(12) INTERNATIONAL APPLICATION PUBLISHED UNDER THE PATENT COOPERATION TREATY (PCT)

(19) World Intellectual Property Organization
International Bureau

(43) International Publication Date
20 September 2018 (20.09.2018)

WIPO | PCT

(10) International Publication Number
WO 2018/166873 A1

(51) International Patent Classification:
C07C 273/04 (2006.01) *C07C 29/15* (2006.01)
C07C 273/10 (2006.01) *C07C 29/151* (2006.01)
C01C 1/04 (2006.01) *C01B 3/02* (2006.01)

(21) International Application Number:
PCT/EP2018/055648

(22) International Filing Date:
07 March 2018 (07.03.2018)

(25) Filing Language: English

(26) Publication Language: English

(30) Priority Data:
PA 2017 00172 12 March 2017 (12.03.2017) DK

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(81) Designated States *(unless otherwise indicated, for every kind of national protection available)*: AE, AG, AL, AM, AO, AT, AU, AZ, BA, BB, BG, BH, BN, BR, BW, BY, BZ, CA, CH, CL, CN, CO, CR, CU, CZ, DE, DJ, DK, DM, DO, DZ, EC, EE, EG, ES, FI, GB, GD, GE, GH, GM, GT, HN, HR, HU, ID, IL, IN, IR, IS, JO, JP, KE, KG, KH, KN, KP, KR, KW, KZ, LA, LC, LK, LR, LS, LU, LY, MA, MD, ME, MG, MK, MN, MW, MX, MY, MZ, NA, NG, NI, NO, NZ, OM, PA, PE, PG, PH, PL, PT, QA, RO, RS, RU, RW, SA, SC, SD, SE, SG, SK, SL, SM, ST, SV, SY, TH, TJ, TM, TN, TR, TT, TZ, UA, UG, US, UZ, VC, VN, ZA, ZM, ZW.

(84) Designated States *(unless otherwise indicated, for every kind of regional protection available)*: ARIPO (BW, GH, GM, KE, LR, LS, MW, MZ, NA, RW, SD, SL, ST, SZ, TZ, UG, ZM, ZW), Eurasian (AM, AZ, BY, KG, KZ, RU, TJ, TM), European (AL, AT, BE, BG, CH, CY, CZ, DE, DK, EE, ES, FI, FR, GB, GR, HR, HU, IE, IS, IT, LT, LU, LV, MC, MK, MT, NL, NO, PL, PT, RO, RS, SE, SI, SK, SM, TR), OAPI (BF, BJ, CF, CG, CI, CM, GA, GN, GQ, GW, KM, ML, MR, NE, SN, TD, TG).

(54) Title: CO-PRODUCTION OF METHANOL, AMMONIA AND UREA

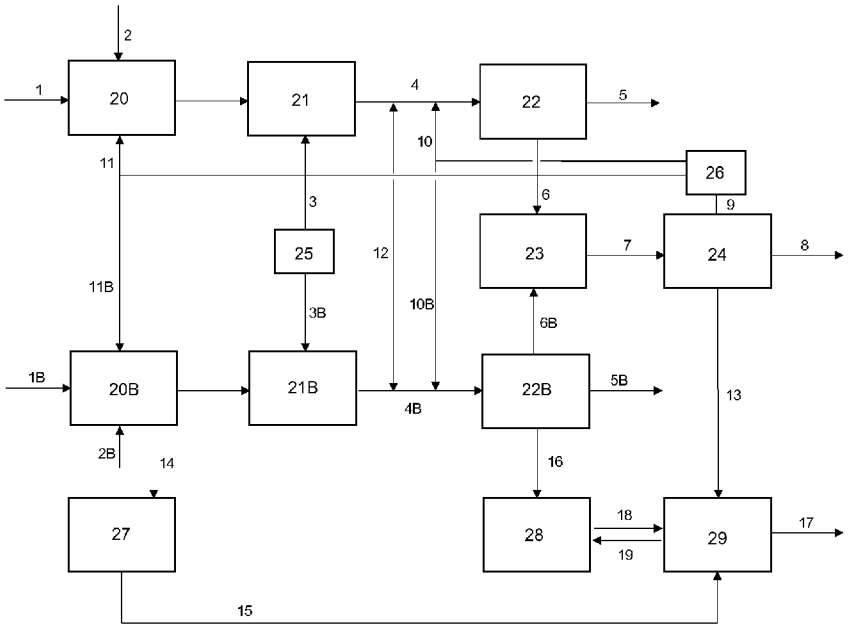


Fig. 1

(57) Abstract: Process for the co-production of methanol and ammonia together with urea production from a hydrocarbon feed without venting to the atmosphere carbon dioxide captured from the methanol or ammonia synthesis gas and without using expensive air separation units and water gas shift. Carbon dioxide removal from flue gas from reforming section to convert partially or fully all ammonia into urea. Synergi of having methanol, ammonia and urea production to produce coating material for the urea production.

