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INTERIM TECHNICAL STATUS REPORT
FOR THE COthane CONTRACT

MASTER

December 1979

Work Performed Under Contract No. AC03-78CS40177

Union Carbide Corporation
Tarrytown Technical Center
Tarrytown, New York

U. S. DEPARTMENT OF ENERGY

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FOR THE COTHANE CONTRACT
(DE-AC03-78CS40177)

DECEMBER, 1979

UNION CARBIDE CORPORATION
TARRYTOWN TECHNICAL CENTER
TARRYTOWN, NEW YORK

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SUMMARY

Each year millions of tons of dilute carbon monoxide in industrial waste streams are flared or vented to the atmosphere. In order to recover this wasted source of energy, Union Carbide has developed the COthane process for the direct conversion of the dilute carbon monoxide in such streams to pipeline quality natural gas (SNG).

During the first step of this two step, multi-bed process, the waste gas is passed over a bed of catalyst where the carbon monoxide disproportionates into ejected carbon dioxide and deposited active carbon. During the second step, moderate pressure steam reacts with the active carbon to form methane and carbon dioxide. Removal of the carbon dioxide and residual water yields pipeline quality SNG.

A three phase program has been planned for the commercialization of the COthane process. Phase I will define the process details. Phase II will further test the process with an integrated pilot unit operating on industrial waste streams. Phase III will monitor the operation of the first commercial unit.

The primary purpose of DOE contract DE-AC03-78CS40177, which runs from 25 September 1978 to 31 July 1980, is to carry out the Phase I effort. This effort has been broken down into five tasks, which cover the bench scale pilot runs (Tasks 1 and 2), the development of an improved catalyst (Task 3), and the technical-economic studies (Tasks 4 and 5). The purpose of this interim report is to describe the preliminary work done under these five tasks during the first fourteen months of the contract.

The bench scale pilot unit has been built and initial test runs started. New catalysts have been developed and tested which incorporate different matrixing agents, metal components, stabilizing agents, and synthesis techniques. Finally, the technical and economic studies have shown that, under certain optimistic conditions, a COthane unit handling blast furnace off-gas will produce SNG at \$3.17/MM BTU. This cost is probably competitive with currently priced natural gas, and is certainly competitive with SNG produced by the Lurgi process.

The objective of this contract, to define the basic design parameters of the COthane process, will be achieved by the contract's 31 July, 1980 expiration date. By that time sufficient technical, economic and marketing data will have been obtained to make a decision as to whether or not Phase II should be pursued, and if it should be pursued, to carry out the design of the Phase II pilot plant.

INTRODUCTION

Each year over 24 million tons of dilute carbon monoxide are wastefully flared or vented to the atmosphere by steel mills, foundries, oil refineries, and carbon black plants. This is the energy equivalent of about 570 million cubic feet/day of natural gas. Until now, no economical means of recovering this energy has existed.

Even in its pure state, carbon monoxide is a toxic, low quality fuel that requires special burners for efficient energy recovery. In a very dilute form, it may cost more to burn than the value of the energy produced.

In recent years, efforts have focussed on converting CO in the presence of H₂ to methane (methanation). This appears to be marginally economical for large coal gasification plants where the CO and hydrogen are present together in high concentrations.

In dilute concentrations, the carbon monoxide would first have to be separated. This is an expensive operation by itself. Then hydrogen would have to be added to convert the concentrated CO to a useful methane fuel. The cost of these routine processes of concentration and methanation are presently greater than the value of the energy recovered.

Union Carbide has been working for several years in the fields of catalysis, coal gasification, methanation and related research. As a result of these efforts, a novel process has been developed for the direct concentration and conversion of carbon monoxide in dilute streams to high quality methane (SNG) at a cost believed to be competitive with natural gas. This new process has been dubbed COthane— a contraction of the feed carbon monoxide (CO) and product methane.

