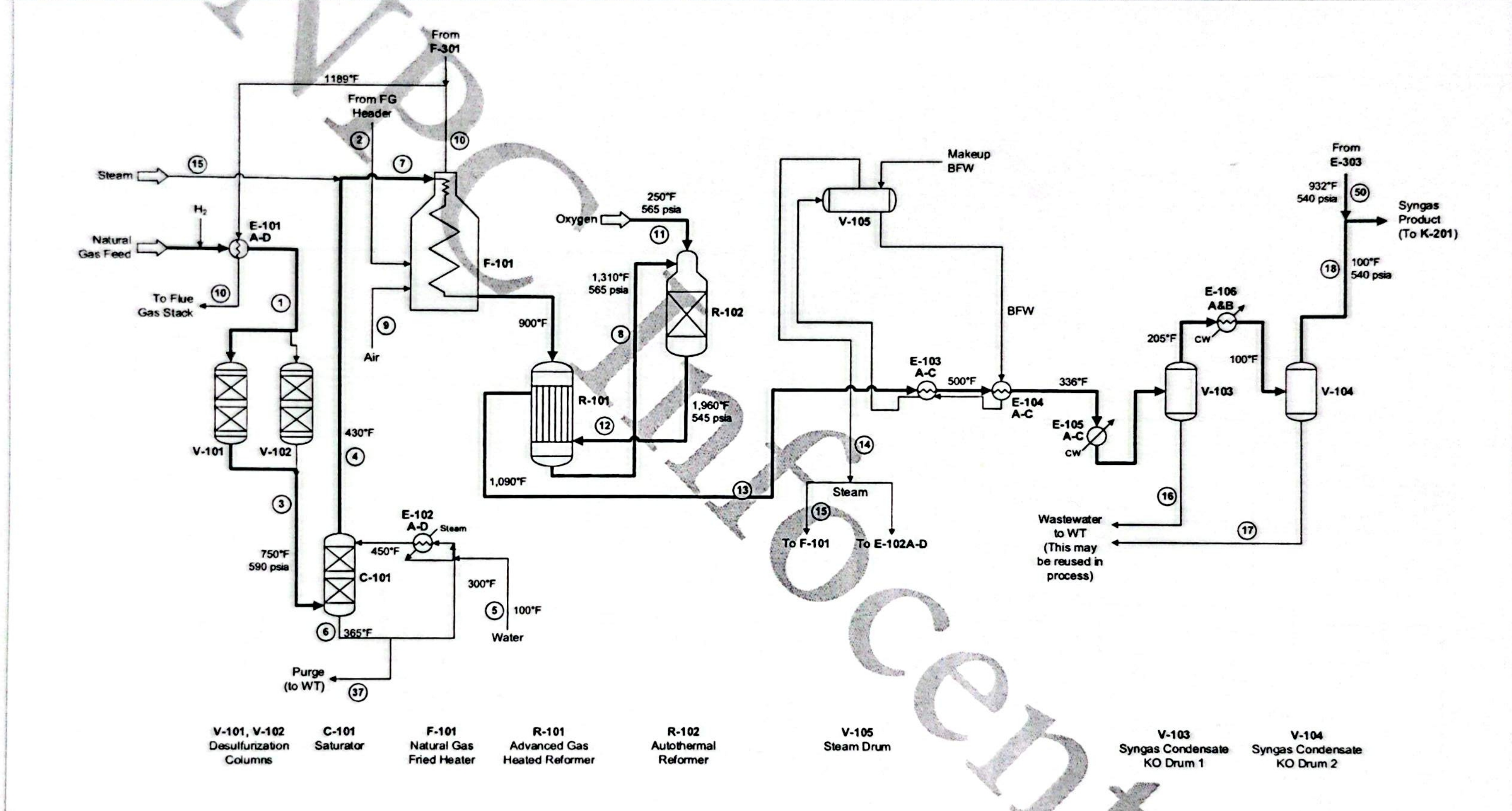
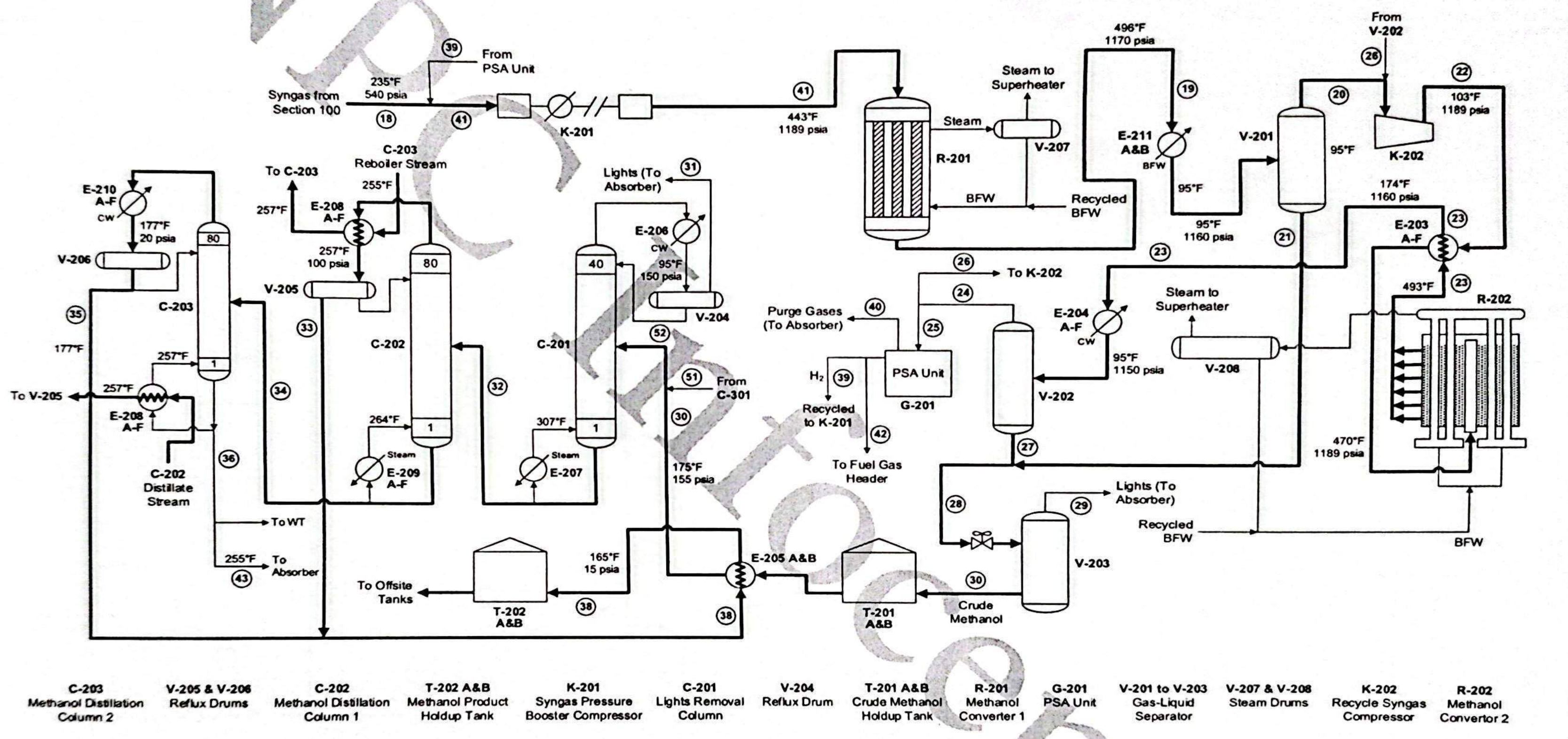
Figure 6.1 (Sheet 1 of 3) Low CO_2 emission methanol production process



