USE OF CONTROLLED THERMONUCLEAR
REACTOR FUSION POWER FOR THE PRODUCTION
OF SYNTHETIC METHANOL FUEL FROM AIR
AND WATER

by

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Abstract

Methanol synthesis from carbon dioxide, water and nuclear fusion energy is extensively investigated in the present study. The entire system is analyzed from the point of view of process design and economic evaluation of various processes. The main potential advantage of a fusion reactor (CTR) for this purpose is that it provides a large source of low cost environmentally acceptable electric power based on an abundant fuel source. Carbon dioxide is obtained by extraction from the atmosphere or from sea water. Hydrogen is obtained by electrolysis of water. Methanol is synthesized by the catalytic reaction of carbon dioxide and hydrogen. The water electrolysis and methanol synthesis units are considered to be technically and commercially available. The benefit of using air or sea water as a source of carbon dioxide is to provide an essentially unlimited renewable and environmentally acceptable source of hydrocarbon fuel. Extraction of carbon dioxide from the atmosphere also allows a high degree of freedom in plant siting.

The significant contribution of the present study is the evaluation of various methods of separation of carbon dioxide from air or sea water. Eight different methods of extraction of carbon dioxide from air are analyzed. They are (1) absorption and stripping of air by water at atmospheric pressure, (2) absorption and stripping of air by water at atmospheric pressure

with a cooling tower as part of the absorption unit, (3) absorption and stripping of air by water at higher pressure, 20 atm, (4) absorption and stripping of air by methanol at 20 atm and -80°F , (5) removal of water vapor by adsorption on molecular sieves and subsequent extraction of carbon dioxide by refrigeration, (6) removal of water vapor by compression refrigeration and subsequent extraction of carbon dioxide by refrigeration, (7) absorption and stripping of air by a dilute aqueous potassium carbonate solution, and (8) removal of water vapor by adsorption on molecular sieves and adsorption/desorption of carbon dioxide from dry air by molecular sieves. A method of stripping of carbon dioxide from sea water is also presented. In order to compare these newly developed methods for CO_2 separation with other conventional non-fossil sources of carbon dioxide, the calcination of limestone is also examined.

It is found that the most economical process of obtaining carbon dioxide is from stripping of sea water. For the extraction of carbon dioxide from air, the process of absorption/stripping of air by dilute potassium carbonate solution is found to be the most economical. The total energy required for methanol synthesis from these sources of carbon dioxide is 3.90 kwh(e)/lb methanol of which 90% is used for generation of hydrogen. The process which consumes the greatest amount of energy is the absorption/stripping of air by water at high pressure and amounts to 13.2 kwh(e)/lb methanol.

Using nuclear fusion power plants of 1000 MW(e) to 9000 MW(e), it is found that the cost of methanol using the extraction of carbon dioxide from air with dilute potassium carbonate solution is estimated to be in the range between \$1.73 and \$2.90/MMBTU (energy equivalent) for plant capacities of 21,400 to 193,000 bbl/day methanol. This methanol cost is competitive with gasoline in the range of 19 ~ 33 cents/gallon. For the process of stripping of carbon dioxide from sea water, the cost is found to lie in the range of \$1.65 to \$2.71/MMBTU (energy equivalent) for plant capacities of 21,700 to 195,000 bbl/day methanol which is competitive with gasoline in the range of 18 ~ 30 cents/gallon.

Projection of methanol demand in the year 2020 is presented based on both its conventional use as chemicals and as a liquid fuel substituting for oil and gas.

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I. Introduction

After 20 years of research specifically aimed at achieving controlled fusion, coupled with concomitant technological developments in that field, in conventional nuclear reactors and in space science; it is believed that fusion power will be in practical use in the near future.^(1,2) With the continued development of plasma confinement, the remaining significant problem of plasma behavior in reactor regimes of temperature and density will soon be solved. The scientific, technical and size limitations of present-day fusion experiments preclude any of them from simultaneously achieving all three of the plasma parameters (temperature, density and confinement time) required for a fusion reactor with large and more complex facilities.

A controlled thermonuclear reactor can produce several forms of energy of use to the chemical industry. These include thermal, electrical, ultra-violet and high energy radiation. Thermal energy can be used for NO formation from air, partial oxidation of methane to acetylene, decomposition of water, coal

gasification, direct reduction of iron ores, ammonia synthesis, manufacture of titanium dioxide, etc. Electrical energy can be used for hydrogen and synthetic fuels, oxygen separation from air, metal refining, the chlor-alkali industry acetylene manufacture via arc process and ozone preparation. Radiation energy can be used for nitrogen fixation, carbon dioxide decomposition, ozone synthesis and water decomposition. Detailed discussions of the application of controlled thermonuclear reactor to chemical technology can be found in references 3 and 4.

There are several advantages of using fusion energy as a future energy source.⁽¹⁾ It is virtually an inexhaustible energy source because the fuels are of negligible cost (compared to fossil fuels) and are universally available. It does not generate a large environmental impact and it is silent. These allow it to be built in urban areas where fossil fuel stations would be unacceptable. Compared to fission and its hazards-radioactive fission products and critical safety requirements, fusion is inherently much safer. While conventional (non-breeding) nuclear power will no doubt have an important role in electric power generation, it is only used as an interim source, in the same sense that oil and eventually coal are necessarily interim sources. The main competitive power source to controlled thermonuclear reactors is the nuclear fission breeder reactor.

Liquid hydrocarbon fuels have become an important part of the energy economy. These fuels have the advantages of being readily stored, transported and utilized in engines. The ability of utilizing nuclear power to form synthetic liquid hydrocarbon fuels from non-fossil fuel sources would provide distinct environmental advantages and add a new dimension to nuclear power. The synthesis of methanol using only CTR energy, atmosphere and water comes closest to simulating the solar energy photosynthetic fuel cycle. The carbon fixed from the atmosphere to methanol is returned to the atmosphere by combustion. Due to the low intensity of solar energy, the photosynthesis process is inefficient, time consuming and sometimes unreliable. The CTR can provide large quantities of environmentally acceptable power for the fixations of carbon as methanol.

Methanol was commercially available from wood distillation before 1923.⁽⁵⁾ At present methanol is manufactured by synthesis from hydrogen and carbon monoxide containing small amounts of carbon dioxide at medium pressure or low pressure. It can also be manufactured from hydrogen and carbon dioxide.^(6,7) Some recent developments on methanol technology and economics are given in reference 8. Synthesis of methanol from carbon dioxide and hydrogen coupled with fusion energy from controlled thermonuclear reactors has some special compelling advantages.

With sufficient energy supply from controlled thermonuclear reactors, one can obtain hydrogen from electrolysis of water and carbon dioxide by separation from air or sea water. The economic advantage of the separation of carbon dioxide from the air is that the concentration of carbon dioxide in the air is fairly constant⁽⁹⁾ and air is readily available at no cost all over the earth. The total quantity of carbon dioxide in the atmosphere to support a synthetic fuel industry is essentially an unlimited supply and when taken in conjunction with the combustion cycle is a renewable resource. The heat content of methanol is 6.00×10^4 Btu/gal which is about half the value of gasoline (1.12×10^5 Btu/gal) and no major changes are needed for the automobile engine if petroleum is replaced by methanol for liquid fuel.⁽⁴⁵⁾

The other advantages of using fusion energy to make methanol are: (1) eliminate the hazards of deep coal mining, (2) eliminate the ecological problems of strip mining, (3) eliminate the sulfur and trace element pollution problems due to burning fossil fuels, (4) a CO_2 balance can be kept in the atmosphere avoiding any possible greenhouse effect, (5) plants will no longer be restricted to sites close to large deposits of fossil fuel, and (6) eliminate transporting oil, off-shore drilling and refinery construction.

The objectives of the present paper are to study the technology, economics and energy requirements of methanol

synthesis from carbon dioxide, hydrogen and energy output from controlled thermonuclear reactor (CTR). With the high energy output from CTR and the known technology of electrolysis of water, one can readily obtain hydrogen from water. The process of methanol synthesis will be described in the next section. The major emphasis of the present paper is on various technologies of separation of carbon dioxide from the atmosphere and sea water and their economic aspects. Comparative study of separation of carbon dioxide from limestone is also investigated. Comparison of various methods are then obtained by their economic feasibility. A projection of the requirement of methanol in the year 2020 is also presented.

II. Overall Process of Methanol Synthesis

The thermodynamics of the reaction between carbon dioxide and hydrogen to form methanol can be found in the literature.⁽⁵⁾ In the catalytic converter, carbon dioxide is converted to methanol by two possible reactions. The first one is in reference 8, page 58.

with heat of reaction $\Delta H_{298}^{\circ} = -12,826 \text{ cal/g-mole}$ and equilibrium

constants $K_p^{\circ}_{298} = 2.10 \times 10^{-1}$, $K_p^{\circ}_{600} = 3.34 \times 10^{-6}$. (Subscript for K_p° indicates temperature in degree Kelvin.) The second reaction involves a two-step reaction

with heat of reaction $\Delta H_{298} = +9838$ cal/g-mole and equilibrium constants $K_p^{\circ}_{298} = 1 \times 10^{-5}$, $K_p^{\circ}_{600} = 3.64 \times 10^{-2}$ and

with heat of reaction $\Delta H_{298}^{\circ} = -21664$ cal/g-mole and $K_p^{\circ}_{298} = 3.44 \times 10^4$, $K_p^{\circ}_{600} = 1.21 \times 10^{-4}$.

Whether the actual reaction is one step or two steps, there is no sufficient experimental data to make a definite statement. The equilibrium constant K_p° for the overall reaction seems to decrease with temperature. On the other hand, pressure has a favorable effect on conversion. In either reaction schemes, the kinetics of methanol synthesis from a mixture of carbon monoxide and carbon dioxide is rather complicated.

