



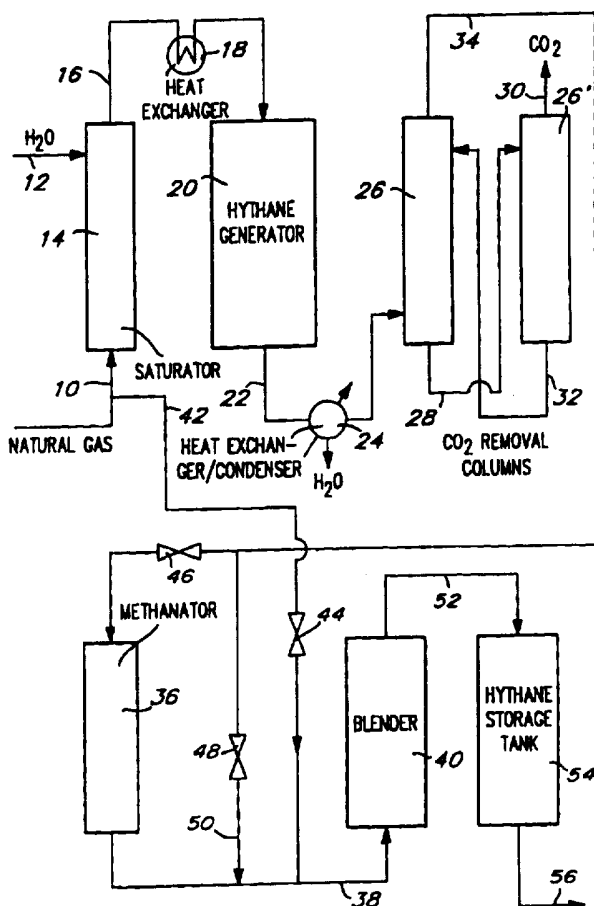
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(21) International Application Number: PCT/CA96/00048 (22) International Filing Date: 23 January 1996 (23.01.96) (30) Priority Data: 2,141,065 25 January 1995 (25.01.95) CA (71) Applicant: GAZ METROPOLITAIN AND COMPANY, LIMITED PARTNERSHIP [CA/CA]; 1717 Du Havre, Montreal, Quebec H2K 2X3 (CA). (71)(72) Applicant and Inventor: PANDEY, Raj, Narain [IN/CA]; 34 Old Colony Trail, Guelph, Ontario N1G 4A9 (CA). (72) Inventors: RATNANI, Kébir; 205 De Bayeux, Boucherville, Quebec J4B 7T9 (CA). AHMED, Shamsuddin; 247 Cole Road, Guelph, Ontario N1G 3K1 (CA). PANDEY, Rupesh, Narain; 34 Old Colony Trail, Guelph, Ontario N1G 4A9 (CA). WILLIAMS, John, R.; 2595 De l'Epervier, Laval, Quebec H7L 3Y6 (CA). (74) Agent: CARRIER, Robert; Swabey Ogilvy Renault, Suite 1600, 1981 McGill College Avenue, Montreal, Quebec H3A 2Y3 (CA).		(81) Designated States: AU, BR, CN, FI, HU, JP, KR, NZ, PL, RU, VN, European patent (AT, BE, CH, DE, DK, ES, FR, GB, GR, IE, IT, LU, MC, NL, PT, SE). Published <i>With international search report.</i> <i>Before the expiration of the time limit for amending the claims and to be republished in the event of the receipt of amendments.</i>

(54) Title: DIRECT CONVERSION OF METHANE TO HYTHANE

(57) Abstract

The direct conversion of methane to hythane, an alternative fuel for internal combustion engines, is effected by oxidizing methane oxidation with water vapor at a temperature of about 400 to about 500 °C and a pressure of about 1 to about 5 atmospheres, in the presence of a catalyst comprising palladium or a binary alloy thereof with another metal selected from the group consisting of elements belonging to Group IB, Group IIA, Group VIII and the lanthanide series of the Periodic Table, the catalyst being supported on a porous carrier. The direct conversion of methane to hythane under conditions of low temperature and pressure in accordance with the present invention enables hythane to be produced not only economically and efficiently, but also at or near vehicle fueling sites, thereby eliminating the storage and transportation of dangerously reactive hydrogen for mixing with methane or natural gas.



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DIRECT CONVERSION OF METHANE TO HYTHANE

TECHNICAL FIELD

5 The present invention relates to a novel process for the efficient and economic production of hythane, an alternative fuel for internal combustion engines. More particularly, the invention relates to a process for the direct conversion of methane to hythane.