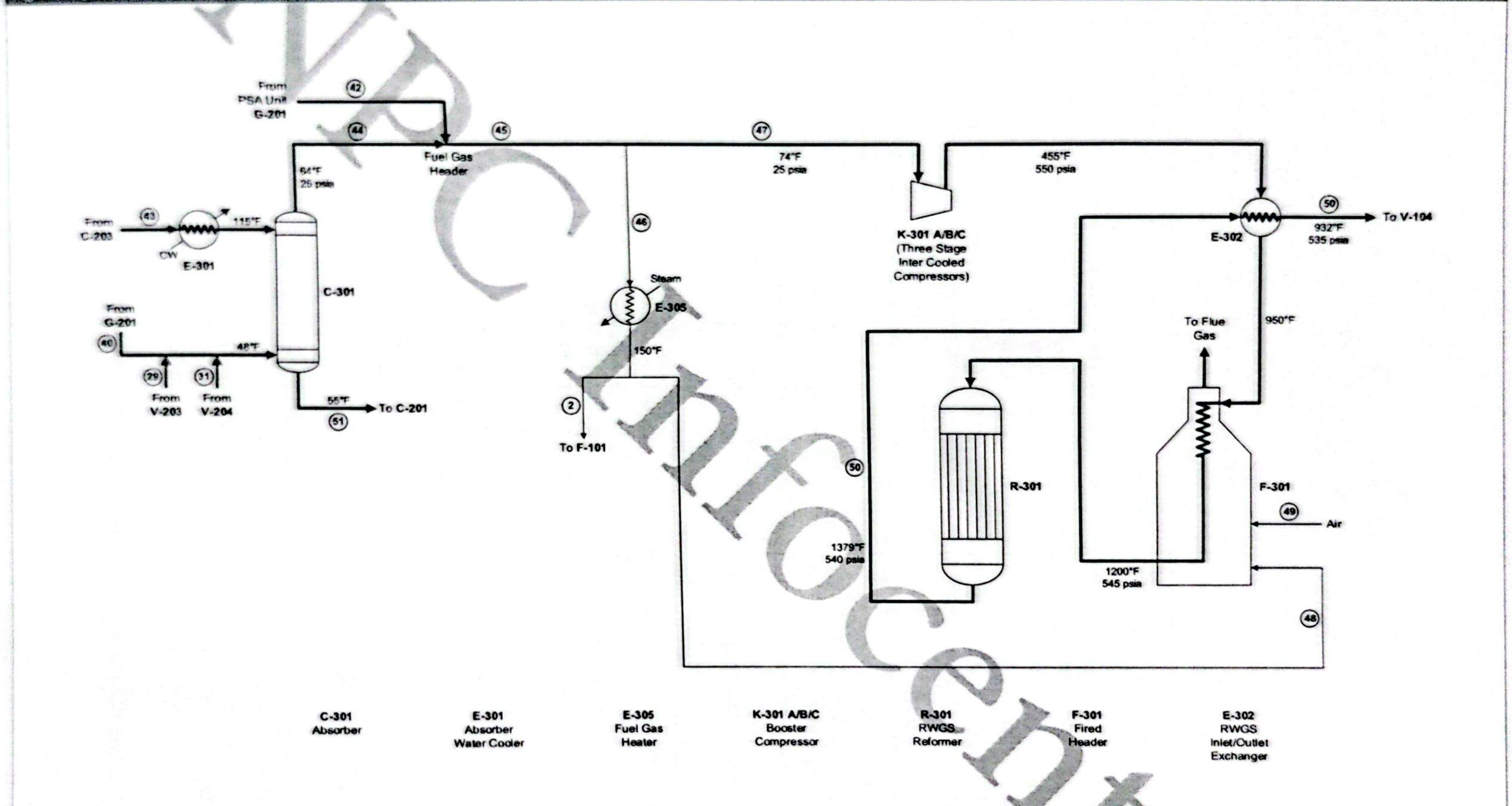
Source: IHS Markit

Figure 6.1 (Sheet 2 of 3) Low CO_2 emission methanol production process



Source: IHS Markit

Figure 6.1 (Sheet 3 of 3) Low CO_2 emission methanol production process



Source: IHS Markit

6 Low carbon dioxide emission methanol production process

In this chapter, we present a technoeconomic evaluation of a large-scale low CO_2 emission methanol production plant using a simulated design of a modified natural gas, two-stage reforming-based process. We have modified the conventional Johnson Matthey/Davy process, which uses the two-stage reforming since we believe it already has the advantage of using effluent from the ATR to heat the GHR. The modification we have introduced is to include an RWGS.

Process description

Table 6.1 lists the design bases and major assumptions for the process. Figure 6.1 in Appendix C presents the main PFD. Table 6.2 shows the flows of the major streams corresponding to Figure 6.1. Tables 6.3 and 6.4 list the major equipment items and utilities consumption summary, respectively. The base capacity of the plant is equivalent to 5,000 mt/d of methanol with 99.85 wt% purity. The plant is assumed to be on the US Gulf Coast.

It is important to note that configuration and design of the process presented by the IHS Markit PEP team are based on our conception and interpretation of the Johnson Matthey/Davy combined-reforming-based methanol production process and the incorporation of the RWGS section to that process. There may be parametric and configurational differences between an actual Johnson Matthey/Davy commercial plant and our simulated design presented for the same.

Section 100—Syngas production

Natural gas, assumed to be available on plant site at a pressure higher than required for the reforming, for example 585 psia at the reformer inlet, is blended with a small amount of H_2. The gas mixture is preheated to 680°F or 360°C in the natural gas/flue gas cross-exchanger (E-101A-D) by the hot flue gas coming from the fired heaters (F-101/F-301). The preheated natural gas, containing 10 wppm of sulfur (assumed value) in the form of sulfur compounds is hydro-desulfurized in a desulfurization column (V-101 or V-102). The two vessels work in sequential order, namely, when V-101 is lined up in operation, the V-102 is on standby, and vice versa. Gas desulfurization takes place in two steps. First, the organic sulfur compounds, usually mercaptans, in the gas react with H_2, getting converted to hydrogen sulfide. A supported cobalt-molybdenum-based catalyst is used for this hydrogenation reaction. The H_2S-containing gas stream then passes over a zinc oxide bed that is in the same vessel, which adsorbs the hydrogen sulfide present in the stream. The V-101 and V-102 are periodically regenerated by passing high-temperature steam or hot air through them.

The desulfurized natural gas is then fed to the saturator (C-101) where it counter currently contacts circulating hot water heated through the saturator feedwater preheater (E-102A-D), connected in parallel. This heating of circulating water in E-102A-D is carried out by the condensing steam. This steam comes in the form of a substream split from the main steam stream generated in the medium-pressure steam waste heat reboiler (E-103A&B). In the heating process, the steam condensate from E-102A-D is subcooled at 300°F or 149°C. The hot condensate is subsequently blended with makeup process water (Stream 5). The resultant condensate mixture at 269°F or 132°C is then passed through E-102A-D, where the temperature of the C-101 circulating water is increased to 450°F or 232°C by heat exchange with the steam described above. A portion of the recirculating water is purged from C-101 on a continuous basis as Stream 37.