The rate of reaction in methanol synthesis depends on many factors, among them temperature, pressure, and composition of the inlet gas mixture. The more important consideration is the catalytic heterogeneous reactions and the selectivity of catalysts for minimizing side reactions. The dominating steps are the adsorptions of carbon dioxide and hydrogen on the

catalytic surface and the desorption of methanol from the catalytic surface. Natta⁽¹⁰⁾ developed a formula for the reaction rate for methanol synthesis as follows.

$$r = \frac{f_{\text{CO}}^2 f_{\text{H}_2} - f_{\text{CH}_3\text{OH}} / K_p^0}{(A + B f_{\text{CO}} + C f_{\text{H}_2} + D f_{\text{CH}_3\text{OH}})} \quad (4)$$

where f is partial fugacity of real gas in reaction mixture, A, B, C and D are functions of the activity of the catalyst and vary with temperature. Formula (4) gives better agreement with experimental results, as well as results in plant operations, than other proposed formulas. The kinetic expression for methanol synthesis from CO_2 or CO feedstock is similar.⁽⁵⁾ Therefore, Eq. (4) can be used for methanol synthesis from carbon dioxide and hydrogen. With the thermodynamic equilibrium data and the rate of methanol formation determined, it is possible to design a full size methanol plant so far as material balance is concerned.

According to reaction (1) for methanol formation, -12,826 cal/g-mole of heat is evolved for the synthesis. With the heat capacity equation given by Kobe and Long,⁽¹¹⁾ one can obtain the heat of reaction for reaction (1) to be -10205 cal/gal-mole (exothermic) at 623°K. For better heat transfer in the catalysts, the quench type reactor of catalyst beds may be more suitable. A heat balance may be worked out with the knowledge of the thermodynamic values.

The key to methanol synthesis is the catalyst. Catalysts may be composed of zinc oxide plus chromium oxide, zinc oxide alone, and copper or copper oxide, or of all three elements, usually in the form of oxides. If methanol is formed by the reactions of carbon monoxide and hydrogen as proposed by the second kind of reaction above, catalysts for this specific kind of reaction should be used. The activity of straight zinc oxide-chromic oxide catalysts without other metal oxides show a minimum activity at a composition of about $\text{Zn}_{73} \text{Cr}_{27}$ at temperatures of 350°C and above or of about $\text{Zn}_{63} \text{Cr}_{37}$ at or below 350°C .⁽¹²⁾ The pressure used in these processes range between 50 and 500 atm. The recent low pressure synthesis process of Imperial Chemical Industries^(8, p.47) operating at 750 psi (50 atm) used copper oxide catalysts which were studied before in early 1930.⁽¹³⁾ The copper based catalysts promoted by zinc oxide and chromic oxide (for example, $\text{Cu}_{49} \text{Zn}_{43} \text{Cr}_8$, $\text{Cu}_{50} \text{Zn}_{40} \text{Bi}_8$, $\text{Cu}_{60} \text{Zn}_{36} \text{Cr}_9$, and $\text{Cu}_{82} \text{Zn}_{16} \text{Cr}_4$) have high selectivity and high initial activity. The only drawback is that the copper based catalyst is sensitive to temperatures above 300°C and very sensitive to sulfur, therefore, it should be operated at temperature below 300°C and the sulfur content of the inlet gas should be maintained much below the level used with the zinc oxide-chromic oxide catalysts. Most new capacity methanol synthesis plants being built are based on the low pressure ICI type process because of the ability to produce high purity synthesis gas.

The electrolytic process produces high-purity hydrogen and consists in passing direct current through an aqueous solution of alkali; decomposing the water according to the following equation:

The theoretical decomposition voltage for this electrolysis is 1.23 volts at room temperature; however, because of overvoltage of oxygen on the anode and also cell resistance itself, voltages of 2.0 to 2.25 volts are usually required. A typical commercial electrolyzer uses a 15% NaOH electrolyte solution, with an iron cathode and a nickel-plated-iron anode, has an asbestos diaphragm separating the electrode compartments, and operates at temperatures from 60 to 70°C. The gas is about 99.7% pure and is suitable even for hydrogenation for edible oils. There are two types of cells: the bipolar and unipolar types. The unipolar cells may be open or closed tanks.

Process technology of the extraction of carbon dioxide from the atmosphere is not available from the literatures and therefore emphasis on methods of carbon dioxide separation unit is analyzed and will be presented in more detail below.

III. Processes for Extraction and Recovery of CO₂ from Atmosphere and Sea Water

Concentration of carbon dioxide in the atmosphere is about 0.036% by volume. The sublimation temperature of carbon

dioxide is 194°K at atmospheric pressure and is the second compound with the highest boiling point in the air (water vapor has the highest boiling point). Since air is available at no cost everywhere on the earth, it is a very desirable raw material for the sources of carbon dioxide for methanol synthesis. Several methods can be used for separation of carbon dioxide from the atmosphere. These are: (1) absorption of carbon dioxide in water at atmospheric pressure and then stripped the carbon dioxide from the carbonated solution, (2) use of cooling tower which is necessary for removal of waste heat from the controlled thermonuclear reactor power plant as part of the absorption/stripping system like method 1 above, (3) absorption and stripping of carbon dioxide from the air by water at high pressure, (4) absorption and stripping of carbon dioxide from the air by methanol at high pressure and low temperature, (5) removal of water vapor from the air by absorption on molecular sieves and separation of carbon dioxide from the dry air by refrigeration, (6) removal of water vapor from the air by compression refrigeration and separation of carbon dioxide from dry air by refrigeration, (7) absorption and stripping of carbon dioxide from the air by K_2CO_3 solution, and (8) separation of water vapor and carbon dioxide from air by adsorption/desorption on molecular sieves. Another abundant source of CO_2 is from sea water so a method of separation of carbon dioxide from sea water

is analyzed. Descriptions and design calculations for each process are given below.

1. Absorption/Stripping of Carbon Dioxide from Atmosphere by Water at Atmospheric Pressure

Absorption of carbon dioxide in water is of some industrial importance, particularly in ammonia synthesis. The use of water as an absorbent offers several advantages as compared, for example, to monoethanolamine solutions: (1) inexpensive readily available solvent, (2) carbon dioxide can be stripped out from the solution with less energy, (3) simpler plant design, (4) no heat load, (5) water not reactive with COS, and other possible trace constituents.

Energy supply for the methanol synthesis process is from the fusion energy of controlled thermonuclear reactor. A conceptual flow design of the metal-steam power conversion system is shown in Fig. 1. (2,48) The Tokamak reactor is of toroidal shape. The first wall is cooled by liquid lithium flowing in the blanket segments, which run parallel to the toroidal magnetic field and have a radial thickness of 1 meter. Tritium is bred by neutron absorption in the lithium. A typical breeding ratio is 1.3, giving a doubling time of about a month. A set of tubes installed in the lithium blanket utilizes the heat generated in the blanket to boil metal which generates power in a metal vapor turbine. The turbine metal exhaust condenses in a metal-condenser-steam-boiler which generates steam for a conventional steam cycle.

Of the 1000 megawatts from the blanket, 564 MW of electrical power is produced at 56.4 percent efficiency, and the useful electrical power output is 518 MW.

A flow diagram of methanol synthesis from carbon dioxide, hydrogen and nuclear energy is shown in Fig. 2. Carbon dioxide is obtained from the absorption/stripping of air by water at atmospheric pressure. Hydrogen is generated by electrolysis of water. Both hydrogen generation and methanol synthesis have been discussed before. Design calculation of extraction of carbon dioxide from air by absorption and stripping towers is presented in more detail below.

It is found by preliminary calculations that the unit capital investment (\$/bbl/day) of the CO₂ separation unit decreases with the increase of CO₂ production rate. Therefore, one would like to build a large CO₂ separation unit so that the unit capital investment will decrease. However, engineering and structural difficulty will prevent the size of the tower over a certain limit. For example, the diameter of the absorption tower is estimated about 1275 ft in order to get 10⁵ lb/hr CO₂ produced. This dimension of the tower is about 3 times larger than the present conventional cooling towers. Based on this engineering and economic reasoning, 10⁴ lb/hr CO₂ is used as the design separation unit.

The following conditions are set for the design calculations:

(1) Absorption efficiency of carbon dioxide by water is 95%, water flow rate in the towers is 20% more than minimum liquid flow rate.

(2) Concentration of CO_2 in inlet air is 0.036% by volume and concentration of CO_2 in inlet water stream is practically zero.

(3) Tower packed with 1 1/2 inch, wet-packed intalox saddles and tower is operated at 60% of flooding velocity.

(4) Operating at atmospheric pressure and 15°C . Liquid vapor solubility equilibrium relationship at 15°C is $y = 1200 x$ where x , y are the mole fractions of carbon dioxide in the liquid and gas phases respectively.

The rate of air flow at both ends of the tower is the same, so that it is convenient to find the exit air composition by using molal ratio as follows: $Y_0 = 3.6 \times 10^{-4} / 0.99964 = 3.601 \times 10^{-4}$ mole CO_2 /mole CO_2 free air. At the dilute end of the tower, for 95 percent absorption, $Y_1 = 0.05 \times 3.601 \times 10^{-4} = 1.8 \times 10^{-5} \approx y$.

Since both the operating and equilibrium lines are almost straight, the minimum water rate corresponds to

equilibrium at the rich end of the tower is $(X_o)_{\max} = 3.6 \times 10^4 / 1200 = 0.3 \times 10^{-6}$ mole of CO_2 /mole of CO_2 free water. With these known mole fractions, the actual slope of the operating line is then 1.44×10^3 and the mole fraction of carbon dioxide in exit liquid stream is 2.5×10^{-7} . The water and air flow rates are 1.64×10^{10} lb/hr and 1.83×10^7 lb/hr respectively. The design calculations here take into account of dissolution of N_2 and O_2 in water. Therefore, the concentrations of nitrogen, oxygen and carbon dioxide in the liquid phase are 5.89×10^{-4} g-mole/l, 3.21×10^{-4} g-mole/l and 1.67×10^{-5} g-mole/l respectively so the concentration of carbon dioxide in water is higher than that in air relative to the corresponding concentrations of nitrogen and oxygen in water. By flashing out N_2 and O_2 from water before flowing into the stripper, one can get higher concentration of CO_2 from the stripper. The number of transfer unit can be calculated from⁽¹⁵⁾

$$N_{OL} = \frac{X_o - X_1}{(X^* - X)_{\ln}} \quad (5)$$

where

$$(X^* - X)_{\ln} = \frac{(X^* - X)_o - (X^* - X)_1}{\ln [(X^* - X)_o / (X^* - X)_1]} \quad (6)$$

N_{OL} is the number of transfer unit in terms of logarithmic concentration driving force. From the above values of X_0 , X_1 , N_{OL} is determined to be 17.

