Dimethyl Ether Synthesis Plant Preliminary Design, Safety, Operability and Feasibility Study

Group B05 – Periwinkle Consulting

Cover Letter (Executive Summary)

Dear Dr. Nease,

Periwinkle Consulting was retained by Ethereal Energies (EE) to conduct preliminary design, safety, operability and techno-economic analyses on building a plant for the production of dimethyl ether (DME) from methanol. The final report presented below provides an overview of the modified process design for better operability along with a detailed economic analysis followed by discussion of safety systems in place.

Plant modifications aimed to improve the process reliability, operability and flexibility were made to allow the process to better respond to unexpected events and minimize the potential for a shut-down. The reliability of the plant was improved through the addition of redundant sensors to account for potential malfunction of instrumentation. Implementation of backup measures such as parallel configuration, bypasses and swing tanks were added to the system to improve flexibility to ensure alternate pathways due to unexpected changes or equipment failure. The operating window of one of the major units was further analyzed in more detail to determine both optimal and normal operating limits.

Safety consideration is integral to any plant design, so the six layers of safety, pertaining to the modification discussed included basic process control, alarms, safety interlock systems, pressure relief, containment, and emergency response. A HAZOP study was conducted in greater detail on the node between T-201 and E-205 to identify potential process risks and include safeguards to prevent or mitigate any severe consequences.

Economic analysis was done on process equipment and operation, ultimately deeming the process financially feasible due to a positive NPV of approximately $128M with a plant life of 25 years. Sensitivity analysis indicated that MARR has the highest impact on NPV, hence the plant being the most sensitive to changes in MARR. Ultimately, Periwinkle Consulting recommends proceeding with the project due to the positive NPV and the quick payback period.

Sincerely,

Group B05 (Periwinkle Consulting)

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1 Introduction

Ethereal Energies (EE) has retained Periwinkle Consulting (PC) to conduct a preliminary design, safety, operability, and techno-economic analysis of a chemical plant producing 50,000 metric tonnes of Dimethyl Ether (DME) at 99.5 weight% purity per year. DME is produced by a catalytic dehydration of methanol on an alumina catalyst at temperatures greater than 250°C but less than 400°C as depicted in equation 1 [1].

$$2CH_3OH(g) \rightleftharpoons \underset{\textit{dimethyl ether}}{H_3C{-}O{-}CH_3(g)} + H_2O(g) \quad (1)$$

As the need for cleaner alternative fuel technology increases globally, DME has been a strong candidate as a synthetic replacement for diesel fuel in compression ignition engines [2]. With a high cetane number from its low auto-ignition temperature and high oxygen content it can burn well and reaches a superheated state that releases a significant amount of energy, comparable to that of diesel [3]. Furthermore, it can be used to replace liquified petroleum gas as a residential fuel for heating and cooking. This implies that the switch to DME will not only be able to be used as a fuel for vehicles but also for residential settings [4].

Transport Canada commissioned the National Research Council Canada (NRC) to study the production, distribution, handling and use of DME in vehicles [5]. Based on the findings, DME can be produced anywhere using renewable products, providing significant flexibility as facilities can be far from crude oil sources [5]. Transport Canada also found that DME powered vehicle can pass any emission standard in the world, in addition DME fueled vehicles burn fuel more efficiently, are quieter and produce less knocking during engine startup [5]. Based on the 2016 study, approximately 5 to 9 million tons of DME are produced yearly [5]. As there are currently no DME fuel stations in Canada [5], this is a great opportunity for EE to spearhead a strong hold on the DME alternative fuel market.

2 Process Overview

The production of DME is significantly less complicated than most fuel production techniques. The process starts with the introduction of high purity methanol into the system. The methanol is heated so that it is vaporized before it reaches the catalyst filled packed bed reactor (PBR) that produces DME at 80% conversion. The DME leaves the reactor at an approximate temperature of 370°C, and is cooled down through a series of heat exchangers that produce medium-pressure steam and low-pressure steam to offset utility costs. The cooled DME reaches the first distillation column which separates DME from unreacted methanol and water. Pure DME is collected at the condenser as a pressurized liquid which is then stored prior to distribution. The methanol and water collected at the bottoms are sent to a second distillation column to separate the two components. The water collected at the bottoms is sent to an in-house waste water treatment plant, whereas the methanol recovered is recycled back into the system with a fraction of the recycle being flared out to the atmosphere.

2.1 Purpose of Each Unit

In the proposed DME synthesis plant, there are five main categories of units consisting of distillation columns, reactors, pumps, heat exchangers and holding tanks. Specifications for each piece of equipment corresponding to the categories are included in the tables below. Although there are many variations of equipment, the client specified equipment and specifications have been incorporated.

2.1.1 Distillation Columns

Distillation columns are large towers that provide the physical means to separate liquid mixtures based on the difference in volatility (boiling points) between the feed components. There are numerous types of distillation columns, with different configurations that can successfully achieve the desired separation based on the properties of the feed and the desired products. For DME production and the proposed process herein, a plate/tray column is considered as the preferred separation method [6].

Table 1: *Outline of distillation columns used in the process.*

P&ID Tag	Equipment Type	Purpose	Specifications		
T-201	Plate/Tray Column	Separate fluid mixture containing methanol, water and DME. Achieve liquid 99.5% wt. DME at the distillate.	Material	Carbon Steel	
			# of Trays	30	
			Tray Material	Stainless Steel	
			Tray Diameter	Trays 2-16	Trays 17-29
				0.84 m	0.40 m
			Column Height	27.6 m	
T-202	Plate/Tray Column	Separate fluid mixture containing methanol and water. Achieve as much methanol purity as possible. Process achieved 99%wt liquid methanol.	Material	Carbon Steel	
			# of Trays	35	
			Tray Material	Stainless Steel	
			Tray Diameter	Trays 2-24	Trays 25-34
				0.62m	0.56m
			Column Height	32.5 m	

2.1.2 Reactor

A PBR was chosen as the most appropriate reactor for DME synthesis, due to its capability of handling vapour phase reactions with high conversion rates demands in addition to its low operating costs [7]. To obtain the desired conversions, inefficient and unstable arrangements must be avoided to avoid potential fouling. The reactor is sufficiently insulated to avoid run away reactions. A potential downside, of PBR operation is the poor temperature control, thus the inlet temperature must be controlled such that no run-away reactions occur [7].

Table 2: *Outline of the reactor used in the process.*

P&ID Tag	Equipment Type	Purpose	Specifications	
R-201	Packed Bed Reactor	A cylindrical vessel that provides the physical means necessary for methanol to be converted into DME. The reactor is packed with porous catalyst that accelerates the reaction.	Material	Carbon Steel
			Length Diameter	6.3 m, 0.77 m
			Catalyst	Alumina, 10% silica
			Conversion	80% of MeOH

2.1.3 Pumps

Positive displacement (PD) and centrifugal pumps (CP) were used in this process. A PD pump operates in such a way that it produces a constant flow rate at a given speed no matter the discharge pressure. A PD pump achieves this by trapping a fixed amount of fluid and discharging that fluid into the outlet pipe. Given the operating principles of the PD pumps, relief valves must be placed on the discharge side of the pump to avoid an accumulation in pressure that could cause the line to burst [8].

PD pumps were chosen to allow for a consistent flowrate throughout the process to ensure that steady state operation is achieved. This is crucial, particularly for the reactor as the PBR conversion rates is sensitive to flowrate inputs. Centrifugal pumps (CP) on the other hand, transfer rotational energy into the moving fluid. The fluid enters the pump and is whirled radially by the impeller until the fluid leaves the pump [8]. CPs can provide much higher flowrates compared to PD pumps [9]. The flow rates can also be throttled by placing a valve downstream of the pump without having excessive pressure buildup, meaning no relief valves are required. CP pumps are chosen to control the reflux rate for the distillation towers as the CP allow for a more flexible operation due to possible changes in pressure and temperature of the column.