The natural gas from the water-contacting step is nearly saturated with the required reformer steam. Hence, most of the primary reformer process steam is provided in this manner, with some supplemental steam additionally provided (Stream 15) to achieve the desired steam-to-carbon ratio of 1.46:1.

The gas stream leaves the saturator overhead at around 430°F or 221°C and then, after blending with the steam, enters the natural gas-fired heater (F-101) as Stream 7 before feeding into primary reforming stage, the advanced gas-heated reformer (R-101). F-101 is a furnace having internal heating coils through which natural gas flows in the downward direction while the flue or combustion gases travel upward on the outside of tubes. The combustion gas is generated from burning the off-gases that are released from different sections of the plant after the methanol synthesis reactor, which are typically combined into a fuel gas pool. The temperature of the primary reformer feed in this heating process is increased at 900°F or 482°C. The hot flue gases leaving the furnace at 1,100°F, which is approximately 600°C, combine with the hot flue gas leaving the furnace of RWGS pass through E-101 to heat the natural gas feed stream to process at 680°F before leaving the process through the stack system. This consists of the entire process CO_2 released for the process.

The basic design of the R-101 is fundamentally identical to that of a conventional shell-and-tube heat exchanger. Natural gas flows through the tubes while the endotherm for reaction is provided by the hot crude syngas stream coming from the secondary ATR unit (R-102) on the shell side of R-101. The tubes are packed with a Ni-based catalyst. Partial reforming of methane, and complete reforming of C_2-plus hydrocarbons takes place in R-101. The reactor effluents come out of the reactor at 1,300°F or 708°C and 565 psia.

The GHR product stream (Stream 8) is fed into the burner of R-102, and combined with oxygen via a specially designed mixer in the neck of the reformer. The specified purity of the oxygen is 99.5%, and the oxygen is assumed to be produced locally in an integrated air separation plant at the required pressure of about 565 psia. The R-102 is a refractory-lined fixed-bed adiabatic reactor packed with a Ni-based catalyst. As a result of reaction exotherm, the temperature of the syngas product increases to 1,964°F or 1,074°C. The hot syngas, which must undergo cooling before its compression for methanol synthesis, represents substantial potential for waste heat energy recovery.

The first cooling step of the R-102 hot syngas product takes place in R-101, where the hot gases provide endotherm for the primary reforming while flowing over the outside surface of the tubes. The gas is cooled at 1,091°F or 588°C in R-101.

In the next step, medium-pressure steam is generated in the medium-pressure steam waste heat reboilers (E-103A-C). The product gas temperature drops at 500°F or 260°C across E-103A-C. This steam at 540°F or 282°C is divided in two parts. One part goes into E-102A-D to preheat the C-101 circulating water at 450°F. The second part (Stream 15) is used as supplemental steam to makeup required steam-to-carbon ratio in the primary reformer feed gas. By-product steam generation from the process section is somewhat positive, but for economic analysis will be considered zero.

Subsequently, the BFW heater (E-104A-C) preheat the high-pressure BFW, or demineralized water, while further cooling the product syngas to 336°F or 169°C. From this point onward, syngas is cooled and partially condensed through a series of two heat exchangers and two following condensate knockout drums: the E-105A-C syngas partial condenser is followed by the V-103 syngas condensate knockout drum 1, and then the E-106A&B syngas partial condenser followed by the V-104 syngas condensate knockout drum 2. Gas partial condensation and phase separation in the above two steps takes place at 205°F (96°C) and 100°F (38°C), respectively, by cooling the gas against cooling water.

After E-106A&B, the nearly dehydrated syngas (Stream 18) is separated from the top of V-104. All multiunit heat exchangers are connected in parallel.

Table 6.1 Low CO_2 emission methanol production process—Design bases and assumptions

Capacity: 5,000 mtpd of methanol

Plant annual stream factor = 0.9

Plant capacity	5,000 MTPD of methanol of 99.85 wt% purity
Basic raw materials	Natural gas and high-purity oxygen
Syngas generation	
Advanced gas heated reformer	
Catalyst	Ni-based
Inlet gas temperature, °F (°C)	900 (482)
Exit gas temperature, °F (°C)	1,310 (710)
Exit gas pressure, psia	565
Methane conversion, mol%	18
C_2^+ hydrocarbons conversion, mol%	100
Steam/Carbon ratio in feed stream, molar	1.46
Autothermal reforming	
Reformer type	Autothermal reformer (boiling water cooled)
External recycle	No
Reforming catalyst	Proprietary Ni-based supported on a Mg-Al carrier
Steam/carbon ratio, molar	1.0
Inlet gas temperature, °F (°C)	1,310 (710)
Exit gas temperature, °F (°C)	1,964 (1,074)
Methane slippage (dry basis), vol%	1.5
Number of ATRs used (in process)	1
Syngas composition, molar	Stoichiometric ratio: 2.048
Natural gas feed composition, % by volume	
CH_4	94.45
C_2H_6	2.68
C_3H_8	1.48
C_4H_{19}	NEG
C_5H_{12}	NEG
CO_2	0.60
N_2	0.78
Total	**100.00**
Oxygen supply to ATR	
Oxygen purity, mol%	99.5+
O_2/C feed ratio, molar	0.43
Number of air separation units used	1
RWGS (Reverse Water Gas Shift) Reactor	
Advanced Gas Heated Reformer	
Catalyst	Ni-based
Inlet gas temperature, °F (°C)	950 (510)
Exit gas temperature, °F (°C)	1,200 (649)
Exit gas pressure, psia	565
CO_2 conversion, mol%	52.4
C_2^+ hydrocarbons conversion, mol%	100

Table 6.1 Low CO_2 emission methanol production process—Design bases and assumptions (continued)

Capacity: 5,000 mtpd of methanol

Plant annual stream factor = 0.9

RWGS as feed composition, % by volume	
CH_4	38.7
N_2	7.9
H_2	35.8
H_2O	0.6
CO	5.9
CO_2	11.1
Total	**100.00**
RWGS outlet composition, % by volume	
CH_4	44.1
N_2	8.4
H_2	22.6
H_2O	9.9
CO	9.4
CO_2	5.6
Total	**100.00**
Methanol production	
Methanol converters configuration	Two converters connected in series
First converter	
Type	Axial-flow, shell-and-tube type reactor. Cooling is done by boiling water circulating on shell side
Makeup syngas stoichiometric ratio, molar	2.048
Converter syngas feed stoichiometric ratio, molar	2.547
Converter recycle loop	Yes (recycled hydrogen/recycle RWGS syngas)
Ratio of reactor feed syngas to makeup syngas, molar	~1.2
Feed gas pressure, psia	1,189
Exiting gas pressure, psia	1,160
Converter feed stream temperature, °F (°C)	488 (253)
Converter exit stream temperature, °F (°C)	496 (258)
Conversion per pass (based on CO+CO_2), mol%	62.7
Catalyst	A proprietary Cu-ZnO-Al_2O_3-based catalyst, possibly doped with additives/promoters
Second converter	
Type	Radial-flow, shell-and-tube type reactor. Cooling is done by boiling water circulating inside tubes
Converter syngas feed stoichiometric ratio, molar	8.38
Converter recycle loop	Yes, unconverted syngas from converter 1
Ratio of reactor feed syngas to makeup syngas, molar	~4.32
Feed gas pressure, psia	1,189
Exiting gas pressure, psia	1,160
Converter feed stream temperature, °F (°C)	470 (243)
Converter exit stream temperature, °F (°C)	493 (256)
Conversion per pass (based on CO+CO_2), mol%	81.24
Catalyst	A proprietary Cu-ZnO-Al_2O_3-based catalyst, possibly doped with additives/promoters
Product separation	By a series of three distillation columns