THE COTHANE PROCESS

The COthane process(1) for the carbon monoxide-to-methane conversion is a two-step, multiple reactor system. Figure 1 shows that during the first step the waste gas is appropriately pre-treated, preheated, and passed through the first reactor containing a bed of nickel or cobalt catalyst. At ~300°C the catalyst causes the carbon monoxide in the stream to disproportionate into CO₂ and an active carbon species, C*. The CO₂ is carried out of the reactor along with the diluents in the waste gas. The active carbon is deposited at the surface of the catalyst. This first step can be expressed as:

This reaction continues in the first reactor until most of the catalyst has been loaded with active carbon. The waste gas is then diverted to a second reactor containing catalyst which has been stripped of its previously accumulated carbon. Moderate pressure steam is simultaneously introduced into the first reactor where it reacts with the active carbon to form methane and carbon dioxide. This second step can be expressed as:

~~This reaction continues in the first reactor until all of the~~
active carbon has been removed from the catalyst. The duration of this step is the same as that of the first step, so that upon its completion the waste gas will be switched back from the second reactor to the first reactor. This cycle continuously repeats itself.

The wet CO₂ and methane stream produced from the second step is treated to remove the carbon dioxide and water, so that the remaining high quality methane may be piped into the local natural gas distribution system.

The feed pretreatment step shown in Figure 1 will depend on the nature of the constituents in the feed gas. If the feed gas contains either hydrogen, water vapor, or oxygen, these compounds will react with the active carbon as quickly as it is formed to produce either methane (from the hydrogen), methane and carbon dioxide (from the water vapor), or carbon dioxide alone (from the oxygen). Furthermore, the methane formed in this manner will leave the reactor in the effluent during the disproportionation step, and consequently will not be recoverable during the subsequent steaming step. In addition to lowering the overall efficiency of the conversion of the waste carbon monoxide to recoverable methane, the reactions of hydrogen, water vapor, and oxygen with the active carbon are highly exothermic and will increase the cooling requirements for the already highly exothermic disproportionation step. Consequently, all of these feed constituents should be kept to some reasonable fraction of the carbon monoxide content before the feed gas is sent to the COthane reactors. This might be accomplished with a pre-oxidation step for the oxygen, a dehydration step for the water vapor, and either a pre-methanation or partial oxidation⁽²⁾ step for the hydrogen.

The feed gas should also be treated to remove catalyst poisons. Hydrogen sulfide, for instance, requires the use of not only a preliminary treatment step, such as an amine scrubber, but the use of a zinc oxide guard bed as well.

During each of the COrthane process cycles described above, there will be a small fraction of carbon formed during the disproportionation step that will not be removed during the steaming step. This "inactive" carbon will slowly build-up as the number of cycles increases, until its presence on the catalyst hinders the efficiency of the cycle. When the methane production rate drops to a minimum allowable level, the reactor containing this inhibited catalyst is withdrawn from service, replaced with a reactor containing carbon-free catalyst, and is itself subjected to a carbon removal step. This "regeneration" step is usually carried out with a low concentration ($\sim \frac{1}{2}$ vol.%) oxygen stream at a 400°C initiation temperature; when all of the carbon has been oxidized in this manner, the catalyst is then subjected to a hydrogen-containing stream so that the nickel oxide formed during the oxidation step is reduced back to nickel.

Alternatively, the carbon on the catalyst can be selectively oxidized with the carbon dioxide⁽³⁾ produced from the steaming step of the COrthane process via the reaction $C + CO_2 \rightleftharpoons 2CO$. The oxidation of the nickel, via the reaction $Ni + CO_2 \rightleftharpoons NiO + CO$, is suppressed by the addition of a small amount of carbon monoxide into the entering carbon dioxide regeneration gas. This influent carbon monoxide concentration is considerably less than the final carbon monoxide equilibrium concentration resulting from the $C + CO_2 \rightleftharpoons 2CO$ reaction. The regeneration effluent gas, having this equilibrium CO concentration, is fed back to the feed gas entering the reactor that is undergoing the disproportionation step. As a result of this recycle loop, the inactive carbon ends up as product methane and carbon dioxide.