10

BACKGROUND ART

 The growing regulatory and legislative demands for gradual introduction of pollution-free vehicles in
15 Canada and other industrialized nations of the world have intensified research to find clean burning transportation fuels. Compressed natural gas, methanol and oxygenated fuels are among the leading candidates. Recently, blends of 5 - 20 vol. % hydrogen and 80 - 95
20 vol. % natural gas have been receiving increasing interest. These gaseous mixtures are called "hythane" and have been described by F.E. Lynch et al in US Patent N° 5,139,002 as having an effective combustion rate similar to that of gasoline, thereby creating a
25 potentially promising substitute for conventional fuels in spark ignition internal combustion engines as well as in compression ignition engines. The pollution emissions from hythane powered vehicles is shown to be well below that of gasoline engines, due to the clean-
30 burning characteristics of the components of hythane (F. E. Lynch and G. J. Egan, Proceedings of the 4th Cdn. Hydrogen Workshop, Nov. 1-2, 1989).

According to the aforementioned US Patent N° 5,139,002, hythane is produced by blending natural gas and hydrogen in desired proportions. Natural gas is a cheap and abundantly available material. However, this is not the case with hydrogen. At present, hydrogen is principally obtained by an energy- and capital- intensive natural gas steam reforming technology. The process operating under extreme conditions of temperature ($>900^{\circ}\text{C}$) and pressure (>20 atmospheres) produces a mixture of H_2 , CO and CO_2 , having a H_2/CO ratio of about 3 to 5, and containing typically about 15% CO. The main purpose of steam reforming process is to produce syngas. The production of CO-free hydrogen from such a reactor effluent or product mixture requires either separation of hydrogen by membrane diffusion technology or conversion of CO to CO_2 by a shift conversion process under operating conditions vastly different from the primary steam reforming process. Either of these processes adds substantial cost to the already expensive catalytic steam reforming process. Because of such complexity of the overall manufacturing process, hydrogen is an expansive commodity. As a result, $\text{CH}_4\text{-H}_2$ blends have high prices as well. The situation is further worsened by the requirements of cryogenic storage and transportation of the dangerously reactive hydrogen for mixing with methane or natural gas.

DISCLOSURE OF INVENTION

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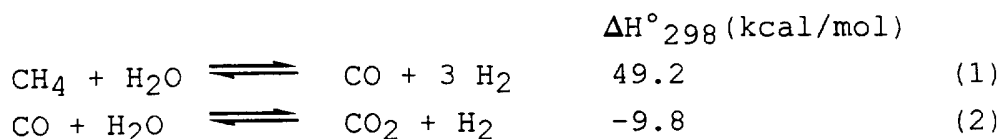
It is therefore an object of the present invention to provide an efficient and economic process for the direct conversion of methane to hythane.

In accordance with the invention, there is thus provided a process for the production of hythane, which comprises subjecting methane to a controlled oxidation with water vapor at a temperature of about 400 to about 500°C and a pressure of about 1 to about 5 atmospheres, in the presence of a catalyst comprising palladium or a binary alloy thereof with another metal selected from the group consisting of elements belonging to Group IB, Group IIA, Group VIII and the lanthanide series of the Periodic Table, the catalyst being supported on a porous carrier.

Applicant has found quite unexpectedly that by using the above catalyst methane or natural gas can be converted directly to hythane under controlled conditions of low temperature and pressure. The absence of severe temperature and pressure conditions keeps the capital and operating costs of the process of the invention within reasonable limits.

The expression "controlled oxidation" as used herein refers to an oxidation reaction which is carried out in the presence of the above catalyst and under the above controlled conditions of low temperature and pressure so that the reaction proceeds to the desired H₂ and CH₄ contents in the gaseous product mixture obtained, that is, about 5-20 vol.% H₂ and about 80-95 vol.% CH₄, corresponding to the composition of hythane.

The main reactions occurring during the controlled oxidation of methane with water vapor in accordance with the present invention are as follows:



5 The transformation of methane or natural gas to
hythane does not require high conversion of CH_4 , but
demands a minimal production of CO. This is achieved in
accordance with the invention by carrying out
controlled oxidation of CH_4 with H_2O at low reaction
10 temperature of 400-500°C, where forward water gas shift
reaction (reaction (2) above) is thermodynamically
favoured. The catalyst used according to the invention
also exhibits the desired dual properties of (a)
moderate activity for the CH_4 - H_2O reaction and (b) high
15 activity for the water gas shift reaction. Under these
conditions, the CO produced by reaction (1) is
converted to CO_2 and H_2 by reaction (2).