Table 3: *Outline of the pumps used in the process.*

P&ID Tag	Equipment Type	Purpose	Specifications	
P-201 A/B	Positive Displacement	Controls the amount of methanol (held constant) at the beginning of the process.	Material	Carbon Steel
			Motor	Electric
P-202 A/B	Centrifugal	Controls the amount of the light product that is liquid going back into T-201.	Material	Carbon Steel
			Motor	Electric
P-203 A/B	Centrifugal	Controls the amount of the light product that is liquid going back into T-202.	Material	Carbon Steel
			Motor	Electric
P-204 A/B	Positive Displacement	Controls the amount of methanol being recycled back to the process.	Material	Carbon Steel
			Motor	Electric

2.1.4 Heat Exchangers/Fire Heaters

All the heat exchangers (HXs) included in this process are of the shell and tube (S/T) type. This type of HX is widely used in industry due to its flexibility to operate at wide range of temperature and pressures [10]. This HX consists of a determined number of tubes placed inside a cylindrical shell. Heat is transferred, as one fluid passes through the tubes while the other fluid flows outside of the tubes. The two variations of S/T used in this process are fixed and floating head. The floating head allows for the tube bundle to be pulled out for cleaning on the tube side and shell side of the process while the fixed head style is not able to do this [10]. Table 4 highlights the main HX used in the process.

Table 4: *Outline of the heat exchangers used in the process*.

P&ID Tag	Equipment Type	Purpose	Specifications	
E-201	Utility Heat Exchanger	Vaporizes liquid methanol before it reaches the reactor. Overcomes the phase change from liquid to vapor methanol.	Material	Carbon Steel
			Actual exchanger area	177.3 m^2
			Utility	MPS
			Heat Duty	4196 kW
E-202	Process-Process Heat Exchanger	Energy recovery by using the hot outlet stream of the reactor to further heat the vapour methanol.	Material	Carbon Steel
			Actual exchanger area	8.9734 m^2
			Utility	Outlet Stream of Reactor
			Heat Duty	175722 cal/sec
E-203	Utility Heat Exchanger	Cools the process to produce MPS	Material	Carbon Steel
			Actual exchanger area	15.9859m^2
			Utility	MPSG
			Heat Duty	120996 cal/sec
E-209			Material	Carbon Steel

	Utility Heat Exchanger	Cools vapour methanol entering the reactor, keeps the inlet to the reactor constant and produces MPS.	Actual exchanger area	3.80369 m^2
			Utility	MPSG
			Heat Duty	75856.6 cal/sec
E-210	Utility Heat Exchanger	Cools the process line before the entering T-201 by producing LPS.	Material	Carbon Steel
			Actual exchanger area	103.824 m^2
			Utility	LPSG
			Heat Duty	482258 cal/sec
FH-201	Utility Fire Heater	Used during start-up achieve the appropriate temperature for the reaction to occur	Material	Carbon Steel
			Heat Duty (at start-up)	545.75 kW
			Utility	Natural Gas

2.1.5 Holding Tank Vessels

There are two types of holding tanks used in this process, holding and condenser tanks. A holding tank was implemented at the beginning of the process to hold methanol, and another holding tank was implemented before each distillation column. Holding tanks allow for the inlet process to remain constant while being able to control the flow downstream. Holding tanks hold a fluid as stock in case the feed were to decrease from steady state, so the process can still produce the appropriate amount of DME for a fixed period. Condenser tanks were also implemented to ensure the vapour fully condenses, and also holds product in case the process deviates from steady state operations.

Table 5: *Outline of the hold tank vessels used in the process.*

P&ID Tag	Equipment Type	Purpose	Specifications	
V-201	Holding Tank	Cylindrical tank that provides a retention time to regulate the amount of methanol going into the process.	Material	Carbon Steel
			Volume	3490 Liters
V-202	Condenser Tank	Used as a total condenser to ensure all light key (DME) product condenses fully to liquid in T-201. Tank is of cylindrical geometry	Material	Carbon Steel
			Volume	2107 Liters
V-203	Condenser Tank	Used as a total condenser to ensure all light key (methanol) product condenses fully to liquid in T-202. Tank is of cylindrical geometry	Material	Carbon Steel
			Volume	1413 Liters
V-204	Swing Vessel	Used if the process deviates from steady state such that T-201 requires partial shutdown till operating condition are changed. Tank is of cylindrical geometry	Material	Carbon Steel
			Volume	3490 Liters
V-205	Swing Vessel	Used if the process deviates from steady state such that T-202 requires partial shutdown till operating condition are changed. Tank is of cylindrical geometry	Material	Carbon Steel
			Volume	3490 Liters

2.2 Alternative Technologies

As need for alternative fuel technology gets increasing attention, specifically to be used in the heavy-duty road sector in Canada, process intensification to produce DME is vital. Currently the designed production of DME from Ethereal Energies is by using the conventional method of dehydration of methanol using solid-acid catalyst, with alumina and 10% silica. However, this process is not ideal as it will cost a lot more to manufacture since methanol itself is an expensive chemical feedstock.

Fortunately, the production of DME is flexible as renewable feedstock can be used produce synthesis gas, which is converted to DME. In a 2008 research study demonstrated the gasification of feedstock called Black liquor (BL). The objective of this study was to demonstrate environmentally optimized biofuel for road transport. Black liquor is a bi-product of the pulping process that is gasified to synthesis gas and then converted to DME, known as BioDME [11].

2.3 Detailed P&ID

Please see attached P&ID document, the process outlets and inlets under normal optimal conditions can be seen in table 6 below.

Table 6: *Describes the process capacity of the DME plant, including temperature, pressure, and mass flow rates of major inlet and outlet streams.*

Stream	**S1 (Inlet to V-201)**	**S23 (Methanol Recycle)**	**S13 (DME outlet)**	**S22 (Water)**	**S24 (To Flare)**
Temperature (°C)	25	26	45	25	110
Pressure (bar)	1	15	10	1	5
Phase	Liquid	Liquid	Liquid	Liquid	Mixed
Methanol (Metric - tonnes/year)	78,900	20,000	280	258	20
Water (Metric – tonnes/year)	0	41	0	22,000	0.04
DME (Metric- tonnes/year)	0	120	56,300	0	0.12

2.4 Process Differences & Redundancies

The overall process was modified from the one proposed as per Project Memo #1 to achieve better operability, reliability, profitability, all while keeping safety in mind. ASPEN-Plus was utilized to simulate various process scenarios. Two process alternatives were identified as potential candidates. The first alternative's process change was the implementation of a process-process heat exchanger (PPHX) and a furnace (fire heater) utility before the reactor. This alternative produced the desired DME throughput, but it was an expensive process due to the constant use of the fire heater, which used natural gas as a utility. The PPHX was not able to reduce the use of the fire heater due to the physical constraints presented by the phase change experienced in the HX. Thus, a second alternative reducing the use of the fire heater was evaluated. This alternative consisted of a HX pre-heating the methanol in (E-201) beyond the phase change profile (i.e 160°C), then further heating the methanol using the PPHX (E-202) up to 280°C, and lastly cooling the methanol up to 250°C using another HX (E-209) that produces MPS. This alternative only requires the use of the furnace at start-up, and then turned off once the reactor gets hot enough. By setting the outlet of the cold stream of E-202 to 280°C, the outlet of the hot stream leaves at 244°C which lets E-203 and E-210 cool down the DME and produce MPS and LPS respectively to offset utility costs.

3 Economics

Several assumptions were required to be able to calculate the capital costs, depreciation, revenues and operating costs for this proposed plant. Some of the most important assumptions included assuming the demand and cost of DME remained constant over the course of its 25-year lifespan. However, in case this wasn't true, a sensitivity analysis was conducted to determine how the profitability would change due to fluctuations in certain variables. Another important assumption was that our plant will be an addition to an already existing plant built within an industrial park to efficiently buy and sell utilities/resources. Should there be any problems in the plant or fluctuation in resource pricing/demands, this plant produces a surplus of 6,300 metric tonnes as contingency that is stored until it can be sold.

3.1 Typical Operating Costs

Operating costs can vary yearly, however based on an assumption of constant price over the years it was determined that a yearly operating cost is approximately $17,515,485. The yearly operating cost can be broken down into categories. It consists of various utilities, catalyst and methanol to produce DME. Furthermore, plant operators and by extension engineers/managers are included in the yearly operating costs through the assumption of hiring 4.5 for every required operator making $60,000/year thereby ensuring every worker is paid and the plant is not understaffed at any given moment. Table 7 identifies the pricing of the utilities calculated for the operation of the plant and the quantity required per year. The production and sale of produced utilities are not considered part of the operating cost and instead contribute to the revenue. Table 8 identifies the pricing of the resources required to produce DME. It is assumed that any water required will be obtained from a lake and filtered by an existing inhouse water filtration system thereby resulting in no additional cost. Natural gas is only used for the startup procedure to get the plant up to the correct temperature. After start up is complete, the fired heater is turned off and therefore natural gas is no longer required. Based on an assumption of requiring a single startup / shut down, per year excess Natural gas is not required. Furthermore, the cost of methanol and catalyst is expected to remain constant over the years.