The absorption of carbon dioxide in water has been shown to be almost entirely liquid-film controlled, presumably because of the relatively low solubility of carbon dioxide. The height of transfer unit in the tower cannot be determined before the dimensions of the tower are known.

The flooding velocity can be found by first calculating $\frac{L}{G} \sqrt{\frac{\rho_G}{\rho_L}} = 21$ where L is the water mass velocity in terms of lb/hr ft^2 , G is the air mass velocity, ρ_G and ρ_L are the densities of air and water respectively. With 1 1/2 inch intalox saddles for tower packing and 60% of flooding velocity, we then have (reference 15, Fig. 37-11)

$$\frac{G^2 \left(\frac{a_p}{\epsilon}\right) (\mu'_L)^{0.2}}{g_c \rho_G \rho_L} = 0.00054$$

where ϵ is percentage void of packing, $\frac{a_p}{\epsilon}$ is the surface to volume ratio, μ'_L is viscosity of liquid and g_c is gravitational constant. Substituting the characteristic values for packing and physical properties for water and air, we then get the air mass flow rate $G = 143 \text{ lb/hr ft}^2$ at 60% of flooding velocity. The diameter of the packed bed is

$$D = \sqrt{\frac{1.83 \times 10^7 \times 4}{143 \pi}} = 403 \text{ ft.}$$

This size is comparable to commercial dimension of cooling tower or large storage tank.

The height of transfer unit can be obtained from the empirical equation (reference 14, Eq. 6-2) and is $H_L = 2.83$ ft. Therefore, the total height of the tower is $N_{OL} H_L = 2.83 \times 17 = 48$ ft.

The pressure drop in the tower is obtained as 2.6 lb/in^2 .⁽¹⁶⁾ The energy required for the CO_2 separation unit is to move the liquid between the absorption and stripping towers. The operating conditions as far as thermal energy balance is concerned are as follows: both the absorption and stripping towers are operated isothermally with the absorption tower at lower temperature, the stripping tower at higher temperature. The sensible heat supplied to raise the temperature of the rich solution from the absorber is obtained through heat exchange from the lean solution from the stripper. The only external heat required for the system is the low pressure steam at 220°F extracted from the turbine of the power plant to desorb carbon dioxide from water. The heat required is 200 Btu/lb CO_2 and amounts to an equivalence of only $0.08 \text{ kwh(e)/lb CH}_3\text{OH}$. Other energy required is the mechanical energy to transport the air and liquid. The mechanical energy balance for liquid transport is

$$\frac{g}{g_c} \Delta Z + \eta W_s + \frac{2f u_L^2}{g_c D} = 0 \quad (7)$$

where the first term is due to potential energy for lifting water to the top of the absorber or stripper, the second term is due to energy supplied by the pump or blower and the third term is due to energy loss by friction in pipes. If we take into account the heads and skirts of the tower, the actual height of the tower would be 1.12 times the height of the packed bed and hence $\Delta Z = 53$ ft. Considering the liquid head for pumping, the suction static head of the pump is about the same as the discharge static head since the height of the absorption and stripping towers is the same. Hence, pumping power for lifting liquid is practically negligible. Efficiency of a pump or blower is usually 80%. In order to transport a large liquid flow rate 1.64×10^{10} lb/hr in a pipe and to avoid supersonic flow, we have to use a pipe with sufficiently large diameter. The friction factor for the carbon steel pipe is 0.004. The frictional loss due to water flow in a large pipe is 0.407 ft-lb_f/lb. To reduce pumping power, plant layout similar to the one employed by DuPont's Belle, West Virginia plant⁽²³⁾ is a good example to follow. The total energy supplied to the pump is $W_s = 0.407$ ft-lb_f/lb. Since 83 pumps are used to transport the fluid to a tower and 10^4 lb/hr of CO₂ is generated (equivalent to 0.727×10^4 lb/hr of methanol), the total energy required for extraction of carbon dioxide is 1.0 kwh(e)/lb CH₃OH. The minimum thermodynamic energy required for CO₂ separation from air is 0.0351 kwh(e)/lb CH₃OH

so the actual energy required by this process is about 28.5 times larger than the minimum values. The process is 3.5% efficient based on the minimum efficiency. This compares to a 10% thermodynamics efficiency for separation of oxygen from air by liquifaction.⁽³⁾ In the following treatment, the unit of energy required for CO₂ separation and methanol synthesis is in terms of kilowatt hour electricity per pound of methanol.

2. Extraction of Carbon Dioxide from Air by Absorption/Stripping with Cooling Tower as Part of the Absorption Unit at Atmospheric Pressure

The future electric utilities have to give more concern to the treatment of waste heat from their electric generating power plants because of the widespread adoption of Federal and state thermal pollution control regulations.⁽¹⁷⁾ Cooling towers were usually used in the past for heat removal and were the main reason for their construction.⁽¹⁸⁻²⁰⁾ A controlled thermonuclear reactor generates 1000 megawatts electricity and will dispose of about 3×10^9 Btu/hr of waste heat as can be seen in Fig. 1. A cooling tower is designed to remove this quantity of heat and is used to absorb part of the carbon dioxide in water. The schematic flow sheet of the methanol synthesis plant is shown in Fig. 3. Cold water of 1.73×10^8 lb/hr from river or pond at 80°F passes through a heat exchanger to remove 3×10^9 Btu/hr of waste heat from the controlled thermonuclear reactor. The exit temperature of the water after the heat exchanger is then raised to 97.3°F.

The absorption/stripping towers for carbon dioxide is the same as method (1) where carbon dioxide is absorbed in water in the absorption tower and is desorbed from the stripping tower.

In order to determine the amount of carbon dioxide absorbed in the cooling tower, one has to know the dimensions of the cooling towers. ⁽²¹⁾ With the given conditions of air and water flowing in the cooling tower, an induced-flow cooling tower of 22 feet tall is designed. Thirty decks of deck "A" packing are used with 9 inch spacing. Five cells of 6 feet modules with dimensions of 120 feet by 150 feet are used. Therefore, the mass velocities of air and water are 1.747×10^3 lb/hr ft² and 1.92×10^3 lb/hr ft² respectively. Total pressure drop in the cooling tower is 0.94 inch water and the brake horsepower required for each cell is 92.5 horsepower.

The water and air flow rates in the cooling tower are 1.73×10^8 lb/hr and 2.58×10^5 lb/hr respectively. Absorption of carbon dioxide in water can be determined by penetration theory ⁽²²⁾ which gives the mass transfer coefficient in liquid phase. The cross sectional area of the tower is 18×10^3 ft² (120 ft x 150 ft). Then the residence time of water in the tower is $t = 0.012$ hr. The mass transfer coefficient $k_L = \sqrt{\frac{D_{\text{CO}_2-\text{H}_2\text{O}}}{\pi t}}$

where $D_{\text{CO}_2-\text{H}_2\text{O}}$ is the molecular diffusivity of carbon dioxide in

water, so k_L can be calculated as 0.084 ft/hr. Now we take a thickness dx of the cooling tower and make a mass balance. The amount of carbon dioxide increased in the bulk liquid phase is equal to the amount of carbon dioxide transferred from the interface to the bulk liquid phase. Hence

$$k_L (\rho_S - \rho_B) (a_\rho dV) = \frac{1.73 \times 10^8}{62.4} d\rho_B \quad (8)$$

where ρ_S, ρ_B is the density of carbon dioxide at the interface and the bulk liquid phase, a_ρ is the surface to volume ratio of packing. Integrate Eq. (8) with the integration limits of density of carbon dioxide from 0 to ρ_B and the height of the cooling tower from 0 to 22 ft. We can then get the total amount of CO_2 dissolved in water which is 24 lb/hr and which corresponds to the mole fraction of carbon dioxide in the outlet liquid phase of the cooling tower as 0.561×10^{-7} .

13% of saturated value of CO_2 is absorbed in the cooling tower. To maintain this CO_2 concentration in the liquid before entering in the absorber, one has (1) to combine this carbonated water from about one hundred cooling towers or (2) to install a storage tank between the cooling tower and the absorber and operate the absorbers when sufficient carbonated water is held in the storage tank or (3) to combine (1) and (2) efficiently to reduce the number of cooling towers and size of storage tank.

Following the same procedure of method (1), one can get the outlet mole fraction of carbon dioxide in water as 0.24×10^{-6} . The diameter of the packed bed is 395 ft. The number of transfer unit and the height of each transfer unit are 11.67 and 2.62 ft respectively. Hence the total height of the packed bed is about 30 ft. Pressure drop in the tower is 1.3 psi. Considering the mechanical energy loss in the unit, it is found that the energy required per pound of methanol generated is 0.78 kwh(e).

3. Absorption/Stripping of Carbon Dioxide from the Air by Water at High Pressure

Absorption and stripping of carbon dioxide from the air by water at high pressure has the advantage of higher solubility of CO_2 in H_2O and hence increases the amount of CO_2 obtained in a shorter column of packed bed. This in turn will decrease the capital investment of the CO_2 -separation unit. The disadvantage is some extra energy required to compress the air. With a power recycle turbine, the energy needed to compress the air can be recovered to 80%. An unusual power-recovery system has been employed at DuPont's Belle, West Virginia Plant.⁽²³⁾

In the present system, 20 atmospheric pressure is employed in the absorber. A schematic flow sheet is shown in Fig. 4. Air of 1.8×10^7 lb/hr is compressed to 20 atmospheres before moving in the absorber. In the packed tower CO_2 is absorbed by counter current water flow of 1.06×10^9 lb/hr. The effluent carbonated water passes through a power-recovery turbine to reclaim some of the power available from the pressure-reduction of the liquid. CO_2 is stripped out from the stripping tower at a rate of 10^4 lb/hr. With this material and energy balances, methanol can be produced at 639 bbl/day. The diameter and height of the packed tower are 220 ft and 17 ft respectively. Wall thickness of the towers is about one foot (by stainless steel) in order to sustain 20 atm. Pressure drop in the

tower is 0.614 psi. Actual energy required to compress air from 1 atm to 20 atm can be recovered 80% by means of a turboexpander and the use of some steam for power generation from the nuclear reactor. The actual power output from the reactor is somewhat smaller than the nominal value because part of the steam has been used for generation of power for compression of inlet air. Compare the dimensions of the absorption tower with the one of operating at 1 atmospheric pressure in method (1). It is seen that the size of the tower is highly reduced. The total energy required to produce one pound of methanol is 9.66 kilowatt hour.