Examples of suitable catalysts which may be used
20 in accordance with the present invention include Pd,
Pd-Ni, Pd-Cu, Pd-Mg, Pd-La and Pd-Ce supported on a
porous carrier such as alumina, silica or zirconia.
Palladium supported on γ -alumina is particularly
preferred. The loadings of the catalytically active
25 components on the carrier can vary between about 0.5
and about 10% by weight, based on the total weight of
the supported catalyst.

According to a preferred embodiment of the
30 invention, an inert porous material such as γ -alumina
is admixed with the supported catalyst to prevent
excessive catalyst cooling (due to the endothermicity
of reaction (1) above) by serving as heat reservoir.
Preferably, the inert porous material is used in an

amount of about one to two-fold the mass of supported catalyst.

5 The composition of the hythane produced by the
process of the invention can be varied by adjusting the
process conditions in terms of $\text{CH}_4/\text{H}_2\text{O}$ feed ratio, dry
gas feed flowrate, catalyst mass and reaction
temperature and pressure. As already indicated above,
the reaction is carried out at a temperature in the
10 range of about 400-500°C and a pressure in the range of
about 1-5 atmospheres. An increase in the reaction
temperature has the effect of producing a hydrogen rich
hythane fuel, whereas an increase in the reaction
pressure decreases the overall conversion of methane,
15 thereby producing a hythane fuel with a lower hydrogen
content. The $\text{CH}_4/\text{H}_2\text{O}$ feed ratio and dry gas feed
flowrate, on the other hand, preferably range from
about 2 to about 5 and from about 20 to about 100
ml (NTP)/min., respectively. A $\text{CH}_4/\text{H}_2\text{O}$ feed ratio of
20 about 3.5 and a dry gas feed flowrate of about 25
ml/min. are preferred. The mass of supported catalyst
may vary within a range of about 0.2 to about 1 g. An
increase in the partial pressure of water vapor in the
feed has the effect of producing a hydrogen rich
25 hythane fuel, whereas an increase in the catalyst mass
or dry gas feed flowrate increases the conversion of
methane and thus increases the hydrogen content of the
fuel.

30 The process of the invention has the distinct
advantage of offering significant cost savings over
conventional processes for hythane production such as
those based on producing hydrogen by electrolysis of
water or by multistep high temperature-high pressure

steam reforming of methane, followed by blending of hydrogen with natural gas. The direct conversion of methane to hythane under conditions of low temperature and pressure in accordance with the present invention
5 enables hythane to be produced not only economically and efficiently, but also at or near vehicle fueling sites, thereby eliminating the storage and transportation of dangerously reactive hydrogen for mixing with methane or natural gas.

10

BRIEF DESCRIPTION OF THE DRAWINGS

Figure 1 is a flow diagram of a process for producing hythane according to the invention.

15

MODES OF CARRYING OUT THE INVENTION

In the process which is schematically illustrated in Fig. 1, purified natural gas and water are fed via
20 feed lines 10 and 12, respectively, to a saturator 14 for saturating the natural gas with water vapor. The water saturated gas stream which is discharged via line 16 is passed through a heat exchanger 18 for pre-heating the gas stream to a temperature of 300-350°C,
25 and then sent to a hythane generator 20 consisting of a downflow reactor containing a fixed catalyst bed maintained at a temperature of 400-450°C. As the pre-heated gas stream enters into the reactor 20, it is heated to the operating temperature of the reactor in
30 order to avoid cooling of the catalyst bed and thereafter passed through the catalyst bed. The transformation of the natural gas to hythane is carried out in the reactor 20 operating at 1 atmosphere absolute and 400-450°C. The reactor effluent which is

discharged via line 22 consists of a gaseous mixture containing methane, hydrogen and small quantities of carbon oxides and unreacted water vapor; in terms of CH_4 and H_2 , the reactor effluent has a composition
5 corresponding to that of hythane fuel.