Source: IHS Markit

Section 200—Methanol production

The cooled and dewatered syngas stream from Section 100 (Stream 18) is blended with the recycled hydrogen stream (Stream 39) coming from the pressure adsorption unit (G-201) and the recycled gas from the R-301 RWGS. The blended stream undergoes compression to 1,189 psia in the syngas pressure booster compressor (K-201), a two-stage, intercooled compressor. Since the gas coming from the RWGS is at a temperature of 915°F or 490°C, the temperature of the blended gas is at 487°F or 253°C. The blended stream enters the methanol converter 1 (R-201) as Stream 41 at 487°F or 253°C and 1,189 psia. R-201 is like a shell-and-tube heat exchanger. The catalyst is loaded inside the tubes, while on the shell side, circulating boiling water removes the heat of reaction and, consequently, turns into saturated steam at 450°F or 232°C and 422 psia. This steam is separated in the steam drum (V-207) and subsequently sent to superheater.

The R-201 product or effluent stream (Stream 19), leaving the converter at 496°F or 258°C and 1,160 psia, the cooled converter product stream is then partially condensed in the converter-1 product partial condenser (E-202A-F) by heat-exchanging with BFW-producing steam. The product stream is further cooled and partially condensed in the converter-1 product partial condenser (E-211A&B) with cooling water. The partially condensed product stream, which mainly consists of methanol, water, unconverted syngas, and a small number of by-products, are phase-separated in the gas-liquid separator (V-201).

The overhead vaporous stream from V-201 (Stream 20) is blended with the recycle syngas stream (Stream 26) coming from G-201; the blended stream is recompressed to 1,189 psia in the recycle syngas compressor (K-202). The compressed blended stream (Stream 22) then passes through E-203A-F—the R-202 feed/product heat exchanger—in which it exchanges heat with the hot product stream (Stream 23) coming from the methanol converter 2 (R-202) at 493°F or 256°C. Stream 22 enters R-202 at 470°F or 243°C and 1,182 psia. R-202's mechanical structure is similar as that of a shell-and-tube heat exchanger. Nevertheless, unlike R-201, in R-202 the coolant, or boiling water, flows through the tubes while the catalyst is packed outside the tubes. Syngas enters the converter through a vertical feed tube, which has side openings along its length. This tube is closed from the top. The converter is basically a radial-flow, steam-raising-type of reactor. The syngas stream enters through the feed tube, passes through openings, and then travels in the radial direction through the catalyst bed. Steam produced in the converter from reaction exotherm is separated in the steam drum (V-208), and subsequently sent to steam superheater (not shown in PFD).

The R-202 product stream (Stream 23) is cooled in E-203A-F by exchanging heat with the reactor feed stream (Stream 22), and in the process its temperature drops down at 167°F or 75°C. Stream 23 is subsequently partially condensed in E-204A-F, the converter 2 product partial condenser, using cooling water as coolant. The partially condensed stream is then phase-separated in the gas-liquid separator (V-202). The condensate streams from V-201 and V-202 (Streams 21 and 27) are combined, and the mixed condensate stream (Stream 28) is flashed and phase-separated in another gas-liquid separator (V-203), expelling out the dissolved gases (Stream 29) from the crude methanol product. The crude methanol product is stored in the crude methanol holdup tank (T-201A&B).

The overhead vaporous stream of V-202 (Stream 24) is split in two parts. One part is recycled to K-202 in the form of Stream 26. The second part, Stream 25, is sent to G-201 to recover H_2 (Stream 39) from the syngas. The H_2-lean gas stream (Stream 40) from G-201 is used as a supplemental fuel in the process.

Section 300—RWGS section

This section has been included in our process to reduce the CO_2 emission compared with the standard Johnson Matthey/Davy process. The gases released from the V-203 gas-liquid separator, because of the pressure letdown, from the C-201 methanol distillation column and from PSA unit G-201—the residue after separating H_2 as described earlier—which are normally used as fuel gas/purge gases are collected and blended and sent to the C-301 water absorption tower. The gas flows counter current to the water, which is fed from the top of the column. The water absorbs the trace methanol contained in these gases. The water leaving C-301 is blended with the feed to column C-201. The column operates at a pressure of 25 psia. Water removed in the C-203 distillation column for methanol purification is cooled to 115°F or 46°C, and recycled to C-301. Gas leaves C-301 at 60°F or 15°C and 25 psia. This gas is then blended with slipstream of H_2-rich gas from G-201. This then forms the fuel gas pool for the process plant. Part of the gas from this fuel gas pool is split to the furnace F-101 for the advanced GHR unit R-101 and furnace F-301 for the RWGS reactor R-301.

The gas that remains from the fuel gas pool is then compressed in RWGS feed compressor K-301 to a pressure of 538 psia. K-301 is a steam-driven, three-stage intercooled compressor system. The compressed gas is preheated through exchanger E-301 to exchange heat with hot gases leaving R-301 RWGS reactor. The gas leaves E-301 at temperature of 950°F or 510°C and fed to furnace F-301. The gas is heated to 1,200°F or 649°C before feeding the R-301 RWGS. The gases exit R-301 at 1,365°F or 740°C. This stream is cooled against the incoming gas in exchanger E-301. This cooled gas is blended with the gas feed to methanol converter 1 R-201. We had described R-201 earlier in Methanol Production—Section 200. The advantage that accrues by using the RWGS is that no gas is flared or vented, thereby increasing the yield of methanol and reducing the total CO_2 emission from the process.

Before going to the distillation section, the crude methanol stream (Stream 30) interchanges heat with the purified methanol product stream (Stream 38) in the methanol product/crude methanol heat exchangers E-205A&B.

Crude methanol is purified in a battery of three distillation columns. The lights removal column (C-201) primarily distills out the dissolved gases present in the crude methanol stream. Light reaction side products, e.g., DME, is also distilled off with the light gases; a tiny amount of methanol is also let above in the distillate to restrict the distillate temperature at 95°F or 35°C, thus eliminating the need for a refrigerant in the column condenser (E-206). The column reboiler is operated at 305°F or 152°C and 165 psia. The methanol-water mixture obtained as C-201 bottoms is then pumped to the methanol distillation column 1 (C-202). C-202 is operated at a condenser pressure of 105 psia—this tower can optionally be operated at a somewhat lower pressure as well. The vaporous distillate stream leaving the top of the column at 260°F or 127°C goes directly into the reboiler of C-203, providing the thermal duty of C-203 reboiler. The condensed methanol-water stream from E-208A-F is collected in the C-202 reflux drum (V-205). A part of the condensed stream is refluxed in C-202. The remaining portion of the condensed stream (Stream 33) is then passed through E-205 after blending the stream with Stream 38 to cool it before the methanol product is collected in the methanol product holdup tank (T-202A&B) for subsequent transfer to the off-site methanol storage tanks (not shown in the PFD).

The bottom-end stream of C-202 is subsequently processed in the methanol distillation column 2 (C-203). This column is operated at atmospheric pressure. High-purity methanol is separated from water in the column overhead stream (Stream 33). This product stream is then cooled after blending it with the overhead methanol product stream of C-202 (Stream 33) in E-205 and sent into T-202A&B for its subsequent transfer to off-site storage tanks. The C-203 bottom-end stream (Stream 36) is comprised of mainly water.

