	34.81
@ 6 mills/kwh(e) ¢/Gallon	68.2	73.6	82.9
Break even crude oil price for gasoline, $/Bbl	$21.00	$24.00	$27.50

1) Power cost and production cost for a dedicated plant would produce 11,000 Bbl/day gasoline.

Table VII

INDUSTRY RESOURCE CAPACITY TO PRODUCE SYNTHETIC CARBONACEOUS FUELS AND FEEDSTOCKS FROM EFFLUENT CO AND CO_2 FEEDSTREAMS AND NUCLEAR POWER

Based on 1975 Production Capacities

Industry	Production capacity Millions tons/yr	CO_2 or CO availability Millions tons/yr	Gasoline production capability Billions gal/yr	No. of 1000 MW(e) nuclear power plants required
Lime and Cement	25 Lime 85 Cement	86 CO_2	9	90
Steel	125 Steel	125 CO	20	125
Coal for Power	500 Coal	1,560 CO_2	165	1,680
Gasoline, US Consumption		(945 CO_2)(1	100	(1,000)(1

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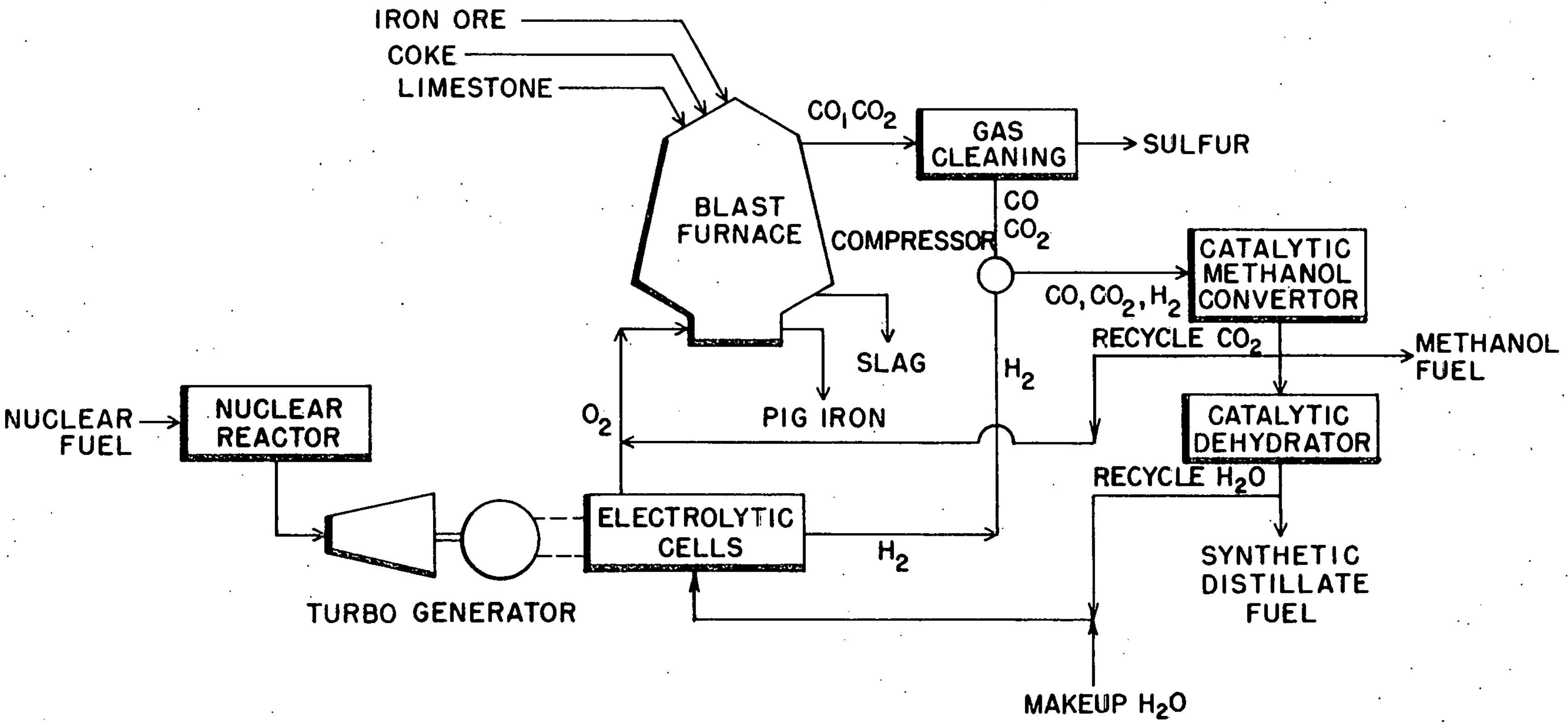
1) Capacity requirement to meet 1975 US consumption of gasoline of approximately 100 billion gal/yr.