The hythane-containing gas stream is passed through a heat exchanger/condenser 24 for recovering useful heat and condensing out excess water vapor, and
10 then sent to CO_2 towers 26, 26'. The removal of CO_2 from the hythane-containing gas stream can be accomplished by one of several standard processes, for example, by scrubbing with a monoethanolamine (MEA) solution. The hythane-containing gas stream passes through a CO_2
15 absorption tower 26 where CO_2 is absorbed at ambient temperature by MEA solution. The spent MEA solution is sent via line 28 to a stripper tower 26' where the absorbed CO_2 is separated from the scrubbing liquor by warming and discharged via line 30 to the atmosphere.
20 The regenerated MEA solution is recycled via line 32 to tower 26.

The gas stream leaving the CO_2 absorption tower 26 is sent via line 34 to a catalytic methanator 36 for
25 converting residual carbon monoxide to methane. The methanation is usually carried out over a silica supported nickel catalyst maintained at a temperature of $300\text{--}350^\circ\text{C}$. The methanator effluent which is discharged via line 38 and contains only CH_4 and H_2 is
30 sent to a blender 40. If the CH_4/H_2 ratio in the hythane fuel thus produced is lower than desired, the ratio can be adjusted in the blender by adding make-up natural gas via line 42 and opening valve 44. On the other hand, if the residual carbon monoxide in the gas

stream leaving the CO₂ absorption tower 26 is within permissible limits, the methanation step can be bypassed as a cost-saving measure by closing valve 46 and opening valve 48 so as to send the gas stream directly to the blender 40 via lines 34, 50 and 38. After blending, the gaseous mixture of CH₄ and H₂ is sent via line 52 to a storage tank 54 for ultimate fueling of automobiles or the like via line 56.

The following non-limiting examples further illustrate the invention.

EXAMPLE 1

A Pd/ γ -Al₂O₃ catalyst containing 2.0 wt.% Pd was prepared by incipient wetness impregnation of a γ -alumina support (20-40 mesh) with an aqueous solution of palladium chloride, followed by drying at 120°C.

A microreactor was packed with an admixture of 0.2 gram of the above catalyst with 0.3 gram of inert γ -Al₂O₃ and placed in a continuous flow system. Prior to the start of the reaction, the mixture of Pd/ γ -Al₂O₃ catalyst and inert γ -Al₂O₃ was calcined at 500°C under N₂ flow for 2 hours, followed by a reduction at 500°C under hydrogen flow for 2 hours. A feed stream consisting of a mixture of CH₄ (43.9 vol.%), water vapor (12.5 vol.%) and N₂ (43.6 vol.%) was passed over the catalyst bed; the CH₄/H₂O feed ratio was thus about 3.5. The nitrogen added to the feed acted as an internal standard. The flow of dry gas at the reactor inlet was maintained approximately at 25 ml/min., the pressure in the reactor being maintained at about 1 atmosphere. The reaction temperature (bed temperature)

was varied from 400 to 450°C. The reactor effluent stream was analyzed for reactants and products (N₂, CH₄, CO, CO₂, H₂O and H₂) by TCD - gas chromatography. The composition of the reactor effluent (dry gas basis
5 excluding internal standard) for the selected temperatures is reported in Table 1.

TABLE 1

Reaction Temperature (°C)	Reactor effluent composition (vol. %)				H ₂ /CO (v/v) ratio in effluent
	CH ₄	H ₂	CO	CO ₂	
400	89.2	9.0	0.40	1.41	22
425	83.4	13.5	0.78	2.33	17
450	76.7	18.9	1.37	3.06	14

10

EXAMPLE 2

A Pd-Ni/ γ -Al₂O₃ catalyst containing 2.0 wt.% Pd
15 and the requisite loading of Ni to give a Pd/Ni atomic ratio of 1 was prepared by incipient wetness impregnation of a γ -alumina support (20-40 mesh) with an aqueous solution of palladium chloride and nickel nitrate hexahydrate, followed by drying at 120°C.

20

A microreactor was packed with an admixture of 0.2 gram of the above catalyst with 0.3 gram of inert γ -Al₂O₃ and placed in a continuous flow system. Prior to the start of the reaction, the mixture of Pd-Ni/ γ -Al₂O₃
25 catalyst and inert γ -Al₂O₃ was calcined at 500°C under N₂ flow for 2 hours, followed by a reduction at 500°C

under hydrogen flow for 2 hours. A feed stream consisting of a mixture of CH₄ (43.9 vol.%), water vapor (12.5 vol.%) and N₂ (43.6 vol.%) was passed over the catalyst bed; the CH₄/H₂O feed ratio was thus about 3.5. The nitrogen added to the feed acted as an internal standard. The flow of dry gas at the reactor inlet was maintained approximately at 25 ml/min., the pressure in the reactor being maintained at about 1 atmosphere. The reaction temperature (bed temperature) was varied from 400 to 450°C. The reactor effluent stream was analyzed for reactants and products (N₂, CH₄, CO, CO₂, H₂O and H₂) by TCD - gas chromatography. The composition of the reactor effluent (dry gas basis excluding internal standard) for the selected temperatures is reported in Table 2.