4. Absorption and Stripping of Carbon Dioxide from Air by Methanol at High Pressure and Low Temperature

Absorption and stripping of carbon dioxide by methanol was discussed by Herbert.⁽²⁴⁾ The advantage of this process is due to the high solubility of carbon dioxide in methanol at low temperature and elevated pressures and ready release from the solvent when the pressure is reduced. Solubility data of CO₂ in methanol and other solvents are given in Table 1. The heat requirements of the process are very low because the solution is cooled by pressure reduction in the regeneration step and the inlet gas is refrigerated by efficient heat exchange with the outgoing carbon dioxide free air. The disadvantages of the process are the complex flow scheme and the vaporization losses of the solvent.

A flow sheet of the process is given in Fig. 5. Inlet air enters the bottom of the absorption tower at elevated pressure, typically 20 atmospheres. It is precooled by the carbon dioxide free cold air stream from the absorber. Air is washed counter-currently with methanol which is fed into the absorber at approximately -80°F . At this operating condition, the mole fraction of methanol in the exit air stream is 4.2×10^{-6} . With an air flow rate of 1.83×10^7 lb/hr and the methanol flow rate of 9.51×10^6 lb/hr, 10^4 lb/hr of carbon dioxide can be generated. The rich solvent from the bottom of the absorber releases the pressure through a power recovery turbine and is stripped of the acid gas by heating with indirect steam in a conventional stripping column, cooled, and recycled into the top of the absorber. The diameter and packed height of the absorption tower are 119 ft and 7 ft respectively. Pressure drop through the packed bed is 0.25 psi. The total power required to compress the air and transport the liquid, etc., is 2.5×10^4 kilowatts. Therefore, energy required to synthesize one pound of methanol is 7.31 kilowatt hour.

5. Removal of Water Vapor from the Air by Adsorption on Molecular Sieves and Separation of Carbon Dioxide from the Dry Air by Refrigeration

Carbon dioxide and water vapor are the two components that have the highest normal boiling point in the air. Among

these two compounds, water will first condense upon cooling. In order to separate carbon dioxide from air, it is therefore necessary to remove water vapor first. Adsorption of water vapor and carbon dioxide on different molecular sieves successively to obtain CO_2 was discussed before.⁽²⁵⁾ The present process concentrates on the method of removing water vapor by molecular sieves and extraction of carbon dioxide from the dry air by refrigeration.

Molecular sieves have been widely used in industry for drying and gas separation. Adsorption of water vapor on molecular sieves have been discussed in the literatures.⁽²⁵⁻²⁷⁾ In the present process, air passes through a bed of molecular sieves (Linde 13X) at suitable pressure for water removal. After one bed is saturated, it can be regenerated and air is allowed to pass through an alternate bed. Dry air is then compressed and cooled to condense CO_2 . Several liquefaction cycles were discussed in the past and Barron⁽⁶⁾ gave a summary of 15 systems for air liquefaction. The precooled Linde-Hampson system and the Linde dual-pressure system have approximately the same performance in practice. Although the Claude system has better performance than the modifications of the basic Linde-Hampson system, it is more complex due to the addition of the expander and therefore increases the capital investment.

The process flow sheet for methanol synthesis using the above CO₂ separation method is given in Fig. 6. Air of 2.67×10^8 lb/hr at 1 atm and 80°F is compressed by a compressor to 200 psia before passing through the molecular sieve. A lower operating pressure may be used in the adsorption bed, but it will increase the volume of the bed and subsequently the capital investment. Assuming the incoming air is saturated with water vapor and the bed is large enough to remove all the water from the air, a water rate of 5.34×10^6 lb/hr can be removed. Part of this water, 1.79×10^5 lb/hr, is used for generation of hydrogen while the remaining part 5.16×10^6 lb/hr is used as coolant in the heat exchanger of the nuclear power plant system. Dry air is then compressed to 2500 psig and cooled to 77°F. The most economical way of cooling the dry air would be to pass it countercurrently with the unliquefied part of the air from the expansion process in a heat exchanger. After the air passes through the expansion valve, part of the CO₂ is liquefied and it is possible to recover it for methanol synthesis. With these operating conditions 9200 bbl/day of methanol can be manufactured.

The energy required from CO₂ separation is mainly from that of compression of air from atmospheric pressure to 200 psia before going into the adsorption bed of H₂O removal and that of refrigeration of carbon dioxide. The compressor

generally operates adiabatically and the energy required to compress air from 1 atm to 200 psia is 4.44 kwh/lb CO₂. Various cryogenic conditions can lead to the refrigeration of carbon dioxide from air. For example, at 41 atmospheric gauge pressure and -238°F, carbon dioxide can condense from the air.⁽⁴⁹⁾ In order to condense carbon dioxide from the air by these operating conditions, one can use compression refrigeration cycles with Freon-22 and methane as refrigerants. Freon-22 can be used as the first refrigerant to cool air from 80°F to -145°F and methane can be used as the second refrigerant to cool air from -145°F to -238°F. The latent heat of evaporation of methane can be supplied by the latent heat of condensation of Freon-22. However, this process shows an energy requirement for separation of carbon dioxide from air to be 17.84 kwh(e)/lb CO₂ which is not an economical method.

In the present study, practical plant operating conditions are used.⁽²⁹⁾ Air is compressed to 2500 psig, cooled to -40°F and then expanded to -270°C through an expansion engine or valve. Compressed air can be precooled through a heat exchanger by the carbon dioxide free dry air before it passes through the expansion valve. Part of the compression power can be recovered by the expansion engine or a turboexpander. So the energy required for the separation of carbon dioxide from dry air is mainly the

compression power of air from 200 psia to 2500 psig. With a two stage compressor of maximum 90% efficiency, the energy required to compress 2.67×10^8 lb/hr of air is 0.81 kwh/lb CO_2 .

Energy required to compress air from 1 atm to 200 psia is 4.44 kwh/lb of CO_2 . Therefore, the total energy needed to produce one pound of methanol is 7.21 kwh.

6. Removal of Water Vapor and Separation of Carbon Dioxide from Dry Air by Refrigeration

Carbon dioxide can be separated from air by refrigeration without going through a distillation process as is done in the separation of air. (29,30) The present process described here makes use of refrigeration to remove water vapor from the air first and then extract carbon dioxide from the dry air by refrigeration.

A schematic flow sheet of methanol synthesis by this method is shown in Fig. 7. The CO_2 extraction unit of refrigeration of dry air is similar to the one of method (5). Water is removed from air by passing through the refrigeration section of the compression refrigeration cycle. Air of 1.34×10^9 lb/hr at 1 atm and 80°F is cooled to -40°F after coming out of the refrigeration section. Under these conditions, 99.65% of the saturated water vapor is removed. (31) In order to remove this quantity of water, 2.68×10^7 lb/hr, 3.33×10^5 tons of refrigerant is needed.

Freon vapor leaving the refrigerator is compressed by the compressor and is condensed in the condensation unit. Heat of condensation in the condenser can be removed by the cold dry air from the water separator. High pressure Freon liquid expands in an expansion engine to Freon vapor which then passes through the refrigerator to remove the heat of condensation of water vapor from the air. Freon circulation rate in the cycle is 6.87×10^7 lb/hr. Condensed water is separated from the dry air in the separator. Part of this water can be used for production of hydrogen by means of electrolysis. Energy required for production of one pound of carbon dioxide by the water removal unit is 0.65 kwh. Combining with 6 kwh/lb CO_2 by the carbon dioxide extraction unit calculated in method (5), the total energy required for production of one pound of methanol is 9.14 kilowatt hours.

7. Absorption/Stripping of Carbon Dioxide from the Air by K_2CO_3 Solution at Atmospheric Pressure

Potassium carbonate solution has been studied and used for absorption and stripping of carbon dioxide from flue gas. Because of its slightly basic nature ($\text{pH} > 7$), it serves as a better solvent than water for absorption of carbon dioxide. Extensive data were reported for absorption and stripping of carbon dioxide by hot potassium carbonate solution at high pressure. (14, p.137-151) However, in the present investigation, it is desired to study the

absorption and stripping of carbon dioxide by dilute potassium carbonate solution at room temperature and atmospheric pressure. Solubility data of carbon dioxide in potassium carbonate solution is obtained from Perry^(16, p.11-14, 51) and is listed in Table 1. Design data is based on 0.01 N K_2CO_3 solution because the ratios of solubilities of nitrogen to oxygen to carbon dioxide in the solution is 0.13 to 0.036 to 1.00 at 77°F and 1 atmospheric total pressure so more than 86% purity of carbon dioxide can be recovered. Higher concentration of potassium carbonate solution will increase the purity of carbon dioxide from the stripper, however the production cost of methanol also increases because a greater loss of potassium carbonate is encountered.

The schematic flow sheet of this process is shown in Figure 8. Since more CO_2 dissolved in the solution, the liquid flow rate becomes 2.62×10^7 lb/hr which is much less than the water rate of 1.64×10^{10} lb/hr in method 1 while air flow rate is the same as 1.83×10^7 lb/hr. The diameter and height of the towers are 122 ft and 21 ft respectively. To save plant area one can build the absorbers and strippers on top of each other. Total energy required for moving the liquid and air is about 0.4 kwh(e)/lb CH_3OH . The potassium carbonate solution can also be used in conjunction with the power plant cooling towers. Carbon dioxide can be desorbed by hydrogen and low pressure steam which supplies the energy for desorption of CO_2 about 260 Btu/lb.

In operation of the stripper, carbon dioxide is stripped off from the solution by steam and hydrogen in the bottom of the stripper. The outlet stream from the top of the stripper passes through a condenser to condense most of the steam. The remaining steam, carbon dioxide and hydrogen then pass through a cooler so that these gases are cooled to 72°F. Heat released by the hot gases is absorbed countercurrently by aqueous carbonate solution from the absorber. With this design, one can obtain concentrated carbon dioxide and hydrogen mixture for catalytic conversion to methanol and reduce the steam supplied to the stripper.