Table 7: *Utilities utilized by the designed plant including the quantities and year price in 2017 Canadian dollars.*

Utilities Used [12]	Price (2017 CAD)	Quantity per Year	Units	Yearly Price
Medium Pressure Steam	$ 0.0138	101056000	kg	$ 1,394,952
Cooling Water	$ 0.00002584	6649188000	L	$ 171,846
Boiler-Feed Water	$ 0.0006	33314640	L	$ 20,664
Natural Gas - Start Up Only	$ 0.1843	79238	kg	$ 14,601
Hydro	$ 75.00	115	MW-hr.	$ 8,625

Table 8: *Resources required including raw materials and manpower to run the plant each year in 2017 Canadian dollars.*

Resources	Price (2017 CAD)	Quantity per Year	Units	Yearly Price
Catalyst	$ 4.5000	9935	MT	$ 44,709
Methanol [14]	$ 191.00	78,894	MT	$ 15,068,754
Operators & Supervisor	$ -	-	-	$ 791,335

3.2 Detailed Estimates of Capital

Based on the provided values of MARR being 18%, debt being 8% and the assumption that 75% of the financing comes from Equity and 25% comes from debt the WACC could be calculated. Furthermore, we are assumed to be taxed 34% as per 4N4 lectures. This allows for the calculation of the weight average capital cost as per equation 2 resulting in 14.82% [15].

$$WACC = (Equity \times MARR) + (Debt_1 \times Debt_2)(1 - Tax) \quad (2)$$

Capital costs were calculated using two separate methods for comparison. Aspen Capital Cost Estimator ("Icarus") and Donald Woods' "Cost Estimation for the Process Industries" text were used and converted to present day Canadian dollars. This comparison identified the shortcomings of Woods estimations as they were limited to certain costing restraints whereas Icarus allowed for the identification of all variables that could potentially result in additional cost for equipment / installation. It was determined that Icarus is the more reliable cost estimator due to all the variables that can be specified and costed accordingly. Valves and pipes were considered as part of the bare module cost for each piece of equipment. Explicit costs for various valves, sensors, alarms and solenoids were not included. Equipment sizing was based on results provided by Aspen Plus and the clients specifications provided at the start of the project. Table 9 contains 2017 Canadian dollars pricing determined by Icarus and Woods as well as the error between the values assuming Icarus is the more accepted value.

Table 9: *Equipment required for the designed plant including a comparison of pricing in 2017 Canadian dollars obtained from two methods (Icarus and Woods) and the error calculated between the costed values.*

Equipment	Icarus Price (CAD/unit)	Woods Price (CAD/Unit)	Error (%)
V-201	$ 118,100	$ 29,912	75%
P201A/B - no motor	$ 119,400	$ 48,242	53%
P201A/B - motors	$ -	$ 8,306	
E-201 (MPS USED)	$ 205,600	$ 184,279	10%
E-202 (Main HX)	$ 136,500	$ 22,202	84%
FHEAT1	$ 351,142	$ 84,325	76%
E-209 (MPS GEN)	$ 73,900	$ 8,814	88%
R-201	$ 293,900	$ 23,178	92%
E-203 (MPS GEN)	$ 112,600	$ 46,922	58%
E-210 (LPS GEN)	$ 117,700	$ 105,211	11%
T-201 (TOWER)	$ 489,500	$ 372,169	24%
T-201 (E-205) [COND]	$ 134,800	$ 141,964	5%
T-201 (P-202 A/B)	$ 115,400	$ 28,795	72%
T-201 (P-202 A/B) - Motor	$ -	$ 3,306	
T-201 (V-202)	$ 125,500	$ 28,480	77%
T-201 (E-204) [REB]	$ 105,200	$ 52,195	50%
T-202 (TOWER)	$ 568,000	$ 308,784	46%
T-202 (E-207) [COND]	$ 106,100	$ 39,745	63%
T-202 (P-203 A/B)	$ 89,400	$ 20,983	76%
T-202 (P-203 A/B) - Motor	$ -	$ 867	
T-202 (V-203)	$ 109,900	$ 13,457	88%
T-202 (E-206) [REB]	$ 141,700	$ 139,666	1%
E-208 (cooling WW)	$ 85,800	$ 28,288	67%
E-211 (cooling recycle)	$ 79,400	$ 20,511	74%
P-202 A/B (recycle)	$ 75,000	$ 24,022	65%
P-202 A/B (recycle) - Motor	$ -	$ 1,900	
V-202	$ 118,100	$ 29,912	75%
V-203	$ 118,100	$ 29,912	75%

V-204	$ 118,100	$ 29,912	75%
V-205	$ 118,100	$ 29,912	75%
Capital Cost	**$ 4,226,942**	**$ 1,906,171**	**55%**

Sample calculations for V-201, R-201, E-202, T-201, P-201 A/B are presented below using Woods method. Icarus uses internal calculation methods and therefore is unable to be presented in the sample calculations.

P-201A/B – Woods Calculation $C_0 = \$1{,}500\ ;\ F_{IF} = 1.75\ ; F_M = 1.55\ ; F_P = 1\ ; n = 0.52\ ; f_p = 0.6\ ; \psi = 0.7;\ CEPCI_{1970} = 123.8\ ;$

$$CEPCI_{2017} = 591, Err = \pm 40\%\ ; Ref. = 10kW\ ; Desired = 10.3968\ kW\ ; Efficiency = 60\%$$

$$F_{CF} = \left(\frac{\frac{Desired}{Efficiency}}{Reference}\right)^n = \left(\frac{\frac{10.3968kW}{0.6}}{10\ kW}\right)^{0.52} = 1.33091, F_\pi = \frac{CEPCI_{2017}}{CEPCI_{1970}} = \frac{591}{123.8} = 4.77383$$

$$C_{FOB} = C_0 \times F_{CF} = \$1{,}500 \times 1.33091 = \$1{,}996, C_{BM} = C_{FOB} \times F_{IF} = \$1{,}996 \times (1.75 - 1) = \$1{,}497$$

$$C_{FOB,adj} = C_{FOB} \times (F_P F_M - 1) = \$1{,}996 \times (1 \times 1.55 - 1) = \$1{,}098$$

$$C_{BM,Adj} = C_{FOB} \times (F_P F_M - 1) \times f_p \psi = \$1{,}996 \times (1 \times 1.55 - 1) \times (0.6) \times (0.7) = \$461$$

$$BM_{1970} = C_{FOB} + C_{BM} + C_{FOB_{adj}} + C_{BM,Adj} = \$1{,}996 + \$1{,}497 + \$1{,}098 + \$461 = \$5{,}053$$

$$BM_{2017} = BM_{1970} \times F_\pi = \$5{,}053 \times 4.77383 = \$24{,}121 \pm 40\%$$

R-201 – Woods Calculation

$$C_0 = \$1{,}900\ ;\ F_{IF} = 3\ ; F_M = 1\ ; F_P = 1\ ; n = 0.62\ ; f_p = 0.6\ ; \psi = 0.7;\ CEPCI_{1970} = 123.8\ ;$$

$$CEPCI_{2017} = 591, Err = \pm 50\%\ ; Ref. = 3.8m^3\ ; Desired = 2.9336742m^3$$

$$F_{CF} = 0.851784,\ \ F_\pi = 4.77383, C_{FOB} = \$1618, C_{BM} = \$3{,}237, C_{FOB,adj} = \$0$$

$$C_{BM,Adj} = \$0, BM_{1970} = \$4{,}855, BM_{2017} = \$23{,}178 \pm 50\%$$

E-202 – Woods Calculation $C_0 = \$8{,}000\ ;\ F_{IF} = 3.14\ ; F_M = 1\ ; F_P = 1.0509\ ; n = 0.71\ ; f_p = 0.6\ ; \psi = 0.7;\ CEPCI_{1970} = 123.8\ ;$

$$CEPCI_{2017} = 591, Err = \pm 40\%\ ; Ref. = 100m^2\ ; Desired = 9m^2$$

$$F_{CF} = 0.1809307\ ,\ \ F_\pi = 4.77383\ , C_{FOB} = \$1{,}447\ , C_{BM} = \$3{,}098\ , C_{FOB,adj} = \$74\ \ , C_{BM,Adj} = \$32$$

$$BM_{1970} = \$4{,}651\ , BM_{2017} = \$22{,}202 \pm 40\%$$

T-201 – Woods Calculation – Only Column included in cost to reduce space taken up by calculations $C_0 = \$65{,}000\ ;\ F_{IF} = 4.16; F_M = 1\ ; F_P = 1.0; n = 0.53\ ; f_p = 0.6\ ; \psi = 0.7;\ CEPCI_{1970} = 123.8\ ;$

$$CEPCI_{2017} = 591, Err = \pm 40\%\ ; Ref. = 100m^2\ ; Desired = 9.6m^2$$

$$F_{CF} = 0.288\ , F_\pi = 4.77383$$

$C_{FOB} = \$18{,}740, C_{BM} = \$59{,}220$

$C_{FOB,adj} = \$0, C_{BM,Adj} = \$0, BM_{1970} = 77{,}960, BM_{2017} = \$372{,}169 \pm 20\%$