[Continued on next page]

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Declarations under Rule 4.17:

— *as to the identity of the inventor (Rule 4.17(i))*
— *as to applicant's entitlement to apply for and be granted a patent (Rule 4.17(ii))*
— *as to the applicant's entitlement to claim the priority of the earlier application (Rule 4.17(iii))*
— *of inventorship (Rule 4.17(iv))*

Published:

— *with international search report (Art. 21(3))*

Title: CO-PRODUCTION OF METHANOL, AMMONIA AND UREA

The present invention relates to a process for the co-production of methanol and ammonia from a hydrocarbon feed without venting to the atmosphere carbon dioxide captured from the methanol and ammonia synthesis gas and without using expensive water gas shift and carbon dioxide removal steps. More particularly the invention is concerned with a sequential and once-through (single pass) process for the co-production of methanol and ammonia without water gas shift and carbon dioxide removal and without air separation unit used in the reforming section of the plant.

Current processes for co-production of methanol and ammonia involve generally parallel processes in which a common reforming section is used to generate a synthesis gas which is split in separate parallel streams, one of which is used for methanol synthesis and the other for ammonia synthesis. The co-production of methanol and ammonia can also be conducted sequentially or in series, where the synthesis gas produced in the reforming section is first converted to methanol and the unreacted gas containing carbon oxides and hydrogen is subsequently used for ammonia synthesis. Water gas shift and/or carbon dioxide removal steps of the synthesis gas stream are required, thus involving the release of CO_2 to the atmosphere and the investment in highly expensive and complicated units for conducting the shift conversion and carbon dioxide removal.

In a first aspect of the present invention is provided a process for co-producing methanol, ammonia and urea which is simpler than current process and which at the same time

enables minimum release of carbon dioxide to the atmosphere.

In a second aspect of the present invention is provided a process for co-producing methanol, ammonia and urea which allows the optimized use of two parallel methanol production lines.

These and other advantages are achieved by the present process for co-producing methanol, ammonia and urea from a hydrocarbon feedstock, said process comprising the steps of

a) from a first methanol processes, comprising a first reforming step and first methanol conversion step obtaining a first effluent comprising methanol and a first gas effluent comprising nitrogen, hydrogen and unconverted carbon oxides, and
from a second methanol processes, comprising a second reforming step and a second methanol conversion step obtaining a second effluent comprising methanol and a second gas effluent comprising nitrogen, hydrogen and unconverted carbon oxides

b) producing an ammonia synthesis gas from the first and/or second gas effluent comprising nitrogen, hydrogen and unconverted carbon oxides in a common catalytic methanation stage and withdrawing said ammonia synthesis gas preferably having a H2:N2 molar ratio of 3:1;

c) catalytically converting the nitrogen and hydrogen of the ammonia synthesis gas in a common ammonia synthesis stage and withdrawing an effluent comprising ammonia and an off-gas stream comprising hydrogen, nitrogen and methane, and

d) reacting at least part of the effluent comprising ammonia with at least part of a flue gas comprising CO2 from at least one of the first and second reforming steps to produce urea.

In other words, the applicant by the present invention provides a process based on two parallel methanol production lines, each comprising a reforming section and a methanol reaction section. Each methanol reaction section providing methanol product which may be send for further processing, storage etc. as well as a gas effluent comprising nitrogen, hydrogen and unconverted carbon oxides which is highly suitable as basis for an ammonia synthesis upon methanation. Furthermore, the aspect of CO2 emission from industrial plants and processes are of general concern rendering the current process of great interest as at least part of CO2 from flue gas from the reforming sections is used to produce urea.

As used herein the term "carbon oxides" means the components carbon monoxide and carbon dioxide.

As used herein, in the catalytic methanation of step (b) the term "by removing the unconverted carbon oxides" means converting the unconverted carbon oxides to methane. This is evidently different from carbon dioxide removal through the use of absorbers in acid gas washes, which the present invention eliminates.

Accordingly, as used herein the term "carbon dioxide removal" means highly expensive CO2-removal stages in the form

of acid gas wash, such as conventional MDEA and carbonate wash processes.