All multiunit heat exchangers are connected in parallel.

Process discussion

The following factors were considered either as criteria or assumptions in our design. Any sizable variation may alter the methanol plant's sectional economics to some extent. However, the variation in the plant's overall economics may not be significant relatively.

Feedstock

We have assumed a methane concentration in the natural gas feedstock of approximately 94 mol%. Other hydrocarbons in the gas include ethane and propane. Higher hydrocarbons (C_4–C_5) may also be present. Those hydrocarbons comprise straight-chained as well as branched hydrocarbons.

The pressure of the feedstock natural gas from the supply source is assumed to be high enough for feeding the gas in the reformer without having to pass through a compressor. In case of lower supply pressure, a compressor would be required. Overall impact on the plant economics caused by that will be minimal. Oxygen, produced onsite from an ASU, is also assumed to be coming at a pressure sufficient for direct feed to reforming.

In the feed gas pretreatment stage, a hydrogenation step is provided to convert the organic sulfur compounds present in feed into hydrogen sulfide. Hydrogenation is carried out using a cobalt-molybdenum catalyst. Zinc oxide absorbent beds follow to remove the hydrogen sulfide formed in the first bed. Two vessels are provided to allow regeneration/replacement of the spent catalyst in a vessel while the plant remains onstream on the second vessel. The feedstock is also assumed to contain none or less than 1 ppm of chloride compounds, which can poison the reforming catalyst. Chloride compounds in feed result in the formation of HCl during hydrogenation. The HCl, in turn, reacts with ZnO forming $ZnCl_2$. At normal operating conditions of ZnO bed, $ZnCl_2$ sublimates and deposits on the downstream catalysts and heat transfer surfaces. Any HCl, if formed, would require an activated alumina absorbent bed (alumina guardbed) placed in between the hydrogenation and ZnO beds.

Unreacted syngas recycling

Our design includes internal recycles of hydrogen (Stream 39) and unreacted syngas (Stream 26).

The internal syngas recycle is comprised of a partial recycling of unreacted syngas obtained after its separation from the crude methanol stream. Besides unconverted syngas, a hydrogen stream is also recycled (Stream 39) primarily to enhance the stoichiometric ratio of syngas feed entering the first methanol converters.

Methanol converters sizing estimate

Converter sizing is primarily based on the volume of catalyst needed for the reaction and the heat transfer flux that is likely to be attained with the available temperature differential of the tube and shell-side fluid. Parameters such as number and length of the tubes may be different from those in an industrial reactor.

Converters feed/product streams configuration and material balance

The flow configuration of the converter's feed/product streams and their material balance around the converters is derived from US patent 2019/0016655, which is assigned to Johnson Matthey/Davy. We also examined another Johnson Matthey/Davy patent, WO 2017/121980, which differed somewhat in configuration from the first patent. Calculation of per-pass conversions of the carbon oxides in methanol converters was estimated from those patents. It is uncertain whether those stream compositions and steam flow rates exist in Johnson Matthey/Davy's commercial plants.

Methanol product purification

We have used boiling water-cooled methanol converters in our design. This type of reactor provides efficient and easy control of the reaction temperature and simultaneously ensures that the released reaction heat can be used efficiently. By controlling the steam pressure, and thus fixing the steam-side temperature over the whole length of the converter, the reaction temperature is controlled throughout the converter. The isothermal nature of the BWR gives a high conversion per unit mass of the catalyst loaded in the reactor.

Two converter vessels (R-201 and R-202) have been used in our design. These converters are connected in series, but operate with common steam pressure. We have used a three-column distillation train to purify crude methanol produced in the converters. The first column removes lights as an overhead stream from the condensing circuit that is operated on total reflux. The second and third columns are methanol purification columns. The second column is operated at medium pressure, while the third column operates at atmospheric pressure. This configuration allows the overhead vapor stream from the second column to condense by providing heat in the third column's reboiler.

Steam consumption

According to our design, 396,900 lb/hr of medium-pressure steam is generated at 540°F or 282°C in the syngas production section (E-103A&B). This steam is divided in two parts. One part goes into E-102A-D to preheat the C-101 circulating water at 450°F or 232°C as described in the process description section. The second part (Stream 15) is used as supplemental steam to achieve the required steam-to-carbon ratio in the primary reformer feed gas. By-product steam generation from Section 100 is zero.

Apart from the above steam, 778,100 lb/hr steam is generated at 450°F and 422 psia in the methanol converters. This steam is superheated in the convection section's furnaces, F-101 and F-301. This superheated medium-pressure steam is used to drive the turbines of the air compressor, syngas pressure booster compressor, BFW pump, and RWGS compressors. In total, these require 88,950 bhp. In addition, cooling of the effluent from the first methanol reactor (R-101) generates 10,447,000 lb/hr steam at 370°F or 188°C at 422 psia. This large amount of steam is generated because of the recycle gas from the RWGS, which when mixed with other feeds heats up the total feed to the R-101. Hence the entire effluent gas from R-101 can be used for heat generation and need to be used to exchange heat with incoming reactant. Part of this steam is used in the reboilers of C-201 and C-202. The excess steam can be used for more power generation or exported.

It should be noted that a detailed heat and steam balance is beyond the scope of this report. We present this section to create the awareness that sufficient steam is generated to make the plant self-sufficient in power generation and steam requirements.

Plant start-up boiler

It may be possible that steam required in the process for reforming, compressors operation, and equipment preheating, such as initial heating of desulfurization columns and reformers, at the time of the plant's cold or warm start-up is available at the site from an adjacent plant. If such steam is not available, which is a more likely situation, then a start-up steam boiler would be needed to supply steam for start-up. A 350,000 lb/hr steam plant is included in the equipment inventory and its financial impact has been taken into account. This boiler will periodically provide backup safety for process operation during possible process upsets by providing supplemental steam. A 350,000 lb/hr capacity of steam generation is considered enough for start-up as per industrial practice.

Materials of construction

Materials of construction for various equipment used in the process are in the major equipment list (Figure 6.3).

Miscellaneous plant sections

Our designed process consists of several other sections and equipment that are not directly related or involved in the production of methanol. These sections are not shown in the PFD in Figure 8.1. Included in the list of major sections are wastewater treatment and recovery system; air separation plant; raw water treatment section; steam, condensate, and electric grid systems; plant flare system, water cooling system; and computerized plant control system. Details of these systems are not in the PFD or in the process description section primarily because of scope limitations for this report. However, the costs of such systems have been taken into consideration in our plant cost estimates.