8. Removal of Water Vapor from Air by Adsorption on Molecular Sieves and Separation of Carbon Dioxide from Dry Air by Adsorption/Desorption on Molecular Sieves

In addition to their use for dehydration as described in process 5 above, molecular sieves can also be used in an adsorption/desorption processes. Most of these take advantage of the fact that the molecular sieves show a high adsorptive selectivity for polar compounds such as CO₂. Carbon dioxide is very strongly adsorbed and can be readily separated from such non-polar systems as oxygen and nitrogen. In this process, water vapor is first removed from the air by adsorption on molecular seive Linde 13X,

similar to process 5. To separate carbon dioxide from dry air, molecular sieve Linde 5A is used.

The process flow sheet for carbon dioxide separation and subsequent methanol synthesis is shown in Fig. 8. For comparison purposes, the operating conditions for this process are made similar to those of method 5. Water saturated air at a rate of 2.51×10^8 lb/hr at 80°F is compressed by a two stage compressor to 200 psia. Water vapor at a rate of 5.03×10^6 lb/hr is adsorbed on Linde 13X adsorption bed and carbon dioxide of 1.27×10^5 lb/hr is subsequently adsorbed on Linde 5A. There are two alternate beds for adsorption/desorption of water and carbon dioxide. To regenerate the bed of Linde 13X, one can use carbon dioxide free dry air from the CO_2 adsorption beds to desorb water. Carbon dioxide can be desorbed by heating the bed or by a clean process gas such as hydrogen from the electrolysis of water. The heat of adsorption of water and carbon dioxide on molecular sieves is 1800 Btu/lb⁽⁴¹⁾ and 403 Btu/lb⁽⁴⁷⁾ respectively which tends to increase the temperature of the exit air stream. The energy requirement for the carbon dioxide separation unit is found to be 6.1 kwh(e)/lb CH_3OH mainly due to compression with power recovery.

9. Sea Water as a Source of Carbon Dioxide

The ocean has an abundance of many elements.⁽⁴⁶⁾
Dissolved carbon dioxide, bicarbonate and carbonate can be used

as a source of carbon dioxide. Although the concentrations of carbon dioxide, bicarbonate and carbonate in sea water vary with pH, salinity, location, temperature, and depth of the ocean, it is estimated that the concentration of dissolved carbon dioxide is about 1.08×10^{-5} g-moles/l which is equivalent to 2.64×10^{-4} cc CO₂/cc liquid sea water at a chloride content of 19 gm/Kg sea water and pH = 8.3. The dissolved oxygen and nitrogen at these conditions and 25°C are 4.95×10^{-3} and 8.96×10^{-3} cc/cc sea water respectively.⁽⁵⁰⁾ Considering the CO₂ content in non-ion paired bicarbonates and carbonates, one can estimate the total concentration of carbon dioxide in sea water as 0.215×10^{-2} g-moles/l. From the two concentrations obtained above, 1.08×10^{-5} and 0.215×10^{-2} g-moles/l, one can easily see that the major sources of carbon dioxide in seawater for methanol synthesis comes from bicarbonates and carbonates. When carbon dioxide is stripped out from sea water at 100°C, the actual CO₂ recovered may be slightly less than the available CO₂ content (0.0526 cc CO₂/cc sea water) in sea water.

A schematic flow sheet of this process is shown in Fig. 9. An absorption tower is not needed as compared to previous methods. Gas-liquid equilibrium data is based on the total CO₂ content which includes dissolved CO₂, non-ion paired bicarbonate and carbonates.

Carbon dioxide can be stripped out from sea water with steam together with hydrogen as an inert carrier gas. By condensing steam from the outlet gas of the stripper, one can obtain a mixture of carbon dioxide and hydrogen in stoichiometric ratio for methanol synthesis. Design calculations show that about 10^4 lb/hr carbon dioxide can be stripped from 1.13×10^8 lb/hr sea water. Stripping tower dimension is smaller than that by method 1 from pure water. Energy required for CO_2 stripping is correspondingly smaller which is calculated to be 0.36 kwh(e)/lb CH_3OH . From the results of previous methods, one can see that the diameter of the stripper is much greater than their corresponding height. The plant may be sited either on shore or on artificial islands offshore.

A summary of dimensions and numbers of absorption and stripping towers required, number of pumps and/or compressors required for nine different CO_2 -separation processes is given in Table 2.

IV. Estimates of Capital Investment and Production Costs of Methanol Synthesis from CO_2 , H_2 and Nuclear Fusion Energy

The energy required for the production of one pound of carbon dioxide and consequently one pound of methanol presented

above is summarized in Table 3. The major energy requirements for methanol synthesis are from electrolysis of water and carbon dioxide separation. The energy required⁽³²⁾ for catalytic conversion from carbon dioxide and hydrogen to methanol is 0.013 kwh(e)/lb CH₃OH which is relatively small compared to the energy needed by the other two units. The energy required for electrolysis of water is 3.5 kwh/lb CH₃OH^(3,33) The production rates of methanol from three different energy outputs, 1000 MW(e), 3000 MW(e) and 9000 MW(e) of controlled thermonuclear reactors are also listed in Table 3 for the nine different methods of carbon dioxide separation.

The capital investment and production cost of methanol synthesis with the application of the energy outputs of controlled thermonuclear reactors for the nine different methods of carbon dioxide extraction are presented in Tables 4-12. The costs of power plants from fusion energy, electrolyzers and the methanol synthesis unit are the same for the nine processes because of their same nature in design and manufacture. The capital costs for power plant, electrolyzer and methanol synthesis are obtained from references 33, 34, and 35 respectively. The cost of the CO₂-separation unit varies with different methods and a description of the distribution of the cost items is given in the following.

Most of the cost evaluation of the equipment is obtained from reference 16 and references 36 thru 39. All the costs presented in this study are based on 1974 values. The absorbers, strippers, pumps and heat exchangers are the major equipment for methods 1-4 and 7-9. Power recovery turbine is added to methods 3-6. For example, the cost of the absorption tower is based on its weight. Carbon steel is used as the material of the towers for absorption/stripping at atmospheric pressure and stainless steel is used as the material of the towers for absorption/stripping at high pressure because of its high allowable tensile strength. The cost of the pumps and compressors was obtained by determining the capacity, power required and pressure developed. The cost of the heat exchanger is based on its heat transfer area. In scaling up to a large production plant, an exponent of 0.6 is used. (36,38,40) An example of the cost distribution of the capital investment for CO₂ separation by method 2 is given in Table 4. Major cost items in method 5 are the compressors for inlet humid air and dry air and the beds of molecular sieves. Because of large quantity of water to be removed, the weight and cost of molecular sieves are 3.56×10^7 lb and $\$3.56 \times 10^7$ for method 5. The major cost of method 6 is a refrigeration unit for water vapor and is about $\$3.7 \times 10^7$. (37) More than 58% of the total cost of capital investment for methanol manufacture is due to the nuclear power plant and electrolyzer

Distribution of the production cost is also presented in Tables 5-14. Raw materials are supplied at nil costs. Catalyst cost is reasonably taken to be 0.25 cents/gal. Capital charge which includes amortization, interest and depreciation, is taken to be 10% per annum of the total capital investment. The production cost ranges from 10.45 to 17.22 cents/gal for method 9 in Table 13 for three different production rates. The thermal values are about \$1.65 to \$2.71 per million Btu which are competitive with the cost of gasoline in the range of 18-30 cents/gal.

Fig. 11 shows the capital investment of methanol synthesis plant vs. production rate by different methods of extraction of carbon dioxide from atmosphere and sea water. Because of the high cost of compression power, one has to invest more in methods 3-6, 8 and 9. The use of cooling tower as part of the absorption unit decreases the dimensions of the absorption and stripping towers and hence reduces the capital investment. Stripping of sea water to obtain carbon dioxide is an advantageous method as a source for methanol synthesis. Since no absorption unit is required, the capital investment for CO₂-separation decreases more than fifty percent as compared to method 1 of separation from air. Tabulated unit capital investment of CO₂-separation unit for nine methods is also given in Table 2.

Fig. 12 is the production cost of methanol as a function of production rate. The production cost varies with different methods of carbon dioxide extraction due to the differences in capital investment and energy requirement for production. Costs of adsorbents and refrigerants are high and hence the production costs are 57.20 and 87.96 cents/gallon respectively for methods 5 and 6. On the other hand, absorption and stripping of carbon dioxide by water offers relatively lower cost at atmospheric pressure. In all cases, production costs decrease with the increase of production rates. The lowest attainable cost in the present study is 10.45 cents/gallon by stripping sea water to obtain carbon dioxide.

V. Alternate Sources of Carbon Dioxide for Methanol Synthesis

Coal is a conventional source for methanol synthesis. However, as mentioned previously due to the polluted oxidation products such as sulfur and nitrogen and the hazards in its production, it is an undesirable long term fuel. Also the cost of coal is increasing and it is a non-renewable resource. An alternate non-fossil fuel source for CO_2 is limestone. Limestone is available in most parts of the United States at a cost of about \$1.50/ton at the mine and can be transported away from the mine at a

cost of about \$3/ton. However, the economics of methanol synthesis will be examined here.

In the United States lime is manufactured at about 25 million tons each year generating some 20 million tons of easily recoverable carbon dioxide with concentration of 30 to 40 percent. Present technology of a lime plant requires an average of about 5 million Btu per ton of lime. Modern vertical kilns approach 3.5 million Btu per ton of lime and by new technology, i.e., catalytic calcination, the energy required might drop to about 3 million Btu per ton. Capital cost of a lime plant of 500,000 tons per year capacity yielding 400,000 tons of carbon dioxide per year is about 8 to 10 million dollars.

With this information for carbon dioxide production, one can calculate the capital investment and production cost of methanol synthesis by means of CTR power. Instead of using natural gas as a fuel for the kiln, one can substitute it directly with CTR thermal energy or hydrogen from water electrolysis. The results are shown in Table 14. Although the capital investment from the carbon dioxide generation unit is lower than most of the methods discussed above from atmosphere, the energy requirement for methanol synthesis is higher (4.44 kwh(e)/lb CH_3OH with electrolyte hydrogen as fuel than methods 7 and 9 discussed above. This results in a lower

production rate for methanol by this process as compared to the other processes discussed except 7 and 9. Adding the cost of limestone and transportation, one finds that the production cost of methanol for this process is not the lowest. Furthermore, ecological problems arise because of the required large mining operations. The production of methanol from sea water or even air, water and fusion energy appears environmentally superior and economically competitive.