Preferably in each of the first and second methanol processes the following steps are carried out

1) producing a methanol synthesis gas containing hydrogen, carbon oxides and nitrogen by steam reforming the hydrocarbon feedstock in a reforming section comprising a primary and a secondary reforming stage;

2) catalytically converting the carbon oxides and hydrogen of the methanol synthesis gas in a once-through methanol synthesis stage and withdrawing an effluent comprising methanol and a gas effluent comprising nitrogen, hydrogen and unconverted carbon oxides;

In various setups the secondary reforming stage is an air-blown secondary reforming stage.

As used herein the term "primary reforming stage" means reforming being conducted in a conventional steam methane reformer (SMR), i.e. tubular reformer with the heat required for the endothermic reforming being provided by radiation heat from burners, such as burners arranged along the walls of the tubular reformer.

As used herein the term "air-blown secondary reforming stage" means reforming being conducted in an autothermal reformer or catalytic partial oxidation reactor using air.

As used herein, the term "once-through methanol synthesis stage" means that methanol is produced in at least one catalytic reactor operating in a single pass configuration, i.e. without significant recirculation (not more than 5%,

i.e. less than 5%, often 0%) of the volume flow of any gas produced in the methanol synthesis back to the at least one methanol reactor of the methanol synthesis stage, particularly the gas effluent containing hydrogen and unconverted carbon oxides.

By providing the two parallel methanol production lines and a common methanation step followed by a common ammonia synthesis a flexible but yet very simple process is provided.

Being able to control the flow/feed of the synthesis gas from the each of the two reforming sections to each of the two methanol synthesis steps is advantageous for several reasons.

For example, the first and second once-through methanol synthesis step of the first and the second methanol processes are controlled at least by controlling the amount of syngas from the first and/or second reforming step fed to the first and/or second methanol production step.

The first and the second methanol processes may be interconnected by a synthesis gas line allowing the synthesis gas from each of the two first and second reforming steps to be distributed to each of the first and second methanol steps which for example may minimize any flaring of synthesis gas to optimize the energy consumption and thereby the operational expenditure (OPEX) and/or maximize the use of an often highly expensive reforming section by utilizing the margins in compressor design and voluntarily different deactivation cycle in the first and second methanol step.

The process may comprise further parallel methanol processes. I.e. one or more additional methanol processes may be worked in the parallel with the twin methanol process of the invention. The parallel one, two, three or more parallel methanol processes may be interconnected by one or more synthesis gas line.

In some embodiments at least a part of the methanol from the first and/or second methanol process is converted into formaldehyde. This formaldehyde or at least part thereof may be absorbed into aqueous urea solution based on at least part of the urea to produce UFC85 or equivalent solutions. The UFC85 may be added to the urea as a coating material. UFC85 is a concentrate comprising 85% formaldehyde + urea and 15% water.

The present application further relates to a plant for co-production of Methanol, ammonia and Urea.

A plant comprises a first and a second parallel methanol process lines, at least one common methanation reactor and at least one common ammonia synthesis step as well as a urea production section. Furthermore, the plant may advantageously comprise an interconnecting synthesis line allowing fluid communication between the first and second methanol process line.

The plant may also comprise a formaldehyde production section and/or a UFC85 production section.

With the simple and low cost process lay-out (as in EP 2192082 B1) one can afford two lines of methanol, where

each one is considered large size, and gain the flexibility that comes with having two lines instead of having one big line.

Typically, the reforming section would be the most costly in terms of capital expenditure (CAPEX) and therefore very limited oversizing of this section would take place. Whereas the synthesis compressor would typically have a margin according to the relevant code and standard, and the methanol synthesis would have extra capacity when the catalyst is new and not yet deactivated. This can be utilized to gain flexibility and to reduce natural gas savings throughout the lifetime of the methanol catalyst lifetime that would be typical designed for 3-4 years. The ammonia synthesis catalyst deactivates much slower and can typically last for 20 years and will therefore not follow the catalyst lifetime cycle as the methanol synthesis catalyst. In principle, the common ammonia synthesis can be kept in operation for the whole lifetime of its catalyst by always being fed with synthesis gas from at least one of the methanol synthesis.

Further flexibility can be obtained by having an interconnecting synthesis gas line between the two front ends (synthesis gas preparation). Most likely the two Methanol synthesis would be operated in a way so they would not have catalyst replacement simultaneously. One methanol synthesis can have end-of-run (EOR) catalyst condition while the other is still in near start-of-run condition and still the total methanol synthesis capacity will be sufficient to utilize the produced synthesis gas.

This synthesis gas line can e.g. be used during start-up of the first methanol process while the second methanol process is in operation by feeding the synthesis gas from the first methanol process being started up to the second methanol process instead of flaring since the first methanol synthesis due to the start-up sequence would not be available for operation.