Combining the disproportionation step reaction (Reaction 1) with the steaming step reaction (Reaction 2) yields the overall C0thane reaction,

Four volumes of carbon monoxide produce one volume of methane. Since one volume of methane has 3.15 times the gross heat of combustion of one volume of carbon monoxide, 78.7% of the gross energy available in the original carbon monoxide remains in the product methane $\{(3.15)(100)/(4) = 78.7\}$.

The flow scheme from the cyclic C0thane process not only has the operating simplicity of the commercial pressure-swing adsorption processes, but, simplistically, it performs the separation of carbon monoxide from its diluents and the subsequent conversion of that CO to moderate pressure methane in a single reactor, without the need for the standard, expensive steps of CO concentration, steam shifting, methanation, and tail gas compression. Consequently, the C0thane process is potentially a relatively uncomplicated and inexpensive way to transform waste carbon monoxide into pipeline quality natural gas.

THE COTHANE PROGRAM

The COthane program is divided into three phases. Phase I will define the basic process details, including design parameters and the preferred catalyst composition. Phase II will prove out the results of Phase I in an integrated pilot unit operating on industrial gases. The engineering data from Phase II will be used to design the first commercial unit under Phase III.

The primary purpose of DOE contract DE-AC03-78CS40177, which runs from 25 September 1978 to 31 July 1980, is to carry out the Phase I effort of developing the COthane process to the point where it is ready to be demonstrated on a large scale process development unit (PDU). This Phase I effort has been broken down into five tasks.

Tasks 1 and 2 call for the construction and operation of a miniature process development unit (mini PDU) to determine the process variables and efficiencies of the COthane process steps for the most promising catalysts, using both a simple CO/N₂ mixture (Task 1) and more complex synthetic feed streams (Task 2). This mini PDU reactor is of a type that will insure at least approximate design data for the type of commercially-sized reactor envisioned from the studies done as part of Tasks 4 and 5. It should be realized that, because of its small size, this mini PDU reactor is expected to yield basic design data for a commercial reactor that will undoubtedly need to be confirmed by runs on a large-scale PDU reactor, during Phase II.

Task 3 calls for a sustained effort to develop a more effective catalyst. Such a catalyst is expected to show improvements in efficiency, stability, resistance to poisons, and physical properties.

Task 4 calls for a technical feasibility study of the various reactor configurations that can be used to carry out the two exothermic steps of the COthane process in a commercially-sized unit. Task 5 calls for the costing of these different possible reactor configurations to determine the most economically and technically attractive design. The mini PDU reactor used in Tasks 1 and 2 is designed to reflect, as closely as possible for a small-scale laboratory reactor, the behavior of the chosen commercial unit. These technical and economic analyses are continually updated as experimental data becomes available.

The preliminary work that has been carried out from 25 September 1978 to 30 November 1979 is reported in this interim report.

TASKS 1 AND 2

The mini PDU reactor has been built, debugged, and subjected to an initial set of trial runs. The design of this reactor and the way it is being operated simulates the design and operation of the presently perceived commercial reactor. This perceived commercial reactor was found from the Task 4 and 5 studies to be the most economically and technically attractive type of reactor for carrying out the two exothermic steps of the COthane process.

TASK 3

The development of a more effective COthane catalyst has entailed the construction of a suitable catalyst screening unit, and the subsequent testing of different catalyst formulations in that screening unit.

Figure 2 shows the flow diagram for the catalyst screening unit. During the four minute disproportionation step 1 liter/minute of a 20 vol.% CO/80 vol.% N₂ gas mixture is fed from a cylinder, through a preheat coil, through the reactor, and on through a back pressure regulator set at 5 psig. The effluent gas is either vented or collected in a gas collection vessel. The preheat coil and reactor are immersed in a fluidized bed temperature bath, which is usually maintained at 270°C. Upon completion of the disproportionation step, the system is pressurized to 150 psig with nitrogen, and the four minute steaming step is begun. A little over 0.2 ml/min of distilled water is pumped through the preheat coil, where it is vaporized, and on through the reactor. The product methane and carbon dioxide are displaced through a back pressure regulator set at 150 psig, and into either the vent line or a second gas collection vessel. The system is then slowly depressurized to ~1 psig over a one minute period, and the released gas is again either vented or collected in the same gas collection vessel used for the steaming step.