<u>V-201 – Woods Calculation</u>

$C_0 = \$1{,}900;\ F_{IF} = 3; F_M = 1; F_P = 1; n = 0.62; f_p = 0.6; \psi = 0.7;\ CEPCI_{1970} = 123.8;$

$CEPCI_{2017} = 591, Err = \pm 50\%\ ; Ref. = 3.8m^3\ ; Desired = 4.42665m^3$

$F_{CF} = 1.099261$, $F_\pi = 4.77383$, $C_{FOB} = \$2{,}089$, $C_{BM} = \$4{,}177$, $C_{FOB,adj} = \$0$ $C_{BM,Adj} = \$0$
$BM_{1970} = \$6{,}266$, $BM_{2017} = \$29{,}912 \pm 50\%$

3.3 Profitability Analysis

The profitability was analyzed using net present value calculations over a span of 25 years. Based on the results generated, this plant would be profitable within a year. Revenues were acquired from the generation of low pressure steam, medium pressure steam and the sale of dimethyl ether for a total of $43,862,351 per year as seen in Table 10. The capital cost of the whole plant is depreciated at the same rate, under the assumption that all equipment depreciates at the same rate and never requires replacing. CRA class 43 was determined to be the class of this plant which resulted in a depreciation rate of 30%. Due to the first year 50% depreciation rate rule, the capital cost was only depreciated 15% the first year and 30% the following years. As per the previously calculated WACC of 14.82%, this plant will generate NPV as seen in Table 11. Additional years in service provide more income though in 2017 dollars it will be worth little. Taxes were taken into account at 34% as certain plant costs are taxable and therefore require to be paid. Though there is some error in the CAPEX as is often the case in scenarios where simulations and outdated resources come into play, based on the amount of Canadian dollars generated through production, this error will have little to no effect on the profitability of the venture as can be seen in Table 11 below.

Table 10: *Revenue generation in the form of quantity and yearly income in 2017 Canadian dollars from various utilities and DME in the proposed plant*.

Revenues [12]	**Price (2017 CAD)**	**Quantity per Year**	**Units**	**Yearly Income**
Low Pressure Steam Gen.	$ 0.0086	22635000	kg	$ 195,280
Medium Pressure Steam Gen.	$ 0.0138	2889640	kg	$ 39,888
Dimethyl Ether [16]	$ 770.88	56,594	MT	$ 43,627,183
Total Revenue	-	-	-	$ 43,862,351

Table 11: *The net present value of the revenue, expenses, depreciation and taxation each year for the proposed plant in 2017 Canadian dollars over the course of 30 years at specific periods.*

Year (n)	Revenue ($)	Eligible Expenses ($)	Ineligible Expenses ($)	BV of Assets ($)	Depreciation ($)	Taxable Income ($)	Tax Paid ($)	Net Cash Flow ($)	PV of Cash Flow ($)	Cumulative PV of CF ($)
0	43,862,351	17,515,485	4,226,942	4,226,942	718,580	25,628,285	8,713,617	13,406,306	13,406,306	13,406,306
1	43,862,351	17,515,485	-	3,508,362	1,192,843	25,154,022	8,552,368	17,794,498	15,497,734	28,904,040
2	43,862,351	17,515,485	-	2,315,519	787,276	25,559,589	8,690,260	17,656,605	13,392,823	42,296,862
3	43,862,351	17,515,485	-	1,528,242	519,602	25,827,263	8,781,269	17,565,596	11,604,068	53,900,930
4	43,862,351	17,515,485	-	1,008,640	342,938	26,003,928	8,841,335	17,505,530	10,071,753	63,972,684
5	43,862,351	17,515,485	-	665,702	226,339	26,120,526	8,880,979	17,465,886	8,751,911	72,724,595
10	43,862,351	17,515,485	-	83,368	28,345	26,318,520	8,948,297	17,398,568	4,368,558	102,106,871
15	43,862,351	17,515,485	-	10,440	3,550	26,343,315	8,956,727	17,390,138	2,187,963	116,808,501
20	43,862,351	17,515,485	-	1,307	445	26,346,421	8,957,783	17,389,082	1,096,291	124,173,939
25	43,862,351	17,515,485	-	164	56	26,346,809	8,957,915	17,388,950	549,332	127,864,572
30	43,862,351	17,515,485	-	21	7	26,346,858	8,957,932	17,388,933	275,262	129,713,892

3.4 Detailed Sensitivity and Risk Analyses

Sensitivity and risk analyses were performed using Excel data tables for various variables that were likely to come under scrutiny over the course of the years. Most notably the prices of DME, methanol, electricity, MARR, and catalyst. Based on the results of these variables, figure 1 was created to examine the sensitivity of each variable on NPV. The steeper the slope, the more sensitive the variable. Due to the low amount of electricity used, price fluctuations did not make a significant difference. The price of methanol had some effect however due to its average price being 191$/MT it did not affect the NPV greatly. The price of DME had a large effect due to its price of 771$/MT with fluctuations in price either greatly increasing or decreasing profits. MARR change was determined to have the largest effect on the NPV due to it being a percentage with high influence on all calculations.

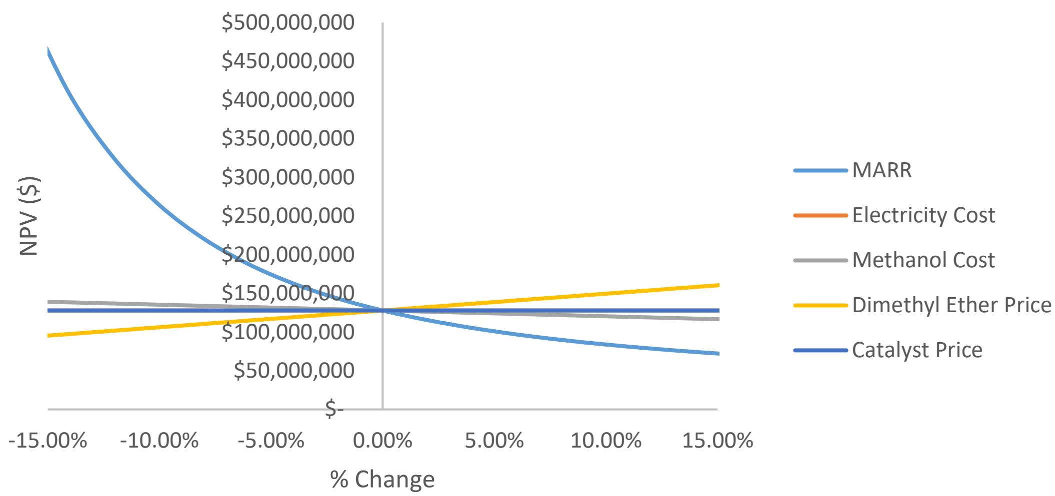

***Figure** 1: A spider plot examining the effects of electricity cost, methanol cost, catalyst price, MARR and dimethyl ether price on the net present value of the proposed plant.*

The present value of cash flow plateaus because as the net cash flow begins to reach a constant yearly value due to all equipment having been depreciated the only change that affects the cash flow is the present-day value of it. Present day value of money decreases with time and therefore contributes less and less to the cumulative present value of cash flow until in today's day the money is worth very little. Therefore a 25-year life span for this plant will ensure the maximum returns on investment. This trend can be observed in figure 2 below. The trend will only change if significant changes are made to the plant later in time (such as replacing equipment or installing new equipment).