Figure 1

SYNTHETIC CARBONACEOUS FUELS & FEEDSTOCK FROM THE STEEL INDUSTRY



REACTIONS:

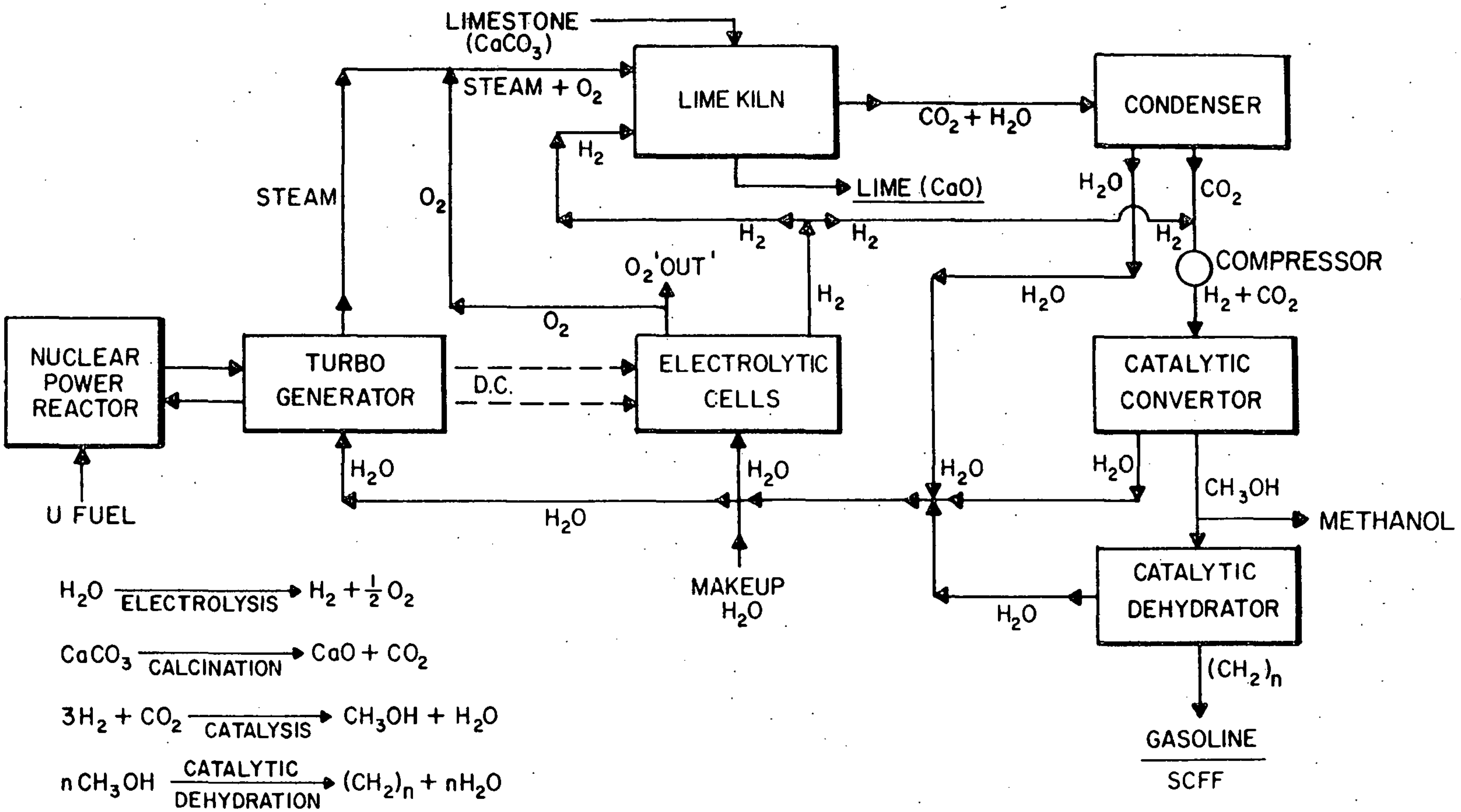
$$2H_2O = 2H_2 + O_2$$

$$CO + 2H_2 = CH_3OH$$

$$nCH_3OH = (CH_2)_n + nH_2O$$

Figure 2

PRODUCTION OF LIME AND BY-PRODUCT SYNTHETIC CARBONACEOUS FUELS & FEEDSTOCKS BY STEAM CALCINATION OF LIMESTONE USING NUCLEAR POWER