TABLE 2

Reaction Temperature (°C)	Reactor effluent composition (vol. %)				H ₂ /CO (v/v) ratio in effluent
	CH ₄	H ₂	CO	CO ₂	
400	92.0	6.5	0.21	1.24	31
425	83.2	13.5	0.56	2.51	24
450	76.6	18.9	1.14	3.29	17

20

EXAMPLE 3

A Pd-Cu/ γ -Al₂O₃ catalyst containing 2.0 wt.% Pd and the requisite loading of Cu to give a Pd/Cu atomic ratio of 1 was prepared by incipient wetness impregnation of a γ -alumina support (20-40 mesh) with

25

an aqueous solution of palladium chloride and copper (+2) nitrate trihydrate, followed by drying at 120°C.

A microreactor was packed with an admixture of 0.2
5 gram of the above catalyst with 0.3 gram of inert γ - Al_2O_3 and placed in a continuous flow system. Prior to the start of the reaction, the mixture of Pd-Cu/ γ - Al_2O_3 catalyst and inert γ - Al_2O_3 was calcined at 500°C under N_2 flow for 2 hours, followed by a reduction at 500°C
10 under hydrogen flow for 2 hours. A feed stream consisting of a mixture of CH_4 (43.9 vol.%), water vapor (12.5 vol.%) and N_2 (43.6 vol.%) was passed over the catalyst bed; the $\text{CH}_4/\text{H}_2\text{O}$ feed ratio was thus about 3.5. The nitrogen added to the feed acted as an
15 internal standard. The flow of dry gas at the reactor inlet was maintained approximately at 25 ml/min., the pressure in the reactor being maintained at about 1 atmosphere. The reaction temperature (bed temperature) was varied from 450 to 500°C. The reactor effluent
20 stream was analyzed for reactants and products (N_2 , CH_4 , CO , CO_2 , H_2O and H_2) by TCD - gas chromatography. The composition of the reactor effluent (dry gas basis excluding internal standard) for the selected temperatures is reported in Table 3.

TABLE 3

Reaction Temperature (°C)	Reactor effluent composition (vol. %)				H ₂ /CO (v/v) ratio in effluent
	CH ₄	H ₂	CO	CO ₂	
450	94.6	4.5	0.31	0.51	15
475	87.1	10.3	0.91	1.53	11
500	79.2	16.9	1.50	2.43	11

5

EXAMPLE 4

A Pd-Mg/ γ -Al₂O₃ catalyst containing 2.0 wt.% Pd and the requisite loading of Mg to give a Pd/Mg atomic ratio of 1 was prepared by incipient wetness impregnation of a γ -alumina support (20-40 mesh) with an aqueous solution of palladium chloride and magnesium nitrate hexahydrate, followed by drying at 120°C.

15 A microreactor was packed with an admixture of 0.2 gram of the above catalyst with 0.3 gram of inert γ -Al₂O₃ and placed in a continuous flow system. Prior to the start of the reaction, the mixture of Pd-Mg/ γ -Al₂O₃ catalyst and inert γ -Al₂O₃ was calcined at 500°C under
20 N₂ flow for 2 hours, followed by a reduction at 500°C under hydrogen flow for 2 hours. A feed stream consisting of a mixture of CH₄ (43.9 vol.%), water vapor (12.5 vol.%) and N₂ (43.6 vol.%) was passed over the catalyst bed; the CH₄/H₂O feed ratio was thus about
25 3.5. The nitrogen added to the feed acted as an internal standard. The flow of dry gas at the reactor

inlet was maintained approximately at 25 ml/min., the pressure in the reactor being maintained at about 1 atmosphere. The reaction temperature (bed temperature) was varied from 400 to 450°C. The reactor effluent stream was analyzed for reactants and products (N₂, CH₄, CO, CO₂, H₂O and H₂) by TCD - gas chromatography. The composition of the reactor effluent (dry gas basis excluding internal standard) for the selected temperatures is reported in Table 4.