VI. Projections of Demand of Methanol in the Year 2020

Methanol production in the United States in 1971 was estimated to be about $1,200 \times 10^6$ gal/year⁽⁶⁾ and is mainly for industrial use such as manufacture of formaldehyde, dimethyl toluene, acetic acid and possibly for fuel cell and water purification. The average annual growth rate of methanol manufacturing since World War II is about 13.6%.⁽⁴²⁾ An estimate projection of the production will be 620×10^9 gal/year in 2020 for industrial uses.

Another potential use of methanol is as high-octane additive for unleaded motor fuel. Methanol has a very high Research Octane Number (blending Value, 133) although its Motor Octane Number is relatively low (105). If methanol can be used as a gasoline substitute, it can also be used as a substitute for natural gas and oil for residential, commercial and transportation purposes such as space heating, air conditioning, cooking, rail

and subway, etc. In 1970, 16% of U. S. petroleum and natural gas needed was imported.⁽⁴⁴⁾ With energy independence aimed in the country, one has to develop other energy sources to compensate for this energy requirement.

Several energy systems have been proposed for U. S. energy demand and fuel mix in the year 2020.^(34,43) The only inputs of the reference system, assuming no fusion power input; are LWR LMFBR, hydro-power, coal, oil and gas. The energy input mix used in this reference energy system reflects, as far as possible, the expected growth or decline in each of the input technologies. The imported energy for crude oil and natural gas for this system is estimated to be 57.2×10^{15} Btu/yr and 9.56×10^{15} Btu/yr respectively. If methanol is used to substitute for these fuels, the production rate of methanol would be $1,043 \times 10^9$ gal/year. This quantity is based on the assumption that methanol is the only fuel substitute for oil and natural gas. Combining with industrial needs described above, the total amount of methanol manufactured will have to be $1,663 \times 10^9$ gal/year. If methanol is produced from sea water and nuclear fusion processes which requires about 3.9 kwh(e)/lb CH_3OH , the total energy needed to produce the quantity of methanol described above will be 6.0×10^6 MW(e).

VII. Conclusions

Several conclusions can be derived from the present study:

1. Fusion energy can be applied to the manufacture of methanol by providing electricity for electrolysis of water and for extraction of carbon dioxide from air or sea water. The CTR synthesis route for fixation of atmospheric carbon dioxide provides an ecologically acceptable and renewable source of liquid hydrocarbon fuel.

2. Design and cost estimates indicate feasibility of the system. Carbon dioxide can be extracted from air by absorption/stripping with water, methanol or potassium carbonate solution, adsorption on molecular sieves and/or refrigeration. Carbon dioxide can also be recovered from sea water. The energy required for carbon dioxide extraction from air ranges from a low value of 0.35 to a high value of 13.16 kwh(e)/lb methanol. The energy required for hydrogen generation is 3.50 kwh(e)/lb methanol. The minimum total energy requirement is therefore 3.86 kwh(e)/lb methanol. The lowest energy requirement is for the process of absorption/stripping using dilute K_2CO_3 and the highest is for the pressure (20 atm) absorption/stripping by water.

3. Because no absorption towers are required, it is found that separation of carbon dioxide from sea water provides the lowest unit capital investment of about \$1146/bbl/day methanol at a plant capacity of 21,700 bbl/day. The production cost of methanol synthesis lies between 10.5 and 17.2 cents/gallon or \$1.65 and

\$2.71/MMBtu equivalent to 18 to 30 cents/gallon of gasoline.

However, the plants must be sited in the vicinity of the oceans.

4. The extraction of carbon dioxide from the atmosphere allows a greater degree of freedom for siting the synthetic fuel plant. Production costs range from 11 to 82 cents/gallon among the eight methods of extraction of carbon dioxide from the atmosphere at production rate in the range of 6,340 to 193,000 bbl/day. The process of absorption/stripping of air by K_2CO_3 solution indicates the lowest production cost and lies between \$1.73 and \$2.89/MMBtu which is economically competitive with gasoline in the range of 19 to 33 cents/gallon.

5. Total capital investment is found to be in the range of \$453 millions to \$2950 millions for methanol production using air or sea water as the resource for carbon dioxide and a nuclear power plant capacity of 1000 MW(e) to 9000MW(e). The unit capital investment in the CO_2 separation plant ranges from \$1146 to \$40,000/bbl/day. The lowest unit capital investment is the process of extraction of carbon dioxide from sea water whereas the highest unit capital investment is the process of separation of water and carbon dioxide from air by refrigeration. Over 58% of the capital investment is attributed to the power plant and electrolyzer. Hence, the capital investment is strongly controlled by the nuclear power plant (CTR).

6. Limestone has also been investigated as a source for carbon dioxide. However, the material and transportation costs make the methanol produced more expensive than the corresponding methods from sea water or air.

7. Projections of U. S. methanol requirement in the year 2020 is in the order of 1,663 billions gallons per year requiring 6 million MW(e) of CTR power for industrial, residential, commercial and transportation use when oil and natural gas are no longer available.

8. The present technologies of methanol synthesis can be conjuncted with other kinds of nuclear fission reactors in addition to controlled thermonuclear fusion reactor equally well. With the existing technology of nuclear fission reactor, the economics of the present technique may eventually look promising.

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Table 1

SOLUBILITY DATA OF CARBON DIOXIDE IN VARIOUS SOLVENTS EXPRESSED IN DIFFERENT UNITS

System	Equilibrium Data	Henry's Law	Concentration Distribution	CO ₂ Solubility ¹
CO ₂ -H ₂ O (25°C)	$y = 1.64 \times 10^3 x$	$p = 29.5 C_l$	$C_g = 1.21 C_l$	2.98×10^{-4}
CO ₂ -CH ₃ OH (-62.2°C, 20 atm)	$y = 0.413x$	$p = 0.339 C_l$	$C_g = 0.018 C_l$	2.01×10^{-2}
CO ₂ - K ₂ CO ₃ (0.01N, 25°C)	$y = \frac{81.6 \times 10^{-6} x^2}{8.10 \times 10^{-9} - x^2}$	$p = \frac{81.6 \times 10^{-6} C_l^2}{25 \times 10^{-6} - C_l^2}$	$C_g = \frac{3.3 \times 10^{-6} C_l^2}{25 \times 10^{-6} - C_l^2}$	0.11
CO ₂ -Sea Water (19 g/l chlorine, 25°C) (dissolved CO ₂ only)	$y = 1.85 \times 10^3 x$	$p = 33.3 C_l$	$C_g = 1.36 C_l$	2.64×10^{-4}
CO ₂ - sea water ²				5.26×10^{-2}

¹Volume of non-ion paired CO₂ in solution in equilibrium with air at 1 atm and 25°C (STP).
cc CO₂ gas (STP)/cc liquid.

²Total CO₂ in sea water, including dissolved and non-ion paired carbonate and bicarbonate.

Note: C_g = concentration of CO₂ in gas phase in moles/l
 C_l = concentration of CO₂ in liquid phase in moles/l
 p = partial pressure of CO₂ in gas phase in atm.
 x = mole fraction of CO₂ in liquid phase
 y = mole fraction of CO₂ in gas phase

Table 2

UNIT CAPITAL INVESTMENT FOR DIFFERENT CO₂-SEPARATION PROCESSES

Summary of Dimensions and Number of Absorption and Stripping Towers Required
and
Number of Pumps and/or Compressors Required

No.	Process	Total CH ₃ OH Production (bbl/day)	Dimensions of absorption stripping towers (ft Diameter x ft Height)	Number of Towers (Absorp- tion and stripping)	Number of pumps and/ or compres- sors	Unit Capital Investment \$/bbl/day
1	H ₂ O absorption/ stripping (atm pressure)	18,600	403x48	60	540	4,282
2	H ₂ O absorption/ stripping (atm pressure) with cooling tower	19,700	395x30	60	540	2,920
3	H ₂ O absorption/ stripping (20 atm)	6,342	220x17	60	508	25,200
4	Methanol absorption/ stripping (20 atm, -80°F)	7,721	119x7	60	507	24,000
5	H ₂ O adsorption on molecular sieves, CO ₂ refrigeration	9,200	Volume of Molecular Sieve Beds 1.58x10 ⁶ ft ³	-	139	25,700
6	H ₂ O refrigera- tion, CO ₂ refrigeration	6,600	-	-	231	40,000

Table 2 (Continued)

No.	Process	Total CH ₃ OH Production (bbl/day)	Dimensions of absorption stripping towers (ft Diameter x ft Height)	Number of Towers (Absorp- tion and stripping)	Number of pumps and/ or compres- sors	Unit Capital Investment \$/bbl/day
7	Aqueous K ₂ CO ₃ absorption/ stripping (atm pressure)	21,400	122x21	60	60	2,082
8.	H ₂ O and CO ₂ adsorption/ desorption from molecu- lar sieves	8,700	Volume of Molecular Sieve Beds 1.75x10 ⁶ ft ³ for H ₂ O adsorption, 8.84x10 ⁴ ft ³ for CO ₂ adsorption/ desorption	-	95	14,460
9.	Stripping of sea water (1 atm)	21,700	140x22	30	30	1,146

Table 3

SUMMARY OF ENERGY REQUIREMENT AND PRODUCTION RATES FOR METHANOL SYNTHESIS
USING VARIOUS CO₂-SEPARATION METHODS

Energy Required for Electrolysis of Water = 3.50 kwh(e)/lb CH₃OH
(80% efficient cells)

No	Method of CO ₂ -Separation	Energy for	Total	Methanol capacity in			% of total energy required for H ₂ O electrolysis
		CO ₂ sept'n kwh(e)/lb CH ₃ OH	energy for CH ₃ OH synth. kwh(e)/lb CH ₃ OH	1000's bbl/day for CTR power levels	3000MW(e)	9000MW(e)	
1.	H ₂ O absorption/stripping (atm. pressure)	1.00	4.50	18.6	55.8	167.3	78
2.	H ₂ O absorption/stripping with cooling tower	0.78	4.28	19.7	59.1	177.0	82
3.	H ₂ O absorption/stripping (20 atm.)	9.66	13.16	6.3	19.0	57.0	27
4.	Methanol absorption/ stripping (20 atm.-80°F)	7.31	10.81	7.7	23.2	69.5	32
5.	H ₂ O adsorption on molecular sieves, CO ₂ refrigeration	7.21	10.71	9.2	27.6	82.7	33
6.	H ₂ O refrigeration, CO ₂ refrigeration	9.14	12.64	6.6	19.8	59.4	28
7.	Aqueous K ₂ CO ₃ absorption/ stripping (atm.pressure)	0.40	3.90	21.4	64.2	193.0	90
8.	H ₂ O and CO ₂ adsorption/ desorption from molecular sieves	6.10	9.60	8.7	26.0	78.0	34
9.	Stripping of sea water (atm.pressure)	0.36	3.86	21.7	65.0	195.0	91

Table 4

CTR-METHANOL SYNTHESIS
DISTRIBUTION OF CAPITAL INVESTMENT FOR THE CO₂ SEPARATION
FROM AIR UNIT

Process No. 2

Water Absorption/Stripping with Part
of CO₂ Absorbed in Cooling Towers

Production Rate = 10,000 lb/hr CO₂

<u>Process Equipment</u>	<u>\$10⁶</u>
Absorption and Stripping Towers	0.4
Pumps	0.2
Tower Packings	3.2
Heat Exchanger	3.3
Other (piping, etc.)	<u>0.1</u>
Total	7.2

A six-tenth factor is used to scale up to larger size plants.