$H_2O \xrightarrow[\text{ELECTROLYSIS}]{} H_2 + \frac{1}{2} O_2$

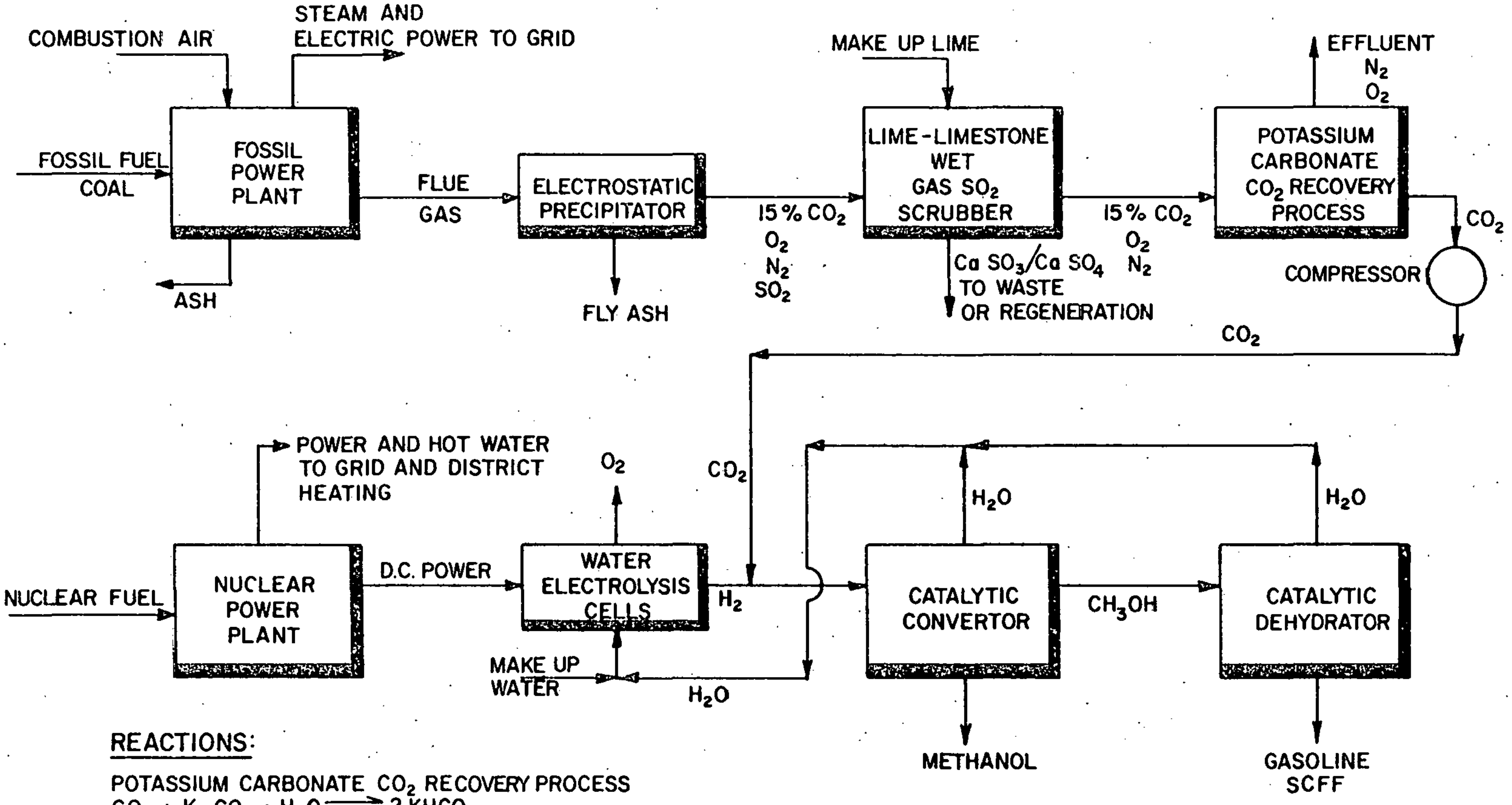
$CaCO_3 \xrightarrow[\text{CALCINATION}]{} CaO + CO_2$

$3H_2 + CO_2 \xrightarrow[\text{CATALYSIS}]{} CH_3OH + H_2O$

$n\,CH_3OH \xrightarrow[\text{DEHYDRATION}]{\text{CATALYTIC}} (CH_2)_n + nH_2O$

Figure 3

CONVERSION OF STACK GAS CO_2 FROM FOSSIL FUEL POWER PLANTS USING NUCLEAR POWER FOR THIS PRODUCTION OF SYNTHETIC CARBONACEOUS FUELS



REACTIONS:

POTASSIUM CARBONATE CO_2 RECOVERY PROCESS
$CO_2 + K_2CO_3 + H_2O \rightleftharpoons 2\,KHCO_3$