The unit repeats this nine minute cycle automatically. During most of the cycles the effluent streams are vented. However, during other, designated cycles the gas samples are collected for the disproportionation and steaming/depressurization steps, and are analyzed for their methane and carbon dioxide contents.

The quantities of these gases indicate the performance of the catalyst during each step of the COthane cycle. The most important measure of the performance of a catalyst is its methane production rate, expressed as ml of methane/gm of catalyst-cycle.

The 0.033 gm moles $\{(4)(0.2)(1)/(24.2) = 0.033\}$ of carbon monoxide fed into the reactor would require, from Equation 3, 0.017 gm moles $\{(0.5)(0.033) = 0.017\}$ of water, if all of the carbon monoxide should disproportionate. The actual amount of water fed to the reactor was 0.044 gm moles $\{(4)(0.2)/(18) = 0.044\}$, or 2.1 times $\{0.044/0.017 = 2.1\}$ greater than would be theoretically required. Furthermore, the 0.8 liters $\{(4)(0.2)(1) = 0.8\}$ of CO fed to the eight grams of catalyst in the reactor is capable of theoretically yielding 25 ml of methane/gm of catalyst $\{(0.25)(0.8)(1000)/(8) = 25\}$ during each complete COthane cycle. Since this amount of methane is considerably greater than the 5 to 10 ml/gm-cycle of methane that was actually produced by most of the tested catalysts during each cycle, it can be concluded that these measured rates were not limited by the amount of CO and water fed to the reactor during each cycle.

Figure 3 shows that the catalyst screening reactor itself consisted of a 5" long, 5/8" O.D. piece of stainless steel tubing with a 20 cm³ capacity that held the 8 gm catalyst sample and ~10 grams of diluent quartz chips. A thermocouple, placed approximately one third the distance down from the top of bed, served as the standard temperature indicator for all of the test runs. In each case the temperature at that point would be close to the temperature of the fluidized temperature bath (typically 270°C) at the beginning of the disproportionation step. That

bed temperature would then climb during the disproportionation step, level out (typically $\sim 300^{\circ}\text{C}$) during the steaming step, and return back down to the fluidized bath temperature by the end of the depressurization step. During the initial runs made with this reactor, a second thermocouple was placed in the space directly above the bed, in order to verify that the 20 feet of 1/8" preheat tubing was sufficiently long to bring the temperatures of both the disproportionation and steaming influents up to the fluidized bath temperature.

The 270°C to 300°C temperature swing experienced by the catalyst bed during each cycle is representative of what the catalyst bed in the presently envisioned commercial reactor would experience under analogous process conditions. Consequently, this catalyst screening unit is considered to be an appropriate tool to test improved catalyst formulations.

Each of these new formulations were placed in the catalyst screening reactor, heated with the fluidized temperature bath to 400°C , reduced with a 4 vol.% H_2 /96 vol.% N_2 stream, cooled to 270°C , and then subjected to the first set of test cycles. After this first set of cycles the catalyst was usually preheated to 400°C , burnt off with a 1 vol.% O_2 , 99 vol.% N_2 stream, reduced with a 4 vol.% H_2 /96 vol.% N_2 stream, cooled down and subjected to the second set of test cycles.

The catalyst formulation used in the COthane process has to catalyze the disproportionation of CO and to support the active carbon formed in the first step, as well as to catalyze the reaction of the active carbon with steam to form equimolar quantities of methane and carbon dioxide. In order to perform these

tasks satisfactorily, the COthane catalyst must a) have a high capacity and selectivity for the formation of active carbon with an attendant resistance to the formation of inactive carbon, b) withstand numerous oxidative regenerations without a significant long term loss in either its capacity, selectivity, or activity, and c) have good mechanical stability.