10

TABLE 4

Reaction Temperature (°C)	Reactor effluent composition (vol. %)				H ₂ /CO (v/v) ratio in effluent
	CH ₄	H ₂	CO	CO ₂	
400	91.2	7.2	0.86	0.71	8.4
425	85.8	11.6	1.32	1.38	8.8
450	80.0	16.1	1.80	2.04	8.9

15 EXAMPLE 5

A Pd-La/ γ -Al₂O₃ catalyst containing 2.0 wt.% Pd and the requisite loading of La to give a Pd/La atomic ratio of 1 was prepared by incipient wetness impregnation of a γ -alumina support (20-40 mesh) with an aqueous solution of palladium chloride and lanthanum nitrate hexahydrate, followed by drying at 120°C.

A microreactor was packed with an admixture of 0.2 gram of the above catalyst with 0.3 gram of inert γ -Al₂O₃ and placed in a continuous flow system. Prior to

the start of the reaction, the mixture of Pd-La/ γ -Al₂O₃ catalyst and inert γ -Al₂O₃ was calcined at 500°C under N₂ flow for 2 hours, followed by a reduction at 500°C under hydrogen flow for 2 hours. A feed stream
5 consisting of a mixture of CH₄ (43.9 vol.%), water vapor (12.5 vol.%) and N₂ (43.6 vol.%) was passed over the catalyst bed; the CH₄/H₂O feed ratio was thus about 3.5. The nitrogen added to the feed acted as an internal standard. The flow of dry gas at the reactor
10 inlet was maintained approximately at 25 ml/min., the pressure in the reactor being maintained at about 1 atmosphere. The reaction temperature (bed temperature) was varied from 400 to 450°C. The reactor effluent stream was analyzed for reactants and products (N₂,
15 CH₄, CO, CO₂, H₂O and H₂) by TCD - gas chromatography. The composition of the reactor effluent (dry gas basis excluding internal standard) for the selected temperatures is reported in Table 5.

20

TABLE 5

Reaction Temperature (°C)	Reactor effluent composition (vol. %)				H ₂ /CO (v/v) ratio in effluent
	CH ₄	H ₂	CO	CO ₂	
400	90.4	7.9	0.77	0.92	10
425	82.0	14.8	0.96	2.24	15
450	76.4	19.0	1.41	3.19	13

EXAMPLE 6

5 A Pd-Ce/ γ -Al₂O₃ catalyst containing 2.0 wt.% Pd and the requisite of Ce to give a Pd/Ce atomic ratio of 1 was prepared by incipient wetness impregnation of a γ -alumina support (20-40 mesh) with an aqueous solution of palladium chloride and cerium nitrate hexahydrate, followed by drying at 120°C.

10

A microreactor was packed with an admixture of 0.2 gram of the above catalyst with 0.3 gram of inert γ -Al₂O₃ and placed in a continuous flow system. Prior to the start of the reaction, the mixture of Pd-Ce/ γ -Al₂O₃ catalyst and inert γ -Al₂O₃ was calcined at 500°C under N₂ flow for 2 hours, followed by a reduction at 500°C under hydrogen flow for 2 hours. A feed stream consisting of a mixture of CH₄ (43.9 vol.%), water vapor (12.5 vol.%) and N₂ (43.6 vol.%) was passed over the catalyst bed; the CH₄/H₂O feed ratio was thus about 3.5. The nitrogen added to the feed acted as an internal standard. The flow of dry gas at the reactor inlet was maintained approximately at 25 ml/min., the pressure in the reactor being maintained at about 1 atmosphere. The reaction temperature (bed temperature) was varied from 400 to 450°C. The reactor effluent stream was analyzed for reactants and products (N₂, CH₄, CO, CO₂, H₂O and H₂) by TCD - gas chromatography. The composition of the reactor effluent (dry gas basis excluding internal standard) for the selected temperatures is reported in Table 6.