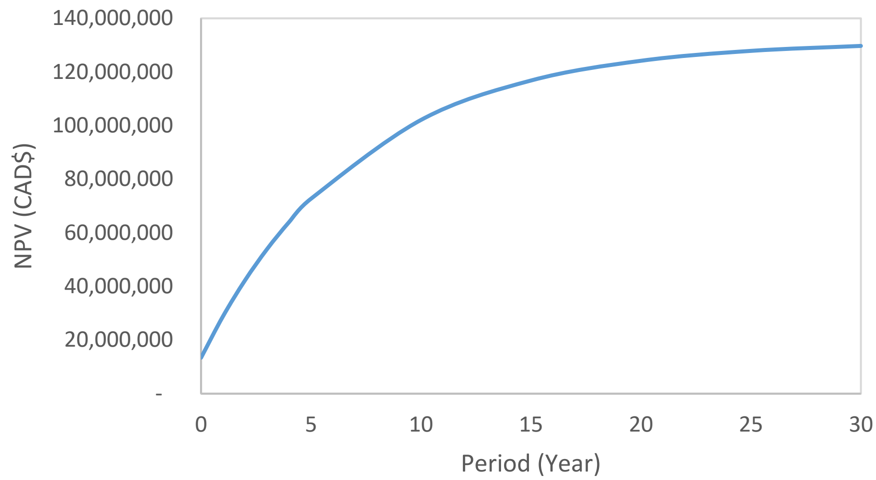

***Figure** 2: The cumulative present value of cash flow over the span of 30 years. It can be observed that the NPV trend is non-linear and begins to plateau around the 30-year mark.*

4 Operability

4.1 Operating Window for Reactor Unit

The operability of the reactor is essential to the overall process therefore steady state conditions are required for proper operation. The operating window of the reactor can be seen below in figure 3. The temperature and pressure of the inlet stream were examined.

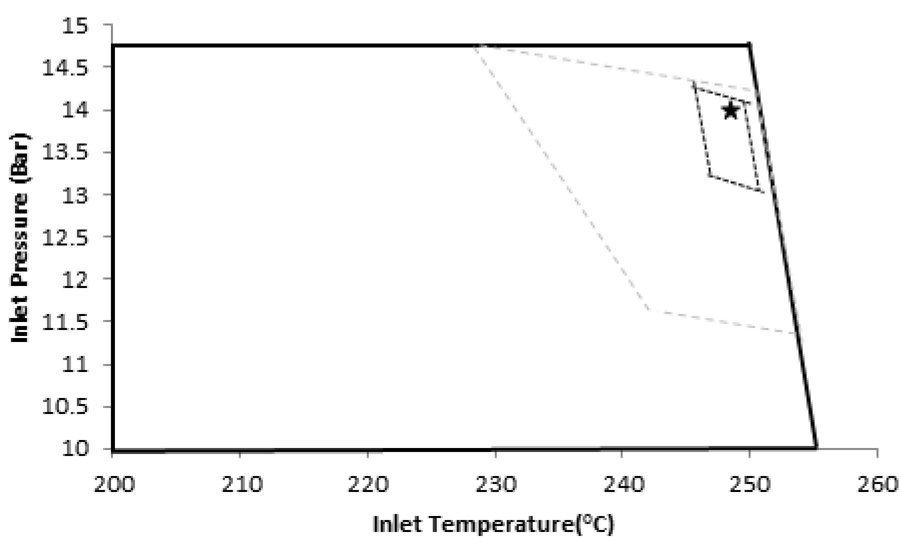

Figure 3: *Operating window of reactor R-201. Black solid lines represent hard constraints, grey dashed lines represent the feasible operation range, back dashed lines represent the preferable operation range and the black star represents the optimal operation point.*

Three operating regions are plotted to represent three states that may occur during operation of the plant. The first region marked by the black lines is defined by the hard constraints of the system, the pressure cannot exceed 14.7 bar and drop below 10 bar and the inlet temperature can't exceed

255°C since the exothermic reaction will increase the temperature of the reactor above 400°C which would cause damage. This region is not a simple rectangle since inlet pressure and inlet pressure both have a positive effect on the reactor temperature and therefore as the pressure increases the temperature of the inlet must decrease to ensure the reactor temperature does not exceed 400 °C. The second operating region marked by the grey dashed lines is where the plant is able to operate but at a lesser efficiency. The reflux, boil up and recycle rates must be adjusted in order to reach the proper percentage weight percentage of DME. The window was constructed by varying temperature and pressure rates in multiple simulations while ensure proper separation of DME is still possible. The last region that is defined by a black dashed line represent the operating window that can exists with minimal changes from the process this region again was defined by multiple simulations by varying temperature and pressure but not allowing the reflux and boil up rates of the distillation columns to change. However, the recycle rate was manipulated to ensure the proper weight percentage of DME product. Lastly the optimal operating point denoted by the star in the figure represents optimal process variables that the system should run at. These temperature and pressure values were used to size the reactor to ensure 80% conversion of methanol.

4.2 Flexibility and Controllability Analyses

This plant has a 95% capacity factor or 8,320 hours of uptime per year. To ensure that no unnecessary downtime occurs the plant has a few modifications to increase this overall operability. Proper recording and measurement of all process lines not only helps with operators and engineers knowing what is happening with their process but it also allows insight if certain equipment is not preforming properly and allows operators to make appropriate changes before damage and shutdown occur. All process lines have the following measurements: flow, temperature and pressure, the measurement is not required if the process did not flow into a major piece of equipment that would have changed the properties. Flow and pressure measurements on vapour lines can be used as redundant measurements to see if any instrumentation is malfunctioning. Major process equipment such as the reactor and distillation columns have redundant sensors to allow better insight to what is happening within the process equipment. For example, the reactor has multiple temperature sensors to allow for the temperature profile of the reactor to be seen to give insight whether any fouling may have occurred. As for the distillation columns, there are redundancies in the level measurement by differential pressure measurement, level transmitters and low and high-level alarms. All the pumps in the plant have a redundant partner and proper isolation to ensure that if any pumps break they can be isolated and flow can be directed to the redundant partner.

Even though this process is assumed to run at steady state, variation in raw materials, utilities and production rates must be taken into consideration. Variation in methanol flow rate is controlled by the first holding tank V-201. This tank allows for a certain optimal flowrate to enter the reactor which is controlled by the level measurement feeding an electrical signal to a control valve to allow more or less flow. Swing tanks at the bottom of the distillation column also allow for fluid to be held if flow rates change such that the distillation column's operability is affected. The reflux ratio and boil up rate would have to be adjusted to accommodate these changes. Changing the production rates of the process would have to be planned to ensure variables such as reflux rate and boil up rate would be calculated and adjusted ahead of time so the purity of the product remains the same. Abrupt changes in the production rates can be handled by the swing tanks in the process until the process returns to the optimal state or separation parameters are changed accordingly. Variation in the composition of the methanol poses more of an issue. Methanol below 95% by weight must be rejected as proper conversion of DME would be hindered. A way to prevent this from happening is to have density measurement at the inlet to ensure all methanol entering the process is above the percent weight limit. Control valves and loops ensure that with any variation in the utility streams, the flow rate would be directly affected to allow for proper heat transfer to occur.

Conversion of methanol to DME is an exothermic process, to take advantage of this property a process-process heat exchanger was implemented to reduce the load on the utility heat exchangers. Since process-process heat exchangers can be somewhat unreliable especially if the process is not operating at steady state the furnace is available if the temperature of the reaction were to drop. A bypass system is also around this process-process heat exchanger in case the reaction was to go out of control and rise in temperature. Other heat exchanger loops were deemed unfeasible since the cost of addition heat exchangers outweighed the potential cost saving. A methanol recycle stream was added to increase material efficiency of the process. This is possible since only 80% conversion occurs in the reactor. In order to recycle this methanol, proper separation from water which occurs in T-202 is necessary. The recycled stream is put back into the holding tank V-201 so the process remains under control.

4.3 Reliability Improvements

Reliability is often a difficult concept for most engineers, operators and businesses to understand. This is caused by the fact that companies/plants must spend money on items that don’t improve their process in real time. Reliability can be broken down into minimizing these five major factors: materials of construction, unfavorable operating conditions, equipment faults/failures, errors by personnel and external disturbances [17]. In this DME process corrosion is not an issue due to the product (DME) and the reactants (methanol) not being corrosive to carbon steel. All piping and equipment are constructed of carbon steel and the temperature and pressure of the process is also within the limits of carbon steel. Therefore, in terms of reliability, materials of construction are a factor that do not need to be considered. As for unfavorable process conditions holding tanks and control valves were added to allow for the process to be easily manipulated to return the process to favourable conditions. To reduce the errors that could be caused by operators, all instrumentation are indicator transmitters that do not require the operator to zero and adjust manually, unless a dramatic unfavorable event has occurred. To reduce the chances of improper operation, all manual valves and controller valves are constructed with a local and remote position indicator that states the percent open or percent close of the valves. For external disturbance, no reliability additions were added since this plant will be constructed inside of another plant and we assume that this plant would have proper procedures in place already.