With methanol processes and interconnecting synthesis gas line, the costly reforming section in both methanol processes can be operated at its maximum capacity and the two methanol synthesis and one common ammonia synthesis can together convert all the synthesis gas being produced.

An Ammonia synthesis loop having a capacity of 750-960 MTPD is considered as small so the most obvious advantages is to have one common ammonia synthesis line to convert the synthesis gas from two methanol synthesis lines. The total ammonia capacity would then add up to a near world class ammonia plant capacity of 1500-1920 and considerable CAPEX savings can be achieved.

Everything else that is related to the ammonia synthesis can also be in common for the two methanol lines such as (and not limited to) process air compressor, methanation section, hydrogen recovery section and ammonia recovery from purge gas. At least same relative CAPEX savings can be achieved for these sections that are considered small.

The Urea and UFC85 process steps described herein may also be used in relation with a single line methanol and ammonia co-production process such as a process/plant basically

based on a single of the herein described methanol processes together with a methanation and ammonia syntheses. For example, a co-production process as described in EP 2192082 B1.

Preferably, the hydrocarbon feedstock is natural gas, for instance in the form of liquified natural gas (LNG) or substitute natural gas (SNG).

By the invention we make direct use of the reactions governing reforming, methanol synthesis and ammonia synthesis so that methanol and ammonia can be co-produced without venting carbon dioxide being captured from the synthesis gas. The production of hydrogen by steam reforming is governed by the endothermic reaction $CH_4 + H_2O = CO + 3\ H_2$, while methanol synthesis in the absence of carbon dioxide is governed by the reaction $CO + 2\ H_2 = CH_3OH$. In the presence of carbon dioxide, methanol is otherwise also generated according to the reaction $CO_2 + 3\ H_2 = CH_3OH + H_2O$. Ideally the feed synthesis gas for methanol production is a gas containing the highest possible molar ratio CO/CO_2. Ammonia synthesis occurs according to the reaction $N_2 + 3\ H_2 = 2\ NH_3$. Since when conducting the overall process, the reforming only produces 3 moles of hydrogen, while methanol synthesis already takes 2 moles of hydrogen and ammonia synthesis requires 3 moles of hydrogen, we restrict on purpose the amount of ammonia to be produced to a third in order to be able to utilize the hydrogen that is available according to 1/3 ($N_2 + 3\ H_2 = 2\ NH_3$). Hence, in a way, by the invention we purposively promote a minimum of flexibility in the product split of methanol and ammonia.

This simple and elegant measure enables the production of about 75-80 wt% methanol and 20-25 wt% ammonia at any time in a process which is simpler and less costly than conventional ones because the process obviates the need to use highly expensive water gas shift stages for the conversion of carbon monoxide into hydrogen and carbon dioxide and also obviates the need to use highly expensive CO_2-removal stages, i.e. acid gas wash, such as the conventional MDEA and carbonate wash processes. Operating costs are also kept at minimum since there is no need for shift catalyst replacement and no need for solvent replenishment in the CO_2-removal processes. This contrasts other combined processes for the production of methanol and ammonia, such as that of JP 2000063115 where highly expensive removal of carbon dioxide via conventional CO_2 stripper or absorber is necessary in order to adjust the CO2/CO ratio in the synthesis gas and thereby achieve flexibility in the process. In addition, since the secondary reforming is conducted in an air-blown secondary reformer (air-blown autothermal reformer) in order to provide for the required nitrogen there is no need for expensive and massive Air Separation Units (ASU), thereby also making the process less costly than current processes where ASU plants are often required for oxygen supply in autothermal reformers and in which the attendant nitrogen production is normally used in a subsequent nitrogen wash.

The process of the present invention is environmentally friendly because there are no emissions to the surroundings of the CO_2 captured from the methanol and ammonia synthesis gas. Practically all carbon monoxide (and carbon dioxide) produced in the process is used for methanol synthesis.

The process is applicable for any plant capacity including large plants producing more than 2000 MTPD ammonia and methanol, for instance 3000, 5000 MTPD or even more.