TABLE 6

Reaction Temperature (°C)	Reactor effluent composition (vol. %)				H ₂ /CO (v/v) ratio in effluent
	CH ₄	H ₂	CO	CO ₂	
400	88.9	9.2	0.59	1.32	16
425	80.9	15.6	0.90	2.52	17
450	74.7	20.6	1.42	3.30	15

CLAIMS:

1. A process for the production of hythane, which comprises oxidizing methane with water vapor at a
5 temperature of about 400 to about 500°C and a pressure of about 1 to about 5 atmospheres, in the presence of a catalyst comprising palladium or a binary alloy thereof with another metal selected from the group consisting of elements belonging to Group IB, Group IIA, Group
10 VIII and the lanthanide series of the Periodic Table, the catalyst being supported on a porous carrier.
2. A process as claimed in claim 1, wherein said oxidation is carried out at atmospheric pressure.
15
3. A process as claimed in claim 1 or 2, wherein said oxidation is carried out at a temperature ranging from about 400 to about 450°C.
- 20 4. A process as claimed in claim 1, wherein said catalyst is selected from the group consisting of palladium, alloys of palladium-nickel, palladium-copper, palladium-magnesium, palladium-lanthanum and palladium-cerium.
25
5. A process as claimed in claim 1 or 4, wherein said carrier is selected from the group consisting of alumina, silica and zirconia.
- 30 6. A process as claimed in claim 1, wherein said catalyst comprises palladium supported on γ -alumina.

7. A process as claimed in claim 1 or 4, wherein said catalyst comprises about 0.5 to about 10% by weight of palladium or of said binary alloy thereof, based on the total weight of the supported catalyst.

5

8. A process as claimed in claim 1, wherein said catalyst comprises about 2% by weight of palladium, based on the total weight of the supported catalyst.

10 9. A process as claimed in claim 4, wherein said catalyst comprises an alloy of palladium with another metal selected from the group consisting of nickel, copper, magnesium, lanthanum and cerium, in which palladium is present in an amount of about 2% by
15 weight, based on the total weight of the supported catalyst, and wherein said alloy has a palladium/other metal atomic ratio of about 1.

20 10. A process as claimed in claim 1, wherein the supported catalyst is used in admixture with an inert porous material.

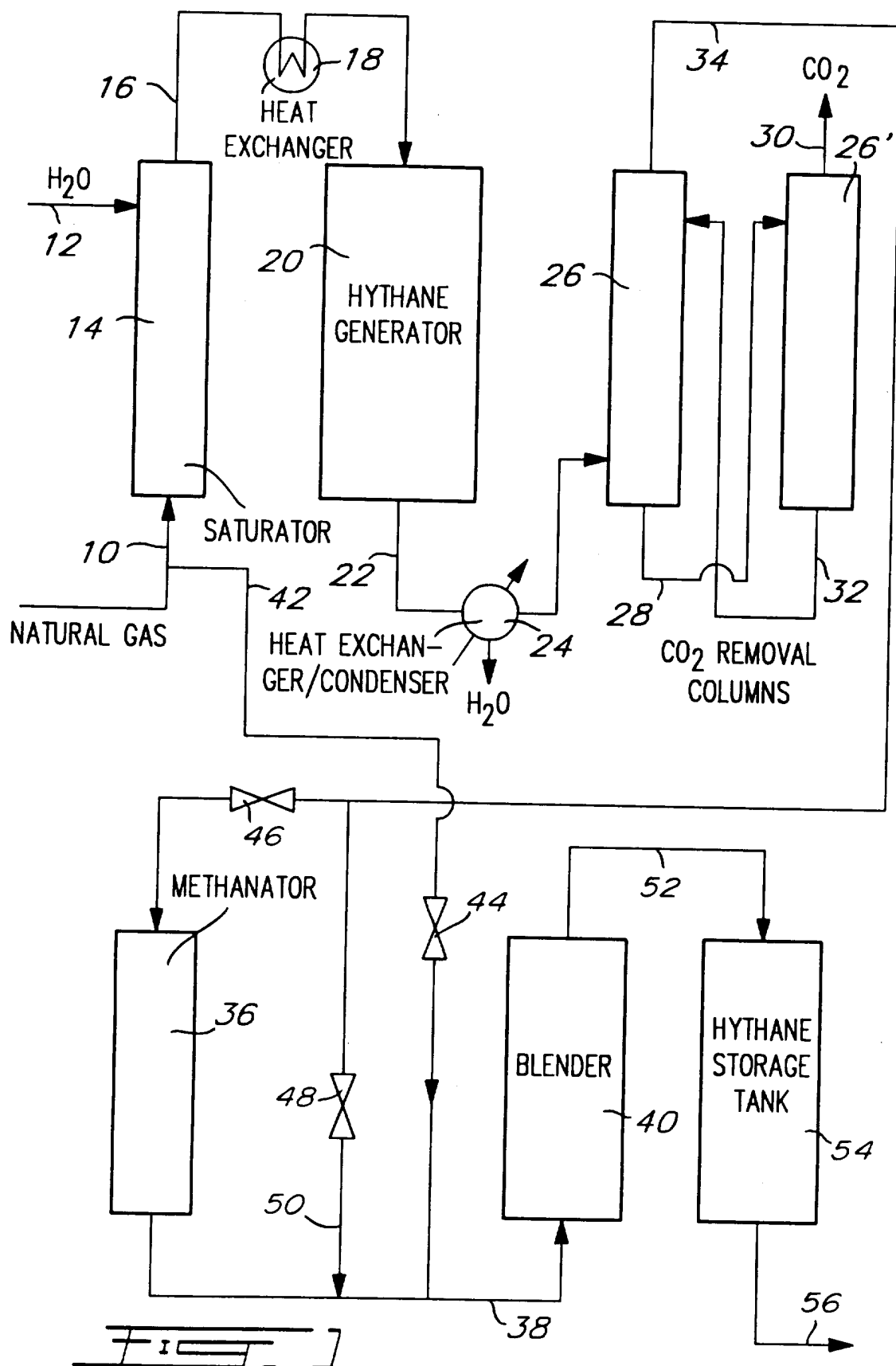
11. A process as claimed in claim 10, wherein said inert porous material comprises γ -alumina.