The reliability of each major process equipment, instrumentation and valves can be seen below in table 12. These values can be used to quantify whether certain equipment is appropriate to add parallel systems and if equipment should be stocked at the plant in case of failure. For this DME process, due to the low reliability percent and high failure rate of pumps, parallel systems were added such that in an event of a failure the pump could be isolated and a second pump would be able to take the load of the process to prevent the plant from shutting down. All pumps in our process have a parallel pump to increase the overall reliability of the system. Also, to increase the reliability of the pumps, vibration monitoring will be implemented on the pumps which will show operators when the pumps are vibrating too much and when cavitation is present. If operators can see when excessive cavitation is occurring, the integrity of the pump could be saved. As for the other lower reliability values such as control/manual valves and instrumentation two approaches are taken to increase the overall reliability. For manual valves, control valves and instrumentation an inventory should be constructed. Inventory is necessary for a multitude of reasons such as: multiple controllers that require process measurement for proper operation, consistent monitoring of the system is required for safe processes which is the number one priority and proper isolation to prevent leakage and safe operation. The last improvement for reliability occurs with the control valves and the strategy that is implemented is having the ability to isolate the control valve and having a bypass line with a manual value that could be manually operated to the percent opening that the control valve was last at. This prevents the process from shutting down in the event of the control valve malfunctioning and or breaking.

Table 12: *Reliability and failure rate of various process equipment, instrumentation and valves.*

Process Equipment	Failure Rate (per year) [18]	Reliability (%)
Pressurized Vessels	0.0007	99.93%
Heat Exchangers	0.002	99.80%
Distillation Columns	0.026	97.43%
Fired Heaters	0.041	95.98%
Pumps	0.089	91.48%
Isolation Valves	0.438	64.53%
Control Valves	0.687	50.31%
Instrumentation	0.841	43.13%

4.4 Start-Up and Shut-Down Procedure

Start-up and shut down is often the most dangerous time in a plant. This is caused by the fact that the plant is typically not running a "predictable" steady state. A yearly shut down of the plant should be planned lasting up to one week to ensure no serious degradation of major process equipment. This shut down time should be planned along service work that must be completed on instrumentation, valves and depending on the damage that may have occurred in major process equipment. During the shutdown period, normal 12-hour shifts will occur, engineering staff should be doubled during this time and daily meetings should occur with all staff to ensure the safety of all workers.

Start-up Procedure:

The plant is divided into three major sections when starting up. Zone one which includes the process from the methanol source up to and including the reactor. Zone two includes the process from the end of the reactor up to and including the first distillation column. The last zone, zone three includes the outlets of first distillation column up to and including the second distillation column as well as the recycle stream. The start-up procedure is outlined below:

1. Leak Testing: While all vents and drains are closed and the recycle line valves are open, pressurize the process with an inert gas (i.e. nitrogen) up to a pressure that is 1 bar below the safety relief valve set point and monitoring the rate of pressure loss. If no safety relief valves are present pressurize the process to a pressure that is 1-2 bar higher than the normal operating pressure. Each zone should be completed one at a time. During these tests, operators should check: joints, flanges, drains, vents, major equipment, valves and process connections for instrumentation with ultrasonic leak detectors to confirm leak points. If the pressure loss is greater than 0.1 bar the start-up procedure should not

continue until the leaking component found via the ultrasonic leak detector is dealt with and the pressure loss is below 0.1 bar.

2. Pressuring-up/Heating-up: Eliminate the inert gas in the system and depressurize the system process. Starting with the distillation column in zone two, feed the condensable component (methanol) into the column. Slowly increase the temperature and pressure of the feed until the column is operating at typical conditions. This can be achieved via infinite boil-up rate/ reflux ratio. Repeat the same process the distillation column in zone three, however the condensable component would be water. It is important to note that the zones should be treated as separate processes and mixing should not occur in this step. This step is necessary as it reduces the thermal stress that occurs when introducing the feed mixture since liquid flashing can occur [19]. Meanwhile in zone one, methanol should be introduced in 10% increments. During this time, the fire heater's temperature should be stepped in a controlled way until the temperature reaches 275°C. Once the feed reaches 100% flow, reduce the load of the fire heater and direct the flow to the bypass to allow the residual heat from process-process heat exchanger and the heat exchanger using the medium pressure steam will maintain the process at operating conditions. All feed should be recycled back into the holding tanks at the beginning of the process until conversion of DME occurs.

3. Column Operations: Once zone one is up to operating conditions, open the isolation valve, limiting the flow to the second zone in controllable steps (25%, 50%, 75% and 100%). Reduce the reflux ratio and discard of vapor product that is not up to specification. During these steps ensure that all other utilities in zone 1 and 2 are up to the operating conditions. Reduce the boil up rate while opening the isolation valve limiting the valve to the third zone. Once zone two is up to operation conditions and producing the vapor product to specification begin the same steps to the column in zone three. Discard all product vapor that is not up to spec.

4. Steady state: Once all zones are running to specification begin to reduce the methanol flow to the storage tank and introduce the recycling of the methanol. Pay close attention to the composition of the outlet steams of the columns when introducing the recycle. If any stream is off specification adjust the reflux ratio or boil up rate as seen fit until the composition is back to specification.

Shut down procedure:

The zones of the plant remain the same in the start-up and shut down of the plant. The shutdown procedure is outlined below:

1. Temperature of reactor feed: gradually reduce the temperature of the methanol going into the reactor, by reducing the MPS used by E-201. As soon as the temperature drops under 200°C, the exothermic reaction will stop from happening and runaway reactions will be avoided.
2. Feed Flow: Reduce the feed flow of methanol gradually (100%, 75%, 50% and 25%) until the feed is completely stopped. This can be achieved by reducing the discharge pressure of pump in zone one.
3. Column rates: Reduce the column rate in zone two by reducing the temperature (via shutting off the utilities) of the column until most feed entering the column is leaving the column in liquid phase. Repeat the process for the column in zone three, external cooling may be required.
4. Atmospheric pressure: Carefully reduce the pressure of the column in zone three until atmospheric. During this time the pressure in column in zone three should be reduced until both columns are at atmospheric. It is important to note that zone three should always have a lower pressure than zone two.
5. Elimination: Eliminate all undesirable liquid materials appropriately and shut off the recycle stream.

6. Opening to atmosphere: Prepare the system to open to atmosphere by ensuring that the pressure of the entire system is at atmosphere and there is no flow in the process.
7. Atmospheric pressure: Carefully reduce the pressure of the column in zone three until atmospheric. During this time, the pressure in column in zone three should be reduced until both columns are at atmospheric. It is important to note that zone three should always have a lower pressure than zone two.
8. Cleaning: Remove all major equipment and clean as required. Remove all isolation and control valves and re-trim the seats and inspect any damage that may have occurred. Repack the reactor in this time as well. If any valve or major piece of equipment is damaged beyond repair replace before starting up again.

5 Safety

In refinery and petrochemical industries, where the work environment involves complex process equipment, hazardous material, and high temperature and pressure operations, safety is of paramount importance. Consequently, safety measures are in place to protect personnel, prevent damage to equipment, and minimize, if not eliminate, environmental damage by ensuring operation within safe limits. For proper operation of equipment, routine maintenance will be required by experienced personnel to decrease the likelihood of equipment failure. Similarly, routine maintenance and inspection will help identify faults to prevent impending failure or deficiencies that could result in process disruption, such as pump seal leakage or exchanger tube leaks. Consequently, frequent maintenance and inspection personnel training is instrumental in ensuring proper state of equipment and that equipment is replaced when need be before significant damage affects downstream processes. Setting standards for investigating incidents to develop learnings and prevention strategies also aid in preventing future incidents by fostering recommendations that could go towards better practices and operating procedures. The six levels of safety that were examined are as follows: Basic Process Control System (BPCS), Alarms, Safety Interlock System (SIS), Pressure Relief, Containment, and Emergency Response.

5.1 Six Levels of Safety

5.1.1 Basic Process Control System

Process control systems are in place as the first level of safety aims to provide regulatory control of the process variables to maintain process within acceptable operating condition, further increasing stability of the process. A basic control loop works by the controller receiving a signal of the error, the deviation from the desired state (set point), and then a pneumatic signal is sent to a control valve to perform the necessary control action. For example, if the temperature of the process stream going through a utility heat exchanger is lower than the set point, a temperature transmitter sends an electrical signal of the temperature reading to the controller. The controller then sends out an electrical signal of the control action to a temperature relay (TY) which sends a pneumatic signal to the valve manipulating the valve opening of the steam. The corrective action would result in the valve opening more to allow a higher steam flow to heat up the process stream to the desired temperature.

