Spring 5-1-2021

AVEVA Competition 2021: Dimethyl Ether Production Process

Nicholas Gaouette

AVEVA COMPETITION 2021: DIMETHYL ETHER PRODUCTION PROCESS

by

Nicholas Gaouette

A thesis submitted to the faculty of The University of Mississippi in partial fulfillment of the requirements of the Sally McDonnell Barksdale Honors College.

Oxford, MS

May 2021

Approved By

____

Advisor: Professor Adam Smith

____

Reader: Professor Mike Gill

____

Reader: Professor David Carroll