The methanol synthesis stage is preferably conducted by conventional means by passing the synthesis gas at high pressure and temperatures, such as 60-150 bars, preferably 120 bars and 150-300°C through at least one methanol reactor containing at least one fixed bed of methanol catalyst. A particularly preferred methanol reactor is a fixed bed reactor cooled by a suitable cooling agent such as boiling water, e.g. boiling water reactor (BWR). In a specific embodiment the methanol synthesis stage in step (b) is conducted by passing the synthesis gas through one boiling water reactor and subsequently through an adiabatic fixed bed reactor, or by passing the synthesis gas through a series of boiling water reactors and subsequently through an adiabatic fixed bed reactor. Preferably the boiling water reactor is in the form of a single reactor of the condensing-methanol type which comprises within a common shell a fixed bed of methanol catalyst particles and cooling means adapted to indirectly cooling the methanol synthesis gas with a cooling agent, and which preferably operates at pressures above 90 bar and below 150 bar, more preferably above 110 bar and below 130 bar, as described in our DK patent applications PA 2008 00261 and PA 2008 00260 filed 25 February 2008. The use of a methanol reactor according to these applications enables operation at pressures much higher than conventional boiling reactors which normally are about 80 bars. In addition, it enables the use of a single reactor rather than two conventional boiling water reactors, thereby significantly reducing plant costs. Fur-

thermore, since the operating pressure in the methanol synthesis stage can be kept as high as about 120 bars or even higher there are significant savings in terms of equipment size and overall investment costs as methanol synthesis is favoured at high pressures.

Accordingly, the invention enables the operation of the methanol and ammonia synthesis section at similar operating pressures, for instance 130 bars, which implies a simplified process with significant savings in size of equipment as mentioned above. Yet it is also possible to operate at two different operating pressures, for instance 80 bars in the methanol synthesis stage and 130 bar in the ammonia synthesis stage, which implies energy savings in the methanol synthesis stage.

In step (a) the effluent containing methanol is preferably a liquid effluent. This effluent is obtained by cooling and condensation of the synthesis gas from the methanol reactors. Accordingly the process of the invention may further comprise cooling the synthesis gas withdrawn from each methanol reactor to condense methanol and passing the gas through a separator, withdrawing a bottom fraction from the separator containing the raw methanol, withdrawing an overhead fraction containing synthesis gas which is passed to the subsequent methanol reactor, and forming a single liquid effluent containing methanol by combining the bottom fractions of the separators of each reactor containing the raw methanol.

It would be understood that the term “methanol reactor” as used herein encompasses adiabatic fixed bed reactors and

cooled reactors such as boiling water reactors and reactors of the condensing-methanol type which comprises within a common shell a fixed bed of methanol catalyst particles and cooling means adapted to indirectly cooling the methanol synthesis gas with a cooling agent.

Since the methanol synthesis stage is once-through, there is no need for recirculation of a part of the overhead fraction from the separator of the adiabatic fixed bed reactor back to the first methanol reactor of the methanol synthesis stage. This contrasts other combined processes for the production of methanol and ammonia, such as that of JP 2000063115 where methanol synthesis involves significant recirculation of product gas.

In step (b) the catalytic methanation stage for conversion of carbon oxides to methane is conducted in at least one methanation reactor, which is preferably an adiabatic reactor containing a fixed bed of methanation catalyst.

In step (c) the ammonia synthesis gas from the methanation stage containing the right proportion of hydrogen and nitrogen (preferably H_2:N_2 molar ratio of 3:1) is optionally passed through a compressor to obtain the required ammonia synthesis pressure, such as 120 to 200 bar, preferably about 130 bar. Ammonia is then produced in a conventional manner by means of an ammonia synthesis loop comprising at least one ammonia converter containing at least one fixed bed of ammonia catalyst, with interbed cooling. The effluent containing ammonia contains also hydrogen, nitrogen and inerts such as methane and argon. Ammonia may be recovered from the effluent containing ammonia as liquid ammonia by

condensation and subsequent separation. Preferably, an off-gas stream containing hydrogen, nitrogen and methane is withdrawn from the ammonia synthesis stage, as also is a hydrogen-rich stream (> 90 vol% H_2). These streams may for instance stem from a purge gas recovery unit. Preferably, this hydrogen stream is added to the methanol synthesis stage, for instance by combining with the methanol synthesis gas. The recycle of this hydrogen-rich stream enables a higher efficiency in the process as useful hydrogen is utilised in the methanol synthesis and subsequent ammonia synthesis rather than simply being used as fuel.

In order to improve the energy efficiency of the process the off-gas (purge) stream containing hydrogen, nitrogen and methane of step (c) is returned to step (a), i.e. it is returned as off-gas fuel to the reforming section of the plant, specifically to the primary reforming stage.

Plant capacity of 3000 MTPD Methanol and 750 MTPD ammonia in co-production is considered to be world class size, but even higher plant capacities are requested. By the present invention, we provide a twin methanol line plant having capacity of 2 x 3000 MTPD Methanol and 2 x 750 MTPD ammonia which can be competitive against other technologies usually considered more optimal for production of synthesis gas for methanol synthesis; such as two-step reforming or ATR.