Table 5

CAPITAL INVESTMENT AND PRODUCTION COST OF METHANOL SYNTHESIS
FROM AIR, WATER AND FUSION ENERGY

CO₂ is Obtained by Absorption/Stripping of Air by Water at Atmospheric Pressure

Process No. 1

<u>Capacity</u>		<u>Capital Investment</u>						
<u>MW(e)</u>	<u>bbl/day</u> <u>CH₃OH</u>	<u>Power Plant</u> <u>\$/kw</u>	<u>Electrolyzer</u> <u>\$/kw</u>	<u>Power Plant</u> <u>Electrolyzer</u> <u>(\$10⁶)</u>	<u>CO₂-</u> <u>separation</u> <u>\$/bbl/day</u>	<u>Methanol</u> <u>Synthesis</u> <u>\$/bbl/day</u>	<u>CO₂-CH₃OH</u> <u>total</u> <u>(\$10⁶)</u>	<u>Total</u> <u>capital</u> <u>(\$10⁶)</u>
1000	18,600	350	28.5	378	4,280	2,400	124	502
3000	55,800	285	22.9	924	2,760	1,558	241	1,165
9000	167,300	225	18.4	2,164	1,700	1,027	456	2,620

<u>Capacity</u>		<u>Production Cost ¢/gal.</u>					<u>Thermal value</u>
<u>MW(e)</u>	<u>bbl/day</u> <u>CH₃OH</u>	<u>Raw Material</u>	<u>Catalyst</u>	<u>Operating &</u> <u>Maintenance</u>	<u>Capital</u> <u>charge</u>	<u>Total</u>	<u>\$/MMBTU</u>
1000	18,600	0	0.25	0.53	21.42	22.2	3.50
3000	55,800	0	0.25	0.41	16.56	17.22	2.71
9000	167,300	0	0.25	0.31	12.43	13.00	2.09

Table 6

CAPITAL INVESTMENT AND PRODUCTION COST OF CH₃OH SYNTHESIS
FROM CO₂, H₂O AND NUCLEAR FUSION³

CO₂ is Obtained from Cooling Tower, Absorption/Stripping Towers

Process No. 2

Capacity		Capital Investment						
MW(e)	bbl/day CH ₃ OH	Power Plant \$/kw	Electrolyzer \$/kw	Power Plant Electrolyzer (\$10 ⁶)	CO ₂ - separation \$/bbl/day	Methanol Synthesis \$/bbl/day	CO ₂ -CH ₃ OH total (\$10 ⁶)	Total capital (\$10 ⁶)
1000	19,700	350	28.5	378	2,920	2,300	102	480
3000	59,100	285	22.9	924	1,877	1,570	203	1,127
9000	177,000	225	18.4	2,164	1,214	960	385	2,549

Capacity		Production Cost ¢/gal.						Thermal value
MW(e)	bbl/day CH ₃ OH	Raw Material	Catalyst	Operating & Maintenance	Capital charge	Total		\$/MMBTU
1000	19,700	0	0.25	0.48	19.33	20.06		3.16
3000	59,100	0	0.25	0.37	15.13	15.76		2.48
9000	177,000	0	0.25	0.29	11.43	11.96		1.89

Table 7

CAPITAL INVESTMENT AND PRODUCTION COST OF METHANOL SYNTHESIS
FROM AIR, WATER AND FUSION ENERGY

CO₂ is Obtained by Absorption/Stripping of Air by Water at High Pressure

Process No. 3

<u>Capacity</u>		<u>Capital Investment</u>						
<u>MW(e)</u>	<u>bbl/day</u> <u>CH₃OH</u>	<u>Power Plant</u> <u>\$/kw</u>	<u>Electrolyzer</u> <u>\$/kw</u>	<u>Power Plant</u> <u>Electrolyzer</u> <u>(\$10⁶)</u>	<u>CO₂-</u> <u>separation</u> <u>\$/bbl/day</u>	<u>Methanol</u> <u>Synthesis</u> <u>\$/bbl/day</u>	<u>CO₂-CH₃OH</u> <u>total</u> <u>(\$10⁶)</u>	<u>Total</u> <u>capital</u> <u>(\$10⁶)</u>
1000	6,342	350	28.5	378	25,200	3,725	183	561
3000	19,000	285	22.9	924	16,200	2,400	353	1,277
9000	57,000	225	18.4	2,164	10,500	1,547	687	2,851

<u>Capacity</u>		<u>Production Cost ¢/gal.</u>					<u>Thermal value</u> <u>\$/MMBTU</u>
<u>MW(e)</u>	<u>bbl/day</u> <u>CH₃OH</u>	<u>Raw Material</u>	<u>Catalyst</u>	<u>Operating &</u> <u>Maintenance</u>	<u>Capital</u> <u>charge</u>	<u>Total</u>	
1000	6,342	0	0.25	1.76	70.2	74.46	11.77
3000	19,000	0	0.25	1.33	53.34	54.92	8.68
9000	57,000	0	0.25	0.59	39.69	40.53	6.41

Table 8

CAPITAL INVESTMENT AND PRODUCTION COST OF METHANOL SYNTHESIS
FROM AIR, WATER AND NUCLEAR FUSION ENERGY

CO₂ is Obtained by Absorption/Stripping of Air by Methanol
at High Pressure and Low Temperature

Process No. 4

Capacity		Capital Investment						
MW(e)	bbl/day CH ₃ OH	Power Plant \$/kw	Electrolyzer \$/kw	Power Plant Electrolyzer (\$10 ⁶)	CO ₂ - separation \$/bbl/day	Methanol Synthesis \$/bbl/day	CO ₂ -CH ₃ OH total (\$10 ⁶)	Total capital (\$10 ⁶)
1000	7,721	350	28.5	378	24,000	3,443	212	590
3000	23,200	285	22.9	924	15,400	2,217	409	1,333
9000	69,500	225	18.4	2,164	9,910	1,430	788	2,952

Capacity		Production Cost ¢/gal.					Thermal value
MW(e)	bbl/day CH ₃ OH	Raw Material	Catalyst	Operating & Maintenance	Capital charge	Total	\$/MMBTU
1000	7,721	0	0.25	1.52	60.65	62.42	9.83
3000	23,200	0	0.25	1.14	45.60	46.99	7.41
9000	69,500	0	0.25	0.84	33.71	34.8	5.48

Table 9

CAPITAL INVESTMENT AND PRODUCTION COST OF METHANOL SYNTHESIS
FROM AIR, WATER AND NUCLEAR FUSION ENERGY

CO₂ is Obtained by Refrigeration of Dry Air
which is Removed by Adsorption on Molecular Sieves

Process No. 5

<u>Capacity</u>		<u>Capital Investment</u>						
<u>MW(e)</u>	<u>bbl/day</u> <u>CH₃OH</u>	<u>Power Plant</u> <u>\$/kw</u>	<u>Electrolyzer</u> <u>\$/kw</u>	<u>Power Plant</u> <u>Electrolyzer</u> <u>(\$10⁶)</u>	<u>CO₂-</u> <u>separation</u> <u>\$/bbl/day</u>	<u>Methanol</u> <u>Synthesis</u> <u>\$/bbl/day</u>	<u>CO₂-CH₃OH</u> <u>total</u> <u>(\$10⁶)</u>	<u>Total</u> <u>capital</u> <u>(\$10⁶)</u>
1000	9,200	350	28.5	378	25,700	3,210	266	644
3000	27,600	285	22.9	924	13,300	2,068	424	1,348
9000	82,700	225	18.4	2,164	6,880	1,334	679	2,843

<u>Capacity</u>		<u>Production Cost ¢/gal.</u>					<u>Thermal value</u>
<u>MW(e)</u>	<u>bbl/day</u> <u>CH₃OH</u>	<u>Raw Material</u>	<u>Catalyst</u>	<u>Operating &</u> <u>Maintenance</u>	<u>Capital</u> <u>charge</u>	<u>Total</u>	<u>\$/MMBTU</u>
1000	9,200	0	0.25	1.39	55.56	57.20	9.12
3000	27,600	0	0.25	0.97	38.76	39.98	6.38
9000	82,700	0	0.25	0.68	27.28	28.21	4.50

Table 10

CAPITAL INVESTMENT AND PRODUCTION COST OF METHANOL SYNTHESIS
FROM AIR, WATER AND FUSION ENERGY

CO₂ is Obtained by Refrigeration of Dry Air which is Obtained
by Removal of Water by Refrigeration