ELECTROLYSIS
$3H_2O = 3H_2 + {}^3\!/_2\,O_2$

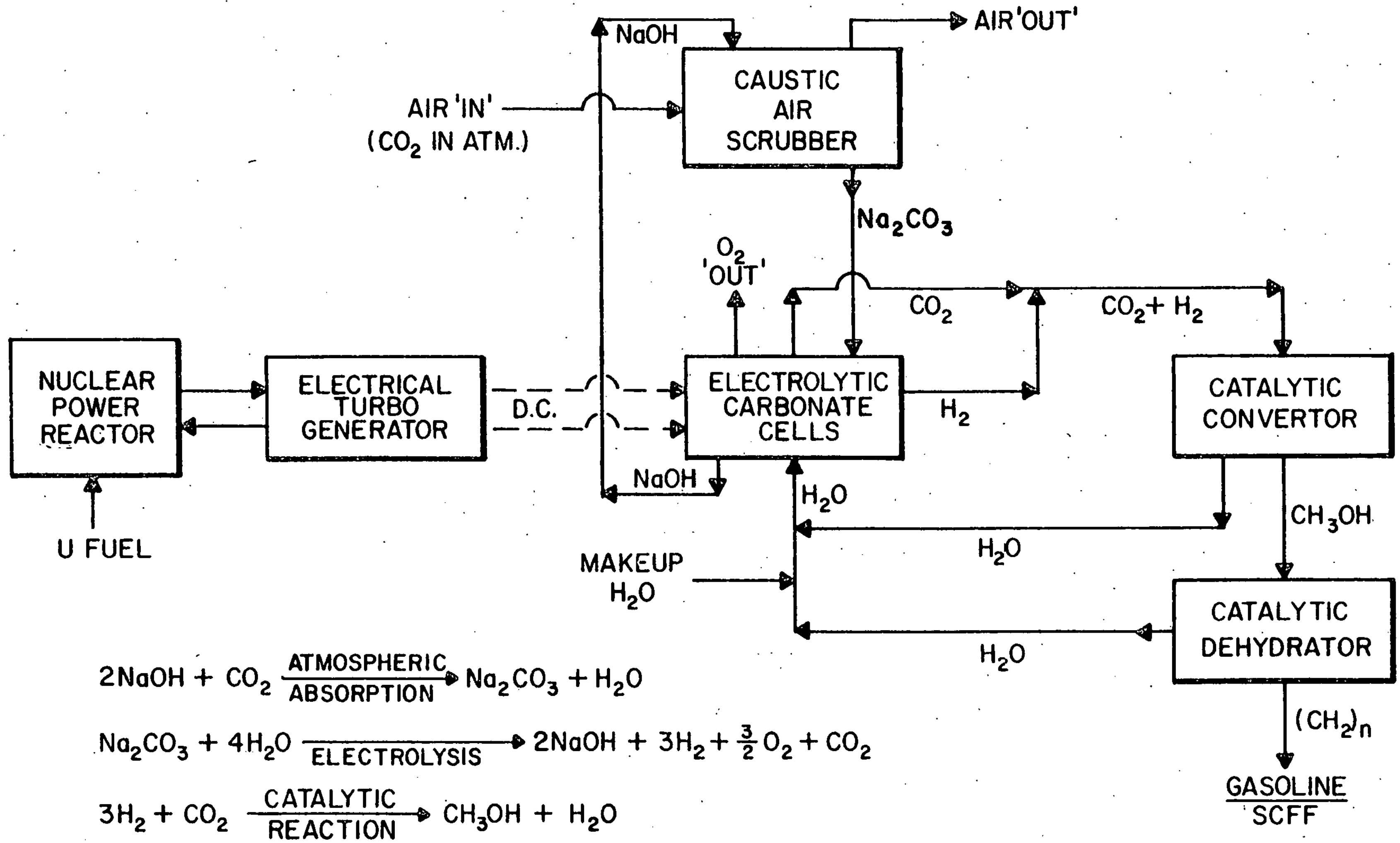
CATALYTIC CONVERTOR
$3H_2 + CO_2 = CH_3OH + H_2O$

CATALYTIC DEHYDRATION
$n\,CH_3OH = (CH_2)_n + n\,H_2O$

Figure 4

SYNTHETIC CARBONACEOUS FUELS AND FEEDSTOCKS (SCFF) FROM NUCLEAR POWER, AIR AND WATER

FIXATION OF ATMOSPHERIC CARBON



$$2NaOH + CO_2 \xrightarrow[\text{ABSORPTION}]{\text{ATMOSPHERIC}} Na_2CO_3 + H_2O$$

$$Na_2CO_3 + 4H_2O \xrightarrow[\text{ELECTROLYSIS}]{} 2NaOH + 3H_2 + \tfrac{3}{2}O_2 + CO_2$$

$$3H_2 + CO_2 \xrightarrow[\text{REACTION}]{\text{CATALYTIC}} CH_3OH + H_2O$$

$$n\,CH_3OH \xrightarrow[\text{DEHYDRATION}]{\text{CATALYTIC}} (CH_2)_n + n\,H_2O$$