The accompanying figure shows a simplified block diagram of the process according to a specific embodiment of the invention including reforming, methanol synthesis stage, methanation stage and ammonia synthesis stage.

In first reforming section, natural gas 1 is added to primary reforming stage 20 (steam methane reformer) under addition of steam 2. The partly reformed gas is then further reformed in air-blown secondary reforming stage 21 (autothermal reformer) under addition of air 3. The methanol synthesis gas 4 containing hydrogen, carbon oxides and nitrogen is cooled in waste heat boiler(s) under the production of steam and then compressed to methanol synthesis pressure (not shown). In first methanol synthesis stage 22 the methanol synthesis gas 4 is converted in once-through operation (single-pass operation, no recirculation) under the production of a liquid effluent 5 containing methanol and a gas effluent 6 containing nitrogen, hydrogen and unconverted carbon oxides. Approximately 80 wt% of the total plant capacity goes to the production of methanol of effluent 5.

In second reforming section, natural gas 1B is added to primary reforming stage 20B (steam methane reformer) under addition of steam 2B. The partly reformed gas is then further reformed in air-blown secondary reforming stage 21B (autothermal reformer) under addition of air 3B. The methanol synthesis gas 4B containing hydrogen, carbon oxides and nitrogen is cooled in waste heat boiler(s) under the production of steam and then compressed to methanol synthesis pressure (not shown). In second methanol synthesis stage 22B the methanol synthesis gas 4B is converted in once-through operation (single-pass operation, no recirculation) under the production of a liquid effluent 5B containing methanol and a gas effluent 6B containing nitrogen, hydrogen and unconverted carbon oxides. Approximately 80 wt% of

the total plant capacity goes to the production of methanol of effluent 5B.

The carbon oxides in gas effluents 6 and 6B are hydrogenated to methane in the common methanation stage 23 thereby generating an ammonia synthesis gas 7 having a H_2:N_2 molar ratio of 3:1. The ammonia synthesis gas 7 is then passed through ammonia synthesis stage 24 under the production of an effluent 8 containing ammonia and an effluent stream 9 containing hydrogen, methane and nitrogen which is treated to give two effluent streams. First effluent stream 11 is returned as off-gas fuel to the primary reforming stage 20. Second effluent stream 10, a hydrogen-rich stream (> 90 vol% H_2) being returned to the methanol synthesis stage 22 by combining with the methanol synthesis stream 4. Approximately 20 wt% of the total plant capacity goes to the production of ammonia in effluent 8. The plant obviates the use of Air Separation Units (ASU) as well as water gas shift and CO_2-removal stages.

The following table shows the temperatures, pressures and flow rates of the different streams for a process according to Fig. 1 where we are able to produce approximately 3000 MTPD methanol and 750 MTPD ammonia despite the use of a difficult feedstock. The feedstock used is heavy natural gas (85 vol% methane):

TABLE

Position	Temp.	Pressure	Flow rate / kmol/h						
	°C	Bar g	H_2O	H_2	N_2	CH_4	CO	CO_2	Ar
4,4B	947	30.1	5890	12023	1414	419	3147	1043	16
6,6B	35	120.3	2.7	4574	1457	463	17	38	20
7	35	119.3		8742	2914	1036			40
10,10B	35	32		1463	66	61			4
11,11B	35	12		167	477	450			16

The process air compressor 25 can be one common or one for each reforming section.

Stream 4/4B can be distributed to the methanol processes 22/22B depending on individual catalyst activity in order to optimize OPEX via the line 12.

Hydrogen rich stream 10/10B and off-gas 11/11B can be distributed to optimize the process requirements in reforming sections and methanol synthesis sections.

By the nature of the co-production of methanol and ammonia from natural gas, there will not be significant amount of CO_2 available in the process gas to be recovered for urea production. If it is desired to use part or all the ammonia production in the presented process, then CO_2 from the flue gas from one or more reforming sections can be recovered in a CO2 removal section 27.

There would typically be sufficient of CO2 in the flue gas from one reforming section to convert all the ammonia from the combined methanol section effluent gas into urea. Thereby an economy of scale can be achieved for the urea plant, section 29.

A widely used coating material in urea production is urea-formaldehyde-concentrate (UFC85) containing up to 85% urea + formaldehyde and balance is water, in order to avoid caking of urea product. The synergy for production of the coating material in this invention is clear since a small stream of methanol 16 is used to produce formaldehyde and urea solution is required for the absorption of formaldehyde to produce UFC85 or more diluted UFC products in section 28.