These characteristics were partially met by the nickel-zirconia catalyst originally developed for the COthane process. This formulation was prepared by the addition of 1 liter of 1.15 molar solution of sodium hydroxide into a 1 liter solution containing 0.5 mole (145.5 gms) of $\text{Ni}(\text{NO}_3) \cdot 6\text{H}_2\text{O}$ and 0.0375 mole (10.0 gms) of $\text{ZrO}(\text{NO}_3)_2 \cdot 2\text{H}_2\text{O}$. The resulting gel was then filtered, washed with distilled water, and subsequently dried at 65°C in air. The resulting catalyst was reduced in a stream of 5% hydrogen in nitrogen with the temperature programmed to increase gradually to 400°C, where it was maintained for several hours.

Efforts to improve the long term capacity, selectivity, and activity of this original formulation have been directed towards trying new matrixing agents, new metal compositions, stabilizing additives, and different synthesis techniques. Furthermore, the attainment of good mechanical stability should result from the continual product development effort which is accompanying the large scale production of five-pound batches of the more promising catalyst formulations required for the mini PDU unit, as dictated by the Task 1 and 2 studies.

Figure 4 shows the results of the second set of cycles for the four catalysts tested in November. They are typical of the results obtained for all of the catalysts tested to date. Most

of the runs were made with the minimum temperature of the cycle at $\sim 270^{\circ}\text{C}$. However, Figure 4 shows that this minimum temperature was raised during the course of several runs to $\sim 290^{\circ}\text{C}$, in order to study the effect of such a temperature increase. Figure 4 also indicates that the methane production rate for most of the catalysts starts out from a wide range of relatively high values and decays to a fairly constant level determined by the catalyst formulation and the minimum bed temperature.

TASK 4

A study of reactor configurations for the COthane process requires a consideration of the amount of heat released during each step of the process. Theoretical calculations showed that ~80 kilocalories are released for the overall reaction $4\text{CO} + 2\text{H}_2\text{O} \rightleftharpoons \text{CH}_4 + 3\text{CO}_2$. Experimental evidence indicates that both the disproportionation and steaming steps are exothermic. Consequently, any reactor system considered for COthane service must be able to handle the maximum temperature rises that occur during these two steps. Furthermore, the overall process must also insure that the total heat released during each complete cycle is removed from the reactor prior to the beginning of the next cycle.

Several reactor configurations were studied from the viewpoint of maximum temperature rise and heat dissipation. They ranged from a simple adiabatic packed bed reactor to a fluidized bed reactor. The modified process steps dictated by each of these designs were thoroughly explored, so that the total cyclic behavior, as well as the inherent restraints, of each of these configurations were defined in detail.

TASK 5

Preliminary economic studies have been carried out on the more promising reactor configurations defined by the Task 4 studies. The feed stream for each of these economic studies was assumed to be 15.7 MM SCFD of water-saturated, 100°F, 0 psig, blast furnace off-gas containing 27.5 vol.% CO. It was also assumed that the catalyst would last two years and would have a 10 ml/gm-catalyst methane production rate.

Figure 5 shows that during the disproportionation step the blast furnace off-gas passes through a dehydrator, an influent/effluent heat exchanger, a ZnO guard bed, a trim preheater, and on to the reactor. During the steaming step 150 psig steam is passed through a preheater and into the reactor; the reaction products pass through a carbon dioxide removal unit, a final dehydrator, and on to the local natural gas pipeline distribution system.

The costing techniques used for this COthane cost study followed the same techniques that were used in a recent ESCOE report(4) for the Lurgi and fourteen other less proven processes. The values for these production costs included feed, operating and capital cost components. The feed gas for the COthane process was assumed to be previously flared off-gas, which had no use to the user, and consequently had a zero value for the study. The yearly operating cost included charges for chemicals, catalysts, labor, maintenance, and insurance, which amounted to \$689,900/yr. The yearly capital cost component was the product of the total capital cost, \$4,088,000, and a 0.115 capital factor, yielding \$470,100/yr. The capital factor was for private financing,

and took into consideration a 17% ROI, the project life, the tax life, the investment tax credit, the tax rate, the interest during construction, and the capital recovery. When the yearly operating and capital costs are combined and divided by the 365,994 MM BTU/yr heat value of the methane produced each year, the production cost becomes \$3.17/MM BTU (mid 1979 dollars).