Table 6.2 Low CO_2 emission methanol production process—Mainstream flows

Capacity: ~5,000 metric tons per day (MTPD) of methanol of 99.85 wt% purity

Plant Annual Stream Factor: 0.9

		Stream flows								
Component	**Mol.wt**	**(1)**	**(2)**	**(3)**	**(4)**	**(5)**	**(6)**	**(7)**	**(8)**	**(9)**
CH_4	16.04	245,277	12,176	245,277	245,212	-	213	245,212	201,681	-
C_2H_6	30.07	13,067	-	13,067	13,064	-	3	13,064	52	-
C_3H_8	44.10	10,606	-	10,606	10,606	-	-	10,606	-	-
C_4H_{10}	58.12	--	-	-	-	-	-	-	-	-
C_5H_{12}	72.15	--	-	-	-	-	-	-	-	-
H_2S	34.02	--	-	-	-	-	-	-	-	-
N_2	18.02	3,542	5,203	3,542	3,540	-	13	3,540	3,540	227,791
H_2	2.02	942	1,725	942	941	-	7	941	30,447	-
H_2O	18.02	--	289	-	389,983	58,331	2,607,714	446,181	319,353	-
Methanol	32.04	--	-	-	-	-	-	-	-	-
CO	28.01	--	2,748	-	-	-	-	-	44,142	-
CO_2	44.01	4,317	10,671	4,317	4,315	-	151	4,315	124,644	-
O_2	32.00	--	-	-	-	-	-	-	-	69,199
Argon	32.04	--	-	-	-	-	-	-	-	-
Sulphur	--	14	-	-	-	-		-	-	-
Alcohols	--	-	-	-	-	-	-	-	-	-
Dimethyl Ether	46.06	-	-	-	-	-	-	-	-	-
Flow lb/hr		**277,765**	**32,812**	**277,751**	**667,661**	**58,331**	**2,608,101**	**723,859**	**723,859**	**296,990**
Flow kg/hr		**125,970**	**14,881**	**125,964**	**302,794**	**26,454**	**1,182,812**	**328,281**	**328,281**	**134,689**

Source: IHS Markit

Table 6.2 Low CO_2 emission methanol production process—Mainstream flows (continued)

Capacity: ~5,000 metric tons per day (MTPD) of methanol of 99.85 wt% purity

Plant Annual Stream Factor: 0.9

		Stream flows								
Component	**Mol. wt.**	**(10)**	**(11)**	**(12)**	**(13)**	**(14)**	**(15)**	**(16)**	**(17)**	**(18)**
CH_4	16.04	-	-	5,932	5932	-	-	-	0	57,046
C_2H_6	30.07	-	-	-	-	-	-	-	0	1
C_3H_8	44.10	-	-	-	-	-	-	-	0	-
C_4H_{10}	58.12	-	-	-	-	-	-	-	0	-
C_5H_{12}	72.15	-	-	-	-	-	-	-	0	-
H_2S	34.02	-	-	-	-	-	-	-	0	-
N_2	18.02	254,954	1,098	4,638	4638	-	-	-	0	24,562
H_2	2.02	-	-	74,186	74,186	-	-	6	0	97,054
H_2O	18.02	47,281	-	366,227	366,227	419,317	56,198	338,989	25,097	17,741
Methanol	32.04	-	-	-	-	-	-	-	0	-
CO	28.01	-	-	365,427	365,427	-	-	26	1	383,944
CO_2	44.01	530,42	-	158,232	158,232	-	-	331	50	174,696
O_2	32.00	5,609	249,685	-	-	-	-	-	0	-
Argon	32.04	-	-	-	-	-	-	-	0	-
Sulphur	–	-	-	-	-	-	-	-	0	-
Alcohols	–	-	-	-	-	-	-	-		-
Dimethyl Ether	46.06	-	-	-	-	-	-	-		-
Flow lb/hr		**360,886**	**250,783**	**974,642**	**974,642**	**419,317**	**56,198**	**339,352**	**251,48**	**755,044**
Flow kg/hr		**163,667**	**113,734**	**442,015**	**442,015**	**190,166**	**25,487**	**153,901**	**11,405**	**342,423**

Source: IHS Markit

Table 6.2 Low CO_2 emission methanol production process—Mainstream flows (continued)

Capacity: ~5,000 metric tons per day (MTPD) of methanol of 99.85 wt% purity

Plant Annual Stream Factor: 0.9

		Stream flows								
Component	**Mol. wt.**	**(19)**	**(20)**	**(21)**	**(22)**	**(23)**	**(24)**	**(25)**	**(26)**	**(27)**
CH_4	16.04	57,046	56,408	638	788,066	788,066	787,501	59,063	728,438	565
C_2H_6	30.07	1	1	-	10	10	10	1	10	-
C_3H_8	44.10	-	-	-	-	-	-	-	-	-
C_4H_{10}	58.12	-	-	-	-	-	-	-	-	-
C_5H_{12}	72.15	-	-	-	-	-	-	-	-	-
H_2S	34.02	-	-	-	-	-	-	-	-	-
N_2	18.02	24,562	24,432	131	336,131	336,131	335,841	25,188	310,653	290
H_2	2.02	69,288	69,185	103	404,450	363,070	362,989	27,224	335,765	81
H_2O	18.02	5,832	28	5,804	922	62,042	962	72	890	61,080
Methanol	32.04	232,713	7,958	224,755	37,896	314,609	32,418	2,431	29,986	282,191
CO	28.01	161,798	161,428	370	321,824	174,779	174,722	13,104	161,618	57
CO_2	44.01	203,828	190,213	13,614	596,729	447,326	441,549	33,116	408,433	5,776
O_2	32.00	-	-	-	-	-	-	-	-	-
Argon	32.04	-	-	-	-	-	-	-	-	-
Sulphur	–	-	-	-	-	-	-	-	-	-
Alcohols	–	-	-	-	-	-	-	-	-	-
Dimethyl Ether	46.06	-	-	-	-	-	-	-	-	-
Flow lb/hr		**755,068**	**509,654**	**245,414**	**2,486,028**	**2,486,034**	**2,135,993**	**160,200**	**1,975,794**	**350,040**
Flow kg/hr		**342,435**	**231,135**	**111,299**	**1,127,451**	**1,127,453**	**968,704**	**72,653**	**896,052**	**158,748**

Source: IHS Markit

Table 6.2 Low CO_2 emission methanol production process—Mainstream flows (continued)

Capacity: ~5,000 metric tons per day (MTPD) of methanol of 99.85 wt% purity
Plant Annual Stream Factor: 0.9

		Stream flows								
Component	Mol. wt.	(28)	(29)	(30)	(31)	(32)	(33)	(34)	(35)	(36)
CH_4	16.04	1,203	1,131	72	72	-	-	-	-	-
C_2H_6	30.07	-	-	-	-	-	-	-	-	-
C_3H_8	44.10	-	-	-	-	-	-	-	-	-
C_4H_{10}	58.12	-	-	-	-	-	-	-	-	-
C_5H_{12}	72.15	-	-	-	-	-	-	-	-	-
H_2S	34.02	-	-	-	-	-	-	-	-	-
N_2	18.02	420	400	20	20	-	-	-	-	-
H_2	2.02	184	181	2	2	-	-	-	-	-
H_2O	18.02	66,884	18	66,867	-	83,228	509	82,720	1	840,563
Methanol	32.04	506,946	884	506,062	175	509,373	508,360	1,240	1,239	6
CO	28.01	426	418	8	8	-	-	-	-	-
CO_2	44.01	19,391	10,812	8,578	8,595	-	-	-	-	-
O_2	32.00	-	-	-	-	-	-	-	-	-
Argon	32.04	-	-	-	-	-	-	-	-	-
Sulphur	–	-	-	-	-	-	-	-	-	-
Alcohols	–	-	-	-	-	-	-	-	-	-
Dimethyl Ether	46.06	-	-	-	-	-	-	-	-	-
Flow lb/hr		**595,455**	**13,845**	**581,609**	**8,873**	**592,602**	**508,869**	**83,960**	**1,241**	**840,569**
Flow kg/hr		**270,047**	**6,279**	**263,768**	**4,024**	**268,754**	**230,780**	**38,077**	**563**	**381,211**