CLAIMS

1. Process for co-producing methanol, ammonia and urea from a hydrocarbon feedstock, said process comprising the steps of

a) from a first methanol process, comprising a first reforming step and first methanol conversion step obtaining a first effluent comprising methanol and a first gas effluent comprising hydrogen, nitrogen and unconverted carbon oxides, and
from a second methanol processes, comprising a second reforming step and a second methanol conversion step obtaining a second effluent comprising methanol and a second gas effluent comprising hydrogen, nitrogen and unconverted carbon oxides

b) producing an ammonia synthesis gas from the first and/or second gas effluent in a common catalytic methanation stage and withdrawing said ammonia synthesis gas preferably having a H_2:N_2 molar ratio of approximately 3:1;

c) catalytically converting the nitrogen and hydrogen of the ammonia synthesis gas in a common ammonia synthesis stage and withdrawing an effluent comprising ammonia and a purge-gas stream comprising hydrogen, nitrogen and/or methane, and

d) reacting at least part of the effluent comprising ammonia with at least part of a flue gas comprising CO2 from at least one of the

first and second reforming steps to produce urea.

2. Process according to claim 1, wherein in each of the first and second methanol processes the following steps are carried out

1) producing a methanol synthesis gas containing hydrogen, carbon oxides and nitrogen by steam reforming the hydrocarbon feedstock in a reforming section comprising a primary and a secondary reforming stage;

2) catalytically converting the carbon oxides and hydrogen of the methanol synthesis gas in a once-through methanol synthesis stage and withdrawing an effluent comprising methanol and a gas effluent comprising nitrogen, hydrogen and unconverted carbon oxides;

3. Process according to claim 2 wherein the secondary reforming stage is an air-blown secondary reforming stage.

4. Process according to any of the preceding claims wherein the hydrocarbon feedstock is natural gas or substitute natural gas (SNG) or a gas comprising >80% methane.

5. Process according to any of the preceding claims wherein the methanol conversion of the first and second once-through methanol synthesis stage of the first and the second methanol processes are controlled at least by controlling the amount of syngas from the first and/or second reforming step fed to the first and/or second methanol production stage.

6. Process according to any of the preceding claims wherein the methanol synthesis is conducted by passing the synthesis gas through one or more boiling water reactor and subsequently through one or more adiabatic fixed bed reactors.

7. Process according to any of the preceding claims wherein the methanol synthesis is conducted by passing the synthesis gas through one or more boiling water reactor and subsequently through one or more gas cooled reactors.

8. Process according to claim 7 wherein the boiling water reactor is in the form of a single reactor of the condensing-methanol type which comprises within a common shell a fixed bed of methanol catalyst particles and cooling means adapted to indirectly cooling the methanol synthesis gas with a cooling agent.

9. Process according to claim 7 or 8 further comprising cooling the synthesis gas withdrawn from each methanol reactor to condense methanol and passing the gas through a separator, withdrawing a bottom fraction from the separator containing the raw methanol, withdrawing an overhead fraction containing synthesis gas which is passed to the subsequent methanol reactor, and forming a single liquid effluent containing methanol by combining the bottom fractions of the separators of each reactor containing the raw methanol.

10. Process according to any preceding claim further comprising withdrawing a hydrogen-rich stream from the purge-gas stream comprising hydrogen, nitrogen and methane and adding this stream to step (a).

11. Process according to any preceding claim further comprising withdrawing an off gas from the purge-gas stream comprising hydrogen, nitrogen and methane and adding this stream to step (a) as fuel.

12. Process according to any preceding claim comprising further parallel methanol processes.

13. Process according to any of the preceding claims wherein at least a part of the methanol from the first and/or second methanol process is converted into formaldehyde.

14. Process according to claim 13 wherein the formaldehyde is absorbed into aqueous urea solution based on at least part of the urea to produce UFC85.

15. Process according to claim 13 or 14 wherein the UFC85 is added to the urea as a coating.

16. Plant comprising a first and a second parallel methanol process lines, at least one common methanation reactor and at least one common ammonia synthesis step.

17. Plant according to claim 16, further comprising an interconnecting synthesis line allowing fluid communication between the first and second methanol process line.