Control valves, operated by pneumatic signals, are designed to fail open or closed with the loss of air pressure due to failure of control loop. The "failure state" is set to that which yields the safest process state. The fail-safe positions of five example control valves are designated as follows:

1. [TV3] - Fail-closed: The higher temperature imposes a more serious safety hazard with increasingly hot streams, this could potentially lead to thermal fatigue of pipes or contribute to runaway reactions downstream at the reactor; therefore the valve is closed to lower temperature.
2. [TV34] - Fail-open: The lower the temperature is, the safer the operating state, so the control valve should fail-safe open to allow for excess cooling duty and yield a cooler process stream
3. [TV7] - Fail-close (control on fuel gas to fired heater). Similar to the above two, the lower temperature is the safer state, so the fuel gas valve is turned shut to ensure fired heater outlet stream does not increase to undesirably high temperatures
4. [LV31] - Fail-open: If the control loop fails, it is safer to fully open the valve to allow drainage of the vessel to methanol recycle, rather than keep have the vessel's level reach its limits
5. [PV17] - Fail-open: Failure of the control system renders the valve fully open to allow for more condensation of the overhead stream; the cooler reflux would then decrease the overhead pressure by decreasing vapour content, lowering it to a safer state, rather than having pressure increase to dangerously high levels

The set points fed to the controller can be dictated by a refinery's optimizer, like a linear programmer (LP), to improve process economics and stability. Typical controlled variables are flow rate, pressure, temperature, and level to regulate the process and keep it within optimal or stable limits. Example of main control loops for major pieces of equipment are outlined in Table 13.

Table 13: *Control objective and loop pairing logic of control loops of major process equipment.*

Valve	**Manipulated Variable**	**Controlled Variable**	**Control Objective and Loop Pairing Logic**
[TV4]	Reactor inlet bypass opening	Reactor effluent temperature	Reactor effluent temperature control Valve opening manipulates how much of the reactor inlet is directed to the reactor feed without undergoing any heating. When the temperature of the reactor outlet is higher than the set point, the by-pass is opened more, so more of the feed is unheated, ultimately resulting in a cooler reactor feed, which translates to a decrease in reactor outlet temperature. Control Valve is fail-open to have a cooler feed to prevent the potential for runaway reactions
[PV17]	Cooling water supply valve opening	Tower overhead pressure	- Tower overhead pressure control Valve opening manipulates the cooling duty by adjusting cooling water supply. If the tower pressure becomes too high because of higher vapour pressure, the valve opens to allow for more cooling, ultimately sending back cooler reflux which results in the condensation of the vapours in the tower, decreasing the vapour pressure. Control valve is fail-open to ensure that maximum cooling duty is supplied in case of control failure to keep down tower pressure to protect the tower from over-pressuring.
[LV20]	Liquid outlet valve opening	Vessel liquid level	- Vessel liquid level control If the liquid level is higher than the set point, the valve manipulating liquid outlet flow opens to "drain" the vessel and lower the liquid to escape.

			Control valve is fail-open to prevent potential flooding or drying out in the event of control system failure which can damage equipment; for example, pump could vibrate and cavitate when no liquid is pumped through.

5.1.2 Alarms

The second level of safety is alarms, which notify operators of unusual plant operation that require attention, such as higher than normal pressure differential across a distillation tower. Set alarms provide both auditory and visual indication, which require manual action from an operator to return the system to desired operating range. Alarms are added throughout the process where appropriate to indicate any significant deviation from safe limits. A summary of the alarms used for towers and vessels are outlined as an example in Table 14.

Table 14: *Alarms used in towers and vessels.*

TAH	The towers and vessels are equipped with high temperature alarms since the high temperatures can lead to over-pressurization and result in potential equipment failure if design temperatures are exceeded. Similarly, high temperature alarms are added to bottom of tower since high temperatures can lead to too much vaporization of the bottoms liquid which could increase upward vapour flow.
PAH/PAL/PDAH	Alarms are added to indicate whether there's too much pressure buildup in the tower/vessel or if there's too low of a pressure, which could lead to weeping in the tower due to minimum vapour flow. Similarly, high pressure differential alarms are added on the towers to indicate whether there might be flooding due to the increased upward vapour flow through the tower that could force the liquid up the column.
LAH/LAL	High level alarms added on the towers and vessels to indicate if liquid fill rates are indicative of flooding. Alternatively, low level alarms are used to indicate emptying of bottom tower or vessel which can cause cavitation of downstream pumps.

5.1.3 Safety Interlock System

SIS is in place in order to prevent the process from reaching "Not to Exceed" limits and restore it to safe, normal operation if safe operating limits (SOLs) were exceeded for a set period of time. The SIS in place are described in Table 15 used to bring process to normal operating limits (NOLs).

Table 15: *SIS description and trigger for each equipment*.

SIS number	**Description**	**Trigger**
SIS 2011 SIS 2021	This interlock is used to prevent the pressure from the tower from reaching dangerously high limits to restore safe conditions. When triggered for "high", the control valve (CV) on the CWS of the overhead condenser will fully open and the CV on tower feed fully closes to prevent any further pressure build up, until tower's pressure reaches NOL, and vice versa.	Tower pressure is higher/lower than SOL for more than five minutes.
SIS 2012 SIS 2022	This interlock is used to prevent flooding of sieve trays indicated by sudden changes in pressure difference. When triggered for "high", the CV on the tower feed fully closes to prevent addition of extra vapour, and CV on MPS flow rate is closed to stop additional vapourization, and vice versa.	Differential pressure is higher than SOL for more than five minutes.

SIS 2013 SIS 2023	This interlock is used to prevent temperature inside tower from exceeding SOL, also contributing to excessive vaporization. When triggered for "high", both CVs on MPS line fully close until safe temperatures are reached, and vice versa.	Tower temperature is higher than SOL for more than five minutes.
SIS 2014 SIS 2024	This interlock is used to prevent the liquid level in the tower from exceeding SOL. When triggered for "high", CV on tower feed fully closes and CV on bottom stream to swing tank fully opens until normal operating range is reached, and vice versa.	Tower liquid level is higher/lower than SOL for more than five minutes
SIS 2041 SIS 2051	This interlock is used to prevent liquid level in vessel from exceeding SOL. When triggered for "high", CV on vessel's liquid outlet fully opens and vice versa.	Vessel liquid level is higher/lower than SOL for more than five minutes
SIS 2001	This interlock is used to prevent fired heater outlet temperature from exceeding SOL. When triggered for "high", both CVs on fuel gas stream fully close, and vice versa.	Fired heater outlet temperature is higher than SOL for more than five minutes.
SIS 2000	This interlock is used to prevent reactor temperature from exceeding SOL. When triggered for "high", CV on reactor feed fully closes, and vice versa.	Reactor temperature is higher than SOL for more than five minutes.

5.1.4 Pressure Relief

Pressure can dramatically increase due to unexpected disturbances, so process equipment will need to be protected from hazards of high pressure (or low due to vacuum): examples of pressure buildup can include fire exposure, pressure accumulation of vapour build-up, blocked discharge, or thermal expansion of liquid in a pipe. Pressure-relief devices are used as self-actuating devices to provide an exit path for the fluid to bring the process back to safe operating condition. The type of pressure relief device used is a pressure relief valve (PRV), which are spring loaded valves that pop open to route the fluid to a header, typically flare, when the accumulated pressure reaches the set pressure of the PRV. The advantage of the PRV is that it remains open until pressure drops below the set pressure before it closes again, returning the process unit to normal operation. PRVs are chosen as the relief device of choice since there are no corrosive process fluids that could damage the PRV. The PRVs are to be sized to relieve, or open, at 110% of the equipment's design pressure, or the maximum allowable working pressure, for the appropriate capacity of the respective units. PRVs should be pop-tested between shut downs to make sure that PRVs will properly relieve when needed. The PRVs added are used to protect the equipment from pressure accumulation and blocked discharge. For example, a PRV is included on discharge of positive displacement pumps since they will continue to pump a fixed volume of fluid, so if the downstream valve was closed (blocked), pipe failure will occur. Consequently, a PRV was added to route fluid back to pump suction to ensure continuous pumping without failure. In addition, PRVs are included on steam side of utility exchangers to prevent build-up of pressure in case of blocked downstream valve, while PRVs are included on the process side of said exchangers to protect from thermal expansion of trapped fluid if outlet is blocked, which could rupture the tubes. All PRVs relieving hydrocarbons will be relieving to flare to combust the hydrocarbons instead of releasing directly to atmosphere, while PRVs relieving steam and water will relieve to atmosphere and grade, respectively.