25

12. A process as claimed in claim 1, wherein said oxidation is carried out at atmospheric pressure with a $\text{CH}_4/\text{H}_2\text{O}$ feed ratio ranging from about 2 to about 5, a dry gas feed flowrate ranging from about 20 to about
30 100 ml/min., and a mass of supported catalyst ranging from about 0.2 to about 1 g.

13. A process as claimed in claim 12, wherein said $\text{CH}_4/\text{H}_2\text{O}$ feed ratio is about 3.5.
14. A process as claimed in claim 12, wherein said dry
5 gas feed flowrate is about 25 ml/min.
15. A process as claimed in claim 12, wherein said mass of supported catalyst is about 0.2 g.
- 10 16. A process as claimed in claim 12, wherein said supported catalyst is used in admixture with an inert porous material.
- 15 17. A process as claimed in claim 16, wherein said inert porous material comprises γ -alumina.
18. A process as claimed in claim 16 or 17, wherein said inert porous material is used in an amount of about one to two-fold the mass of supported catalyst.
- 20 19. A process as claimed in claim 12, 13, 14, 15, 16 or 17, wherein said oxidation is carried out at a temperature ranging from about 400 to about 500°C.
- 25 20. A process as claimed in claim 12, 13, 14, 15, 16 or 17, wherein said catalyst comprises palladium supported on γ -alumina.

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INTERNATIONAL SEARCH REPORT

International application No.

PCT/CA 96/00048

A. CLASSIFICATION OF SUBJECT MATTER

IPC6: C01B 3/38 // B01J 23/44, B01J 23/54, B01J 23/56

According to International Patent Classification (IPC) or to both national classification and IPC

B. FIELDS SEARCHED

Minimum documentation searched (classification system followed by classification symbols)

IPC6: C01B

Documentation searched other than minimum documentation to the extent that such documents are included in the fields searched

Electronic data base consulted during the international search (name of data base and, where practicable, search terms used)

EPODOC, WPIL, PAJ, IFIPAT

C. DOCUMENTS CONSIDERED TO BE RELEVANT

Category*	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
Y	EP, A2, 0212758 (SHELL INTERNATIONALE RESEARCH MAATSCHAPPIJ B.V.), 4 March 1987 (04.03.87), page 2, column 1, line 20 - column 1, line 26 --	1-20
Y	EP, A2, 0112613 (ENGELHARD CORPORATION), 4 July 1984 (04.07.84), page 5, line 4 - line 14; page 13, line 14 - line 18; page 1, line 17 - line 20 --	1-20
A	EP, A2, 0212755 (SHELL INTERNATIONALE RESEARCH MAATSCHAPPIJ B.V.), 4 March 1987 (04.03.87), abstract -- -----	1-20

☐ Further documents are listed in the continuation of Box C. ☒ See patent family annex.

* Special categories of cited documents:

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"X" document of particular relevance: the claimed invention cannot be considered novel or cannot be considered to involve an inventive step when the document is taken alone

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Date of the actual completion of the international search

26 April 1996

Date of mailing of the international search report

07.06.96

Name and mailing address of the International Searching Authority

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INTERNATIONAL SEARCH REPORT

Information on patent family members

01/04/96

International application No.

PCT/CA 96/00048

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