This \$3.16/MM BTU production cost for the COthane process, incorporating the most desirable reactor configuration and assuming optimistic catalyst properties, appears to be competitive with currently priced natural gas. It is also competitive with SNG produced by the Lurgi process. Table I shows that the cost⁽⁴⁾ of SNG produced by a large Lurgi unit situated at an outlying coal deposit (\$4.22/MM BTU) and transported (~\$0.52/MM BTU) to more populated, industrialized areas, where COthane units would probably be located, would amount to \$4.74/MM BTU (mid 1979 dollars).

REMAINING WORK

Work will continue on the Phase I effort until the expiration of the DOE contract on 31 July 1980. By that time the contract's objective of developing the COthane process to the point where it can be demonstrated on the large scale (Phase II) PDU unit will have been met.

The Tasks 1 and 2 mini PDU runs will yield sufficient data to insure the design of both the Phase II large scale PDU unit and the perceived commercial reactor required for the updating of the Task 5 economic studies. The Task 3 catalyst development effort will continue to be aimed at improving the efficiency, stability, and physical properties of the COthane catalyst. The Task 4 technical feasibility studies will translate the data obtained from the Tasks 1 and 2 mini PDU runs into design data for both the Phase II large scale PDU and the commercial reactor envisioned for the Task 5 economic studies. The Task 5 economic studies will update the previously optimistic methane production cost for an envisioned commercial reactor as the data from the other tasks becomes available.

This Phase I work will be matched by an accompanying market potential study, so that by the middle of 1980 the technical viability, economic attractiveness, and marketability of the COthane process will be better defined. These factors will be weighed in order to decide whether or not the COthane program should proceed any further. Therefore, the completion of Phase I will not only yield the data required to design the large scale PDU of Phase II, but it will also supply the bases for a go or no go decision for Phase II.

BIBLIOGRAPHY

- (1) U.S. Patent Application 889,558, "Methanation of Carbon Monoxide Without Prior Separation of Inert Gases", Filed March 23, 1978.
- (2) Pending Patent Application, "Enhanced Methanation of Gas Streams Containing Carbon Monoxide and Hydrogen", Serial No. 83473, October 10, 1979.
- (3) Pending Patent Application, "Methanation of Carbon Monoxide With Improved Catalyst Regeneration", Serial No. 82472, October 9, 1979.
- (4) K. A. Rogers and R. F. Hill, "Coal Conversion Comparisons" FE-2468-51, prepared for the United States Department of Energy under Contract No. EF-77-C-01-2468 by the Engineering Societies Commission on Energy Inc., July, 1979.

TABLE I

COMPARISON BETWEEN THE COTHANE AND LURGI SNG PRODUCTION COSTS⁽¹⁾

<u>PROCESS</u>	<u>PRODUCTION COST \$/MM BTU</u>
Cothane ⁽²⁾	\$3.17
LURGI ⁽³⁾	
With feed coal priced at \$1.33/MM BTU	\$4.22
+ <u>Average 1978 Interstate Pipeline Transportation Cost</u>	\$0.52
TOTAL DELIVERED TO CITY LURGI COST	\$4.74

- (1) Both production costs include a 17% ROI
- (2) Based on a Cothane unit using 15.7 MM SCFD of blast furnace off-gas.
- (3) Based on a Lurgi SNG plant using 25,000 tons of dry coal/day situated near extensive coal deposits.

FIGURE 1
SCHEMATIC DIAGRAM FOR THE COTHANE PROCESS

FIGURE 2
FLOW DIAGRAM FOR THE CATALYST TESTING UNIT

FIGURE 3
STAINLESS STEEL PACKED BED REACTOR

FIGURE 4

NOVEMBER CATALYST SCREENING RUNS WITH THE TEST REACTOR

FIGURE 5
THE COTHANE PROCESS HANDLING BLAST FURNACE OFF-GAS