Source: IHS Markit

Table 6.2 Low CO_2 emission methanol production process—Mainstream flows (continued)
Capacity: ~5,000 metric tons per day (MTPD) of methanol of 99.85 wt% purity
Plant Annual Stream Factor: 0.9

		Stream flows								
Component	**Mol. wt.**	**(37)**	**(38)**	**(39)**	**(40)**	**(41)**	**(42)**	**(43)**	**(44)**	**(45)**
CH_4	16.04	1	-	343	58,572	57,046	147	-	59,775	59,923
C_2H_6	30.07	-	-	-	1	1	-	-	1	1
C_3H_8	44.10	-	-	-	-	-	-	-	-	-
C_4H_{10}	58.12	-	-	-	-	-	-	-	-	-
C_5H_{12}	72.15	-	-	-	-	-	-	-	-	-
H_2S	34.02	-	-	-	-	-	-	-	-	-
N_2	18.02	-	-	-	25,188	24,562	-	-	25,608	25,608
H_2	2.02	-	-	18,918	199	97,054	8,108	-	382	8,490
H_2O	18.02	9,015	510	-	72	17,741	-	17,697	1,423	1,423
Methanol	32.04	-	509,599	-	2,431	-	-	-	-	-
CO	28.01	-	-	-	13,104	383,944	-	-	13,530	13,530
CO_2	44.01	1	-	-	33,116	174,696	-	8	52,516	52,516
O_2	32.00	-	-	-	-	-	-	-	-	-
Argon	32.04	-	-	-	-	-	-	-	-	-
Sulphur	–	-	-	-	-	-	-	-	-	-
Alcohols	–	-	-	-	-	-	-	-	-	-
Dimethyl Ether	46.06	-	-	-	-	-	-	-	-	-
Flow lb/hr		**9,016**	**510,109**	**19,261**	**132,684**	**755,044**	**8,255**	**17,705**	**153,236**	**161,491**
Flow kg/hr		**4,089**	**231,342**	**8,735**	**60,174**	**342,423**	**3,744**	**8,030**	**69,495**	**73,238**

Source: IHS Markit

Table 6.2 Low CO_2 emission methanol production process—Mainstream flows (continued)

Capacity: ~5,000 metric tons per day (MTPD) of methanol of 99.85 wt% purity

Plant Annual Stream Factor: 0.9

		Stream flows						
Component	**Mol. wt.**	**(46)**	**(47)**	**(48)**	**(49)**	**(50)**	**(51)**	**(52)**
CH_4	16.04	13,321	46,602	1,147	-	50,771	-	72
C_2H_6	30.07	-	1	-	-	1	-	-
C_3H_8	44.10	-	-	-	-	-	-	-
C_4H_{10}	58.12	-	-	-	-	-	-	-
C_5H_{12}	72.15	-	-	-	-	-	-	-
H_2S	34.02	-	-	-	-	-	-	-
N_2	18.02	5,693	19,916	490	21,467	19,924	-	20
H_2	2.02	1,887	6,603	163	-	3,957	-	2
H_2O	18.02	316	1,106	27	-	15,600	16,363	83,228
Methanol	32.04	-	-	-	-	-	3,491	509,548
CO	28.01	3,008	10,523	259	-	18,545	-	8
CO_2	44.01	11,674	40,842	1,006	-	16,845	17	8,595
O_2	32.00	-	-	-	6,521	-	-	-
Argon	32.04	-	-	-	-	-	-	-
Sulphur	–	-	-	-	-	-	-	-
Alcohols	–	-	-	-	-	-	-	-
Dimethyl Ether	46.06	-	-	-	-	-	-	-
Flow lb/hr		35,899	125,592	3,092	27,988	125,643	19,871	601,474
Flow kg/hr		16,281	56,958	1,402	27,988	56,981	9,012	272,777

Source: IHS Markit

Table 6.3 Low CO_2 emission methanol production process—Major equipment

Capacity: ~5,000 metric tons per day (Mtpd) of methanol of 99.85 wt% purity

Plant annual stream factor: 0.9

Equipment number	Name	Size		Material of construction	Remarks
Reactors					
R-101	Advanced gas-heated reformer			Shell: C.S. Tubes: Cr-Mo-NB	Tube dia = 3-inch. Tube length = 35 feet. Number of tubes = 1,600. Tubes contain 2,750 ft^3 of Ni-based catalyst
R-102	Autothermal reformer	36,250 12.5 X 39.5 ft (HT)	gal	Shell: C.S.lining: Refractory	Filled with 3,075 ft^3 of a nickel-based catalyst in the SMR section of reformer
R-201	Methanol converter 1	4,250 ft^3 of catalyst		Shell: C.S. Tubes: 304 SS	Boiling-water reactor. Coolant on shell. Side copper-based catalyst inside tubes. The converter contains 4,800 tubes. Tube id = 2.5 inch. Tube length = 26 feet. Shell dia = 24 ft shell length
R-202	Methanol converter 2	5,500 ft^3 of catalyst		Shell: 304 SS Tubes: 304 SS	Boiling-water reactor. Coolant on tube tubes: 304 SS side. Copper-based catalyst on shell side. The converter contains 6,000 tubes. Tube id = 2 inch. Tube length = 30 feet. Shell dia = 23 ft shell length = 38 ft
R-301	Rwgs reactor reformer			Shell: C.S. Tubes: Cr-Mo-NB	Tube dia = 3-inch. Tube length = 35 feet Number of tubes = 1,600. Tubes contain 2,750 ft^3 of Ni-based catalyst
Columns					
C-101	Saturator	13 65	ft dia ft	Shell: C.S. Packing: 304 SS	Column is packed with pall rings. Packed height = 50 feet
C-201	Lights removal column	12 90	ft dia ft	Shell: 304 SS Trays: 304 SS	40 valve trays, 24-inch spacing
C-202	Methanol distillation column 1	16 170	ft dia ft	Shell: 304 SS Trays: 304 SS	80 valve trays, 24-inch spacing
C-203	Methanol distillation column 2	20 170	ft dia ft	Shell: 304 SS Trays: 304 SS	80 valve trays, 24-inch spacing
C-301	Methanol absorber	9 45	ft dia ft	Shell: C.S. Packing: 304 SS	Column is packed with pall rings Packed height = 50 feet
Compressors					
K-101	Combustion air blower	3,625	bhp	C.S.	Not shown in PFD. Electric-driven
K-102	Main air compressor (ASU part)	29,460	bhp	C.S.	Turbo-driven from internally-produced steam
K-201	Syngas pressure booster	39,215	bhp	C.S. Two-stage, water-cooled	Turbo-driven from internally-produced steam
K-202	Recycle syngas compressor	3,250	bhp	C.S.	Turbo-driven from internally-produced steam
K-301	Recycle syngas compressor	3,250	bhp	C.S.	Turbo-driven from internally-produced steam
Heat exchangers					
E-101A-D	NG/flue gas cross-exchanger	13,725 31.0	sq ft each MMBtu/hr each	Shell: C.S. Tubes: C.S.	