Process No. 6

<u>Capacity</u>		<u>Capital Investment</u>						
<u>MW(e)</u>	<u>bbl/day</u> <u>CH₃OH</u>	<u>Power Plant</u> <u>\$/kw</u>	<u>Electrolyzer</u> <u>\$/kw</u>	<u>Power Plant</u> <u>Electrolyzer</u> <u>(\$10⁶)</u>	<u>CO₂-</u> <u>separation</u> <u>\$/bbl/day</u>	<u>Methanol</u> <u>Synthesis</u> <u>\$/bbl/day</u>	<u>CO₂-CH₃OH</u> <u>total</u> <u>(\$10⁶)</u>	<u>Total</u> <u>capital</u> <u>(\$10⁶)</u>
1000	6,600	350	28.5	378	40,000	3,300	285	663
3000	19,800	285	22.9	924	21,000	2,150	458	1,382
9000	59,400	225	18.4	2,164	11,000	1,400	736	2,900

<u>Capacity</u>		<u>Production Cost ¢/gal.</u>					<u>Thermal value</u> <u>\$/MMBTU</u>
<u>MW(e)</u>	<u>bbl/day</u> <u>CH₃OH</u>	<u>Raw Material</u>	<u>Catalyst</u>	<u>Operating &</u> <u>Maintenance</u>	<u>Capital</u> <u>charge</u>	<u>Total</u>	
1000	6,600	0	0.25	1.99	79.72	81.96	12.96
3000	19,800	0	0.25	1.38	55.39	57.02	9.01
9000	59,400	0	0.25	0.97	38.74	39.96	6.32

Table 11

CAPITAL INVESTMENT AND PRODUCTION COST OF METHANOL SYNTHESIS
FROM AIR, WATER AND NUCLEAR FUSION ENERGY

CO₂ is Obtained by Absorption/Stripping of Air with K₂CO₃ Solution

Process No. 7

Capacity		Capital Investment						
MW(e)	bbl/day CH ₃ OH	Power Plant \$/kw	Electrolyzer \$/kw	Power Plant Electrolyzer (\$10 ⁶)	CO ₂ - separation \$/bbl/day	Methanol Synthesis \$/bbl/day	CO ₂ -CH ₃ OH total (\$10 ⁶)	Total capital (\$10 ⁶)
1000	21,400	350	28.5	378	2,082	2,290	94	472
3000	64,200	285	22.9	924	1,341	1,480	181	1,105
9000	193,000	225	18.4	2,164	863	970	353	2,517

Capacity		Production Cost ¢/gal.						
MW(e)	bbl/day CH ₃ OH	Raw Material	K ₂ CO ₃	Catalyst	Operating & Maintenance	Capital charge	Total	Thermal value \$/MMBTU
1000	21,400	0	0.12	0.25	0.44	17.50	18.31	2.89
3000	64,200	0	0.12	0.25	0.34	13.66	14.37	2.27
9000	193,000	0	0.12	0.25	0.26	10.35	10.98	1.73

Table 12

CAPITAL INVESTMENT AND PRODUCTION COST OF CH₃OH SYNTHESIS
FROM CO₂, H₂O AND NUCLEAR FUSION ENERGY

CO₂ is Obtained by Absorption/Desorption from Air by Molecular Sieves

Process No. 8

Capacity		Capital Investment						
MW(e)	bbl/day CH ₃ OH	Power Plant \$/kw	Electrolyzer \$/kw	Power Plant Electrolyzer (\$10 ⁶)	CO ₂ - separation \$/bbl/day	Methanol Synthesis \$/bbl/day	CO ₂ -CH ₃ OH total (\$10 ⁶)	Total capital (\$10 ⁶)
1000	8,700	350	28.5	378	12,000	3,300	133	511
3000	26,000	285	22.9	924	5,800	2,100	205	1,129
9000	78,000	225	18.4	2,164	2,800	1,380	326	2,490

Capacity		Production Cost ¢/gal.					Thermal value \$/MMBTU
MW(e)	bbl/day CH ₃ OH	Raw Material	Catalyst	Operating & Maintenance	Capital charge	Total	
1000	8,700	0	0.25	1.17	46.61	48.03	7.58
3000	26,000	0	0.25	0.86	34.46	35.57	5.62
9000	78,000	0	0.25	0.63	25.23	26.21	4.14

Table 13

CAPITAL INVESTMENT AND PRODUCTION COST OF CH₃OH SYNTHESIS
FROM CO₂, H₂O AND NUCLEAR FUSION

CO₂ is Obtained by Stripping from Sea Water

Process No. 9

Capacity		Capital Investment						
MW(e)	bbl/day CH ₃ OH	Power Plant \$/kw	Electrolyzer \$/kw	Power Plant Electrolyzer (\$10 ⁶)	CO ₂ - separation \$/bbl/day	Methanol Synthesis \$/bbl/day	CO ₂ -CH ₃ OH total (\$10 ⁶)	Total capital (\$10 ⁶)
1000	21,700	350	28.5	378	1,146	2,300	75	453
3000	65,000	285	22.9	924	738	1,500	145	1,069
9000	195,000	225	18.4	2,164	480	970	283	2,447

Capacity		Production Cost ¢/gal.					Thermal value
MW(e)	bbl/day CH ₃ OH	Raw Material	Catalyst	Operating & Maintenance	Capital charge	Total	\$/MMBTU
1000	21,700	0	0.25	0.41	16.56	17.22	2.79
3000	65,000	0	0.25	0.32	13.05	13.84	2.14
9000	195,000	0	0.25	0.25	9.95	10.45	1.65

Table 14

CAPITAL INVESTMENT AND PRODUCTION COST OF CH₃OH SYNTHESIS
FROM CO₂, H₂O AND NUCLEAR FUSION (CTR) POWER
CO₂ is Obtained from Lime Plant

Capacity		Capital Investment						
MW(e)	bbl/day CH ₃ OH	Power Plant \$/kw	Electrolyzer \$/kw	Power Plant Electrolyzer (\$10 ⁶)	CO ₂ - separation \$/bbl/day	Methanol Synthesis \$/bbl/day	CO ₂ -CH ₃ OH total (\$10 ⁶)	Total capital (\$10 ⁶)
1000	22,400	350	28.5	378	700	2,248	66	444
3000	61,000	285	22.9	924	453	1,450	128	1,052
9000	201,000	225	18.4	2,164	292	934	246	2,410

Capacity		Production Cost ¢/gal.						
MW(e)	bbl/day CH ₃ OH	Raw Material	Trans- portation	Catalyst	Operating & Maintenance	Capital charge	Total	Thermal value \$/MMBTU
1000	22,400	1.5	3.0	0.25	0.39	15.73	20.37	3.29
3000	67,100	1.5	3.0	0.25	0.31	12.44	17.00	2.69
9000	201,000	1.5	3.0	0.25	0.24	9.52	14.01	2.21

Flow Diagram for the Metal-Steam
Binary-Vapor Cycle Power Conversion
System Applied to Controlled Thermo-
nuclear Fusion Reactor

Figure 1

Figure 2

Process 1 - Methanol Synthesis from CO_2 , H_2O and Nuclear Fusion Energy (CTR) - CO_2 is Obtained from Absorption/Stripping of Air by Water at Atmospheric Pressure.

Process 2 - Methanol Synthesis from CO₂, H₂O and Nuclear Fusion Energy (CTR) - CO₂ is Obtained from Absorption/Stripping of Air by H₂O. Part of the CO₂ is Absorbed in H₂O in the Cooling Tower.

Figure 3

Figure 5

Process 4 - Methanol Synthesis from CO₂, Water and Nuclear Fusion Energy (CTR) - CO₂ is Obtained from Absorption/Stripping by Methanol.

Figure 7

Process 6 - Methanol Synthesis from CO₂, H₂O and Nuclear Fusion Energy (CTR) - CO₂ is Extracted from Air by Refrigeration.

Process 5 - Methanol Synthesis from CO₂, Water and Nuclear Fusion Energy (CTR) - CO₂ is Extracted from Air by Refrigeration and Water Vapor is Removed by Absorption on Molecular Sieves.

Figure 6

Figure 8

Process 7 - Methanol Synthesis from CO_2 , H_2O and Nuclear Fusion Energy (CTR) - CO_2 is Obtained from Absorption of Air by K_2CO_3 Solution and Desorption by Hydrogen.

Process 8 - Methanol Synthesis from CO₂, Water and Nuclear Fusion Energy (CTR) - CO₂ and H₂O is Removed from Air by Adsorption/Desorption from Molecular Sieves.

Figure 9

Process 9 - Methanol Synthesis from CO_2 , H_2O and Nuclear Fusion Energy (CTR) - CO_2 is Obtained from Stripping of Seawater.

Figure 10

- (1) H₂O ABSORPTION/STRIPPING (atm PRESSURE)
- (2) H₂O ABSORPTION/STRIPPING (atm PRESSURE)
WITH COOLING TOWER
- (3) H₂O ABSORPTION/STRIPPING (20 atm)
- (4) CH₃OH ABSORPTION/STRIPPING (20 atm, -80°F)
- (5) H₂O ADSORPTION ON MOLECULAR SIEVES,
CO₂ REFRIGERATION
- (6) H₂O REFRIGERATION, CO₂ REFRIGERATION
- (7) K₂CO₃ ABSORPTION/STRIPPING (atm PRESSURE)
- (8) H₂O & CO₂ ADSORPTION/DESORPTION BY
MOLECULAR SIEVES
- (9) STRIPPING SEA WATER AT 1 atm PRESSURE

Figure 11 - CAPITAL INVESTMENT OF METHANOL SYNTHESIS
vs PRODUCTION RATE BY DIFFERENT PROCESSES
OF CO₂ RECOVERY

- (1) H₂O ABSORPTION/STRIPPING (atm PRESSURE)
- (2) H₂O ABSORPTION/STRIPPING (atm PRESSURE)
WITH COOLING TOWER
- (3) H₂O ABSORPTION/STRIPPING (20 atm)
- (4) CH₃OH ABSORPTION/STRIPPING(20 atm,-80°F)
- (5) H₂O ADSORPTION ON MOLECULAR SIEVES,
CO₂ REFRIGERATION
- (6) H₂O REFRIGERATION, CO₂ REFRIGERATION
- (7) K₂CO₃ ABSORPTION/STRIPPING (atm PRESSURE)
- (8) H₂O & CO₂ ADSORPTION/DESORPTION BY
MOLECULAR SIEVES
- (9) STRIPPING SEA WATER AT 1 atm PRESSURE

Figure 12 — PRODUCTION COST OF METHANOL SYNTHESIS
vs PRODUCTION RATE BY DIFFERENT PROCESSES
OF CO₂ RECOVERY