18. Plant according to claim 16 or 17, further comprising a Urea production section

19. Plant according to any of claims 16 - 18, further comprising a formaldehyde production section.

20. Plant according to any of claims 16 or 19, further comprising a UFC85 production section.

21. Plant according to any of claims 16 – 20 arranged to carry out the process according to any of the claims 1 – 15.

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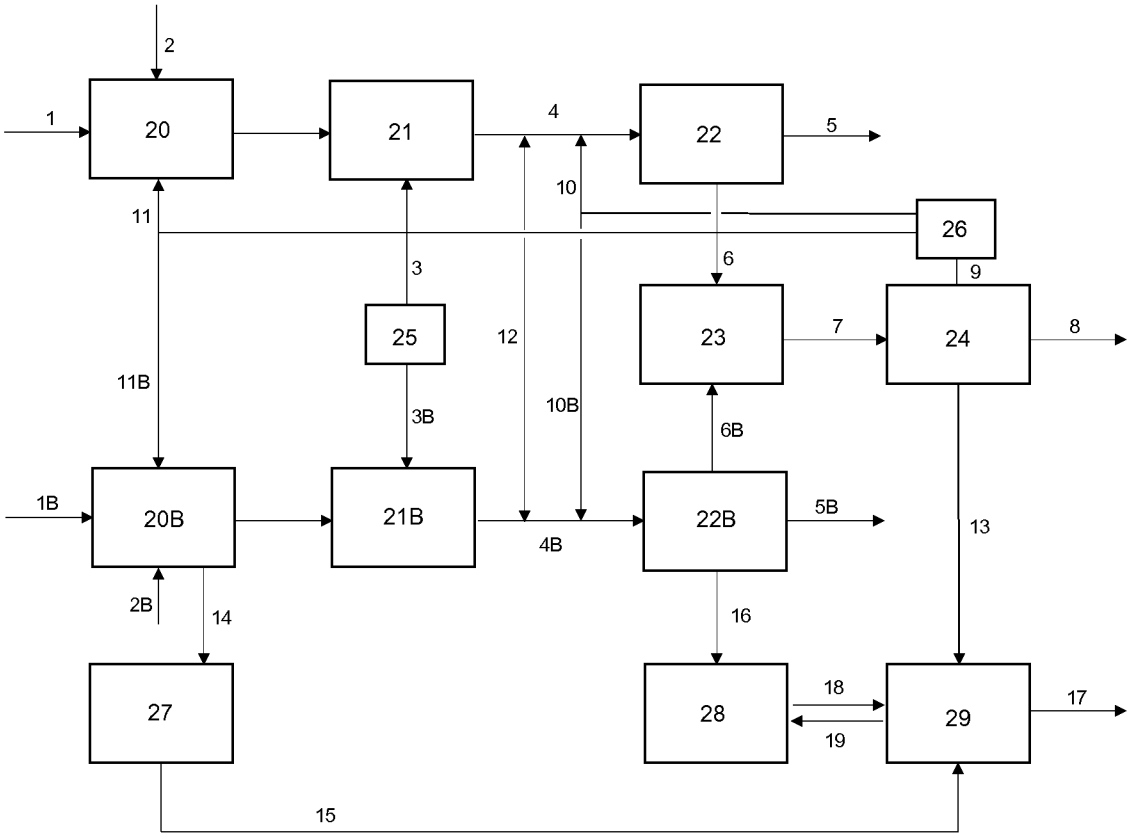

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Fig. 1

INTERNATIONAL SEARCH REPORT

International application No
PCT/EP2018/055648

A. CLASSIFICATION OF SUBJECT MATTER
INV. C07C273/04 C07C273/10 C01C1/04 C07C29/15 C07C29/151
C01B3/02
ADD.
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)
C07C C01B C01C

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)
EPO-Internal, CHEM ABS Data, WPI Data

C. DOCUMENTS CONSIDERED TO BE RELEVANT

Category*	Citation of document, with indication, where appropriate, of the relevant passages	Relevant to claim No.
Y	WO 02/096845 A2 (UHDE GMBH [DE]; BAEHNISCH HANS-JOACHIM [DE]) 5 December 2002 (2002-12-05) paragraphs [0006], [0009], [0010], [0013] - [0017]; figure 2	1-21
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A	paragraphs [0001], [0010], [0016], [0021], [0023], [0025] - [0027], [0030]; figure 1	6-8, 13-15, 18-21
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A	paragraph [0013]; figure 1	6-8, 13-17

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[X] Further documents are listed in the continuation of Box C.

[X] See patent family annex.

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Date of the actual completion of the international search	Date of mailing of the international search report
8 May 2018	16/05/2018

Name and mailing address of the ISA/	Authorized officer
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INTERNATIONAL SEARCH REPORT

International application No
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Y	WO 2017/025272 A1 (HALDOR TOPSØE AS [DK]) 16 February 2017 (2017-02-16)	6-8
A	page 9, line 19 - page 10, line 16	1-5,9-21

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A	page 14, line 4 - page 15, line 14; figure 1	1-12,16, 17

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