Table 6.3 Low CO_2 emission methanol production process—Major equipment (continued)

Capacity: ~5,000 metric tons per day (Mtpd) of methanol of 99.85 wt% purity

Plant annual stream factor: 0.9

Equipment number	Name	Size		Material of construction	Remarks
E-102A-D	Saturator feedwater preheater	24,400	sq ft each	Shell: C.S	
		78.6	MMBtu/hr each	Tubes: C.S	
E-103A-C	MP steam waste heat reboiler	13,900	sq ft each	Shell: C.S	
		96.2	MMBtu/hr each	Tubes: C.S	
E-104A-C	BFW heater	24,450	sq ft each	Shell: C.S	
		53.0	MMBtu/hr each	Tubes: 316 SS	
E-105A-C	Syngas partial condenser	19,500	sq ft each	Shell: C.S	
		98.8	MMBtu/hr each	Tubes: 316 SS	
E-106A&B	Syngas partial condenser	23,200	sq ft each	Shell: C.S	
		31.3	MMBtu/hr each	Tubes: 316 SS	
E-201A-F	Converter-1 product partial condenser	22,800	sq ft each	Shell: C.S	
		33.7	MMBtu/hr each	Tubes: 316 SS	
E-202A-F	Converter-2 feed/converter-2 product heat-exchanger	33,900	sq ft each	Shell: C.S	
		81.7	MMBtu/hr each	Tubes: 316 SS	
E-203A-F	Converter-2 product partial condenser	24,500	sq ft each	Shell: 316 SS	
		40.8	MMBtu/hr each	Tubes: 316 SS	
E-204A&B	Methano product/crude Methanol heat-exchanger	24,800	sq ft each	Shell: C.S	
		15.2	MMBtu/hr each	Tubes: 316 SS	
E-205	C-201 column condenser	9,400	sq ft	Shell: C.S	
		8.4	MMBtu/hr	Tubes: SS	
E-206	C-201 column reboiler	22,950	sq ft	Shell: C.S	
		111.0	MMBtu/hr	Tubes: 316 SS	
E-207A-F	C-202 column condenser/ C-203 column reboiler	17,150	sq ft each	Shell: C.S	
		36.0	MMBtu/hr each	Tubes: 316 SS	
E-208A-F	C-202 column reboiler	16,800	sq ft each	Shell: C.S	
		35.4	MMBtu/hr each	Tubes: 316 SS	
E-209A-F	C-203 column condenser	19,025	sq ft each	Shell: C.S	
		35.6	MMBtu/hr each	Tubes: 316 SS	
E-2010A-H	Converter-1 product Partial condenser	4.940	sq ft each	Shell: C.S	
		39.0	MMBtu/hr each	Tubes: 316 SS	
E-301	Absorber water cooler	87.2	sq ft	Shell: 316 S.S.	
		1.94	MMBtu/hr each	Tubes: 316 SS	
E-302	Rwgs effluent/feed exchanger	555	sq ft	Shell: 316 S.S.	
		40.0	MMBtu/hr each	Tubes: 316 SS	
E-303	1st stage cooler rwgs compressor	1,615	sq ft	Shell: 316 S.S.	
		11.3	MMBtu/hr each	Tubes: 316 SS	
E-304	2nd stage cooler rwgs compressor	1,035	sq ft	Shell: 316 S.S.	
		9.2	MMBtu/hr each	Tubes: 316 SS	
E-305	Furnace fuel gas heater	241.9	sq ft	Shell: 316 S.S.	
		1.35	MMBtu/hr	Tubes: 316 SS	
Pressure vessels					
V-101 & V-102	Desulfurization column	23,500	gal each	C.S.	
V-103	Syngas condensate ko drum 1	7,000	gal	304 SS	
V-104	Syngas condensate ko drum 2	4,000	gal	304 SS	

Table 6.3 Low CO_2 emission methanol production process—Major equipment (continued)

Capacity: ~5,000 metric tons per day (Mtpd) of methanol of 99.85 wt% purity

Plant annual stream factor: 0.9

Equipment number	Name	Size		Material of construction	Remarks
V-105	Steam drum	15,000	gal	C.S	
V-201	Gas-liquid separator	12,550	gal	316 SS clad	
V-202	Gas-liquid separator	17,100	gal	316 SS clad	
V-203	Gas-liquid separator	29,250	gal	316 SS clad	
V-204	Reflux drum	1,600	gal	316 SS clad	
V-205	Reflux drum	28,600	gal	316 SS clad	
V-206	Reflux drum	19,500	gal	316 SS clad	
V-207	Steam drum	24,300	gal	C.S.	
V-208	Steam drum	21,500	gal	C.S.	
Tanks					
T-201A&B	Crude methanol holdup tank	945,750	gal	304 SS	12-hour storage capacity in each tank
T-202A&B	Methanol product holdup tank	875,100	gal	C.S.	12-hour storage capacity in each tank
T-251A-H	Methanol offsite storage tank	1,750,200	gal	C.S	8-day total storage capacity
T-252A&B	Byproducts offsite storage tank	66,000	gal	C.S	
Furnaces					
F-101	Furnace	200	MMBtu/hr each	Cr-Mo coil	Fired by plant generated fuel gas
F-301	Furnace	23	MMBtu/hr each	Cr-Mo coil	Fired by plant generated fuel gas
Package units					
G-101	Air separation unit	2,500	tpd		O_2 purity is close to 99%
G-102	Hydrogen production unit	160,600	lb/hr		H_2 extraction is done at 325–350 psia through PSA. Extraction rate = 27,690 lb/hr

Pumps

Section	Operating	Spares	Operating bhp	
100	4	4	2,030	Includes power for BFW pump/s
200	11	11	906	

Source: IHS Markit

Table 6.4 Methanol production by Johnson Matthey/Davy two-stage reforming-based process—Utilities summary

Capacity: ~5,000 metric tons per day (mtpd) of methanol of 99.85 wt% purity

Plant annual stream factor: 0.9

	Units	Battery limits total	Section 100	Section 200	Section 300
Average consumptions					
Cooling water	gpm	97,332	36,170	56,630	4,532
Electricity	kW	7,975	6,841	1,134	-
Peak demands					
Cooling water	gpm	116,798	43,404	67,956	5,438
Electricity	kW	8,135	6,925	1,210	-

Note: No net natural gas is consumed in the process as a utility, because the furnace duties are met using fuel gas generated in the process.

Source: IHS Markit

Cost estimates

The cost estimates are based on the process scheme described above and on the following assumptions:

- PEP Cost Index is 1,237 (second quarter of 2021)
- Plant is a grassroots project
- Stream factor is 0.90
- Construction is overnight at a US Gulf Coast location
- License fee or royalty is not included in estimates
- Plant front-end costs are not included in our estimates. These include project feasibility study and engineering design costs, government permits costs, land cost, land development costs, technology license and royalty fees, working capital, initial catalyst-charge costs, spare parts cost, plant start-up costs, staff training cost, project financing and borrowing costs, and more

Fixed capital costs

Table 6.5 shows that the estimated TFC investment for a grassroots plant producing 5,000 mt/d of methanol with 99.85 wt% purity is about $1.75 billion—the cost has been rounded. This is equivalent to $958 per annual metric ton of methanol product ($1.75 × 10^9)/(5,000 × 365 tons). BLI accounts for approximately 71.4% of the TFC and stands at about $1.25 billion (rounded cost). The BLI includes a contingency allowance of 25%. Battery limits equipment FOB cost is about $603 million, including the packaged cost of ASU and hydrogen separation unit (pressure swing adsorption). The ASU, reactors, heat exchangers, compressors, and pressure swing adsorption unit are the major cost components in battery limits equipment.

Direct installation costs of FOB equipment—including materials and installation labor costs for instrumentation, electrical equipment, piping, fittings, valves, insulation, support and structural materials, paint, etc—stand at about $308 million. Indirect costs are equivalent to about $295 million, and about $109 million is earmarked for unscheduled equipment.