5.1.5 Containment

The fifth layer of safety is containment, which diverts contents of from each process unit upon the activation of the aforementioned safety measures, such as SIS and relief. Liquid contents from vessels are routed to swing tanks V-204 and V-205 during process upsets for temporary storage until normal operation is restored. Discharge of PRVs are routed to a flare header for safe disposal of hydrocarbons through combustion to break down hydrocarbons rather than venting to atmosphere.

5.1.6 Emergency Response

Four steps are required to ensure proper disaster prevention emergency response [20]. Firstly, understanding the threat of each equipment and how to deal with disasters that could occur with that single piece of equipment. Secondly, a subsystem of personnel must be established that defines roles for both normal and emergency operation. Personnel are held accountable for their section and are required to provide appropriate training to other workers. Thirdly, a physical subsystem must be constructed which includes transportation systems, waste removal systems, communication systems and storage areas if any emergencies were to arise. Lastly, an external subsystem such as the local fire department and police enforcement should be established. All individuals should receive proper training on the hazards of all chemicals used in this process and the ways to deal with potential hazards if equipment has failed in the case of an emergency. Response time, procedures to combat DME and methanol fires, procedures on how to handle pressurized vessels and procedures on the potential dangers of methanol and DME vapors must be established by the external individuals.

5.2 HAZOP

Table 16: A detailed HAZOP study on the T-201 overhead to E-205 node. Numbers in columns **ii-iv** refer to the specific causes in column **i**.

Guide Word	Deviation	i. Cause	ii. Consequence	iii. Action	iv. Safeguard
No	No overhead flow	1. No tower feed flow 2. Feed valve failure/closure	1,2. Tower not operational 2. Feed pipe rupture	1. Start process 2. Open manual bypass around feed CV	2. Feed valve bypass loop
Less	Less overhead flow	1. Feed pipe blockage 2. Less feed flow 3. Pipe blockage in overhead line 4. Low vapour feed (feed temperature below bubble point)	1. Feed pipe rupture 2-4. Low flow to exchanger can result in fouling of exchanger due to lower CW rate 4. Changes in DME quality	1. Open feed manual bypass 3. Open overhead manual bypass 4. Increase boil-up rate	3. Bypass lines 4. Control of column temperature and pressure
More	More overhead flow	1. Increased vapourization of bottoms liquid (more vapour flow upwards) 2. Superheated feed	1,2. Changes in DME quality 1,2. Increase in column pressure	1,2. increase reflux	1-2. Tower pressure control to adjust overhead vapour/liquid distribution

Low	Low overhead pressure	1. Failure of T-201 pressure control system (pressure controller opens CWS valve) 2. Reflux flow meter indicates false low flow (flow controller opens valve more than needed) 3. Reflux control valve is stuck fully open	1. Too much cooling duty 2,3. Composition of DME is affected	1. Open manual bypass on process side of E-205 to have higher reflux temperature 2. Isolate reflux control valve and control flow by by-pass valve.	1. Low Pressure Alarm
High	High overhead pressure	1. Failure of T-201 pressure control system (pressure controller closes CWS valve) 2. Reflux flow meter indicates false high flow (flow controller closes valve) 3. Reflux pump stops working 4. Pipe blockage (between tower overhead and E-205) 5. Closure of block valve (upstream of E-205)	1,2. Loss of cooling capacity 4,5. Accumulator/drum may run dry with no feed and pump cavitates 1-5. Reduction of reflux 1-5. Increase of temperature profile 4,5. Build-up of fluid may damage pipes 1-3. Composition of DME is affected with higher heavy-key in overhead	4. Regular maintenance of overhead line 1. Open manual bypass around pressure control valve	4,5. Pressure alarm 3. Spare reflux pump 4,5. Relief from pressure accumulation
Low	Low overhead temperature	1. Utility Stream control valve fully open (malfunctioned) 2. Low operating temperature in the tower 3. Fouling in E-205 in heat exchanger 4. Decrease vapour flow	1,3. Cooler reflux feed into column 1,3. Flashing occurring over the valve due to pressure drop 1-3, Improper separation of DME and methanol 2. Improper feed temperature (lower) 2,4. Decrease in reflux flow	1,3. Open manual bypass around pressure control valve and throttle to the manual valve to position of control valve before the malfunction	1,3. Temperature measurement on overhead outlet 1,3. Manual bypass around control valve 2. Temperature measurement on tower
High	High overhead temperature	1. Utility Stream control valve fully closed (malfunctioned) 2. Higher operating temperature in the tower 3. Fouling in E-205 in heat exchanger 4. E-205 bypass valve on process side is open (less is being cooled, thus hotter reflux) 5. Reboiler rate too high 6. Increase vapour flow	1-4. Improper operation of the total condenser 4. Hotter reflux feed into column 1-4. Partial liquid/vapour fluid back into tower 1-4. Cavitation in the condenser pump (vapour in reflux feed) 1-4. Seat in valves (control and manual) damaged causing leaking 5. Heavy vapours pass up the tower increasing the temperature 2,6. Increased reflux flow	1,3. Open manual bypass around pressure control valve and throttle to the manual valve to position of control valve before the malfunction	1,3. Temperature measurement on overhead outlet 1,3. Manual bypass around control valve 2. Temperature measurement on tower

6 Conclusions

Periwinkle Consulting has conducted a DME synthesis plant preliminary design, safety, operability, and economic feasibility study of a process capable of producing 50,000 MT of DME at 99.5% wt. The report herein, has reemphasized the use of DME as an alternative fuel replacement, given its efficient and cleaner combustion. The proposed process herein produces DME via catalytic methanol dehydration, and effectively meets Ethereal Energies process specifications. An economic analysis concluded that the yearly total revenues and total operating cost of the plant were $43.6M & $17.5M respectively. Furthermore, assuming a plant life of 25 years a positive net present value of $127.9M was obtained, which means that the overall project is both profitable and economically feasible. The sensitivity analysis concluded that MARR impacted NPV the most.

An operating window for the reactor unit was created to provide further recommendations to plant operators on the preferable and optimal operating conditions. A flexibility and reliability study was conducted to ensure that plan runs at optimal conditions when deviations from steady state operations are presented, and lifetime of equipment is extended. A start-up and shut-down procedure was also proposed to ensure the plant operates efficiently and damage of equipment is reduced. Lastly, respective six layers of safety were implemented throughout the plant design, to ensure that the plant operates in a safely manner. A HAZOP study was conducted on the node, T-201 and E-205 to identify in detail any risks involved in the process and include safeguards that would reduce or prevent any severe consequences.

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8 Contribution of Each Group Member

Group Member Name - Student Number	**List of Significant Section Contribution**
Gabriel Sanocki - 001207344	1 – Introduction 2 - Process Overview 2.1 - Purpose of Each Unit 2.1.2 – Reactor 3 – Economics 3.1 - Typical Operating Costs 3.2 - Detailed Estimates of Capital 3.3 - Profitability Analysis 3.4 - Detailed Sensitivity and Risk Analyses 8 – Contribution of Each Group Member Formatting & Review
Moustafa Kasem - 001217322	5 – Safety 5.1 – Six Levels of Safety 5.1.1 – Basic Process Control System 5.1.2 – Alarms 5.1.3 – Safety Interlock System 5.1.4 – Pressure Relief 5.1.5 – Containment 5.2 – HAZOP Cover Letter
Jay Modi - 001217136	2.2 – Alternative Technologies 2.3 – Detailed P&ID 5.2 – HAZOP 7- References Formatting and Review
Lucas Cordy - 001217901	2.1.3 - Pumps 2.1.4 - Heat Exchangers 2.1.5 - Holding Tank Vessels 4 - Operability 4.1 - Operating Window for Reactor Unit 4.2 - Flexibility and Controllability Analyses 4.3 - Reliability Improvements 4.4 - Start-Up and Shut-Down Procedure 5.1.6 – Emergency Response
Jesus Flores - 001221010	2 - Process Overview 2.1 - Purpose of Each Unit 2.1.1 - Distillation Columns 2.4 – Process Differences & Redundancies 3 - Economics 3.1 - Typical Operating Costs 3.2 - Detailed Estimates of Capital 3.3 - Profitability Analysis 3.4 - Detailed Sensitivity and Risk Analyses Aspen Programming 6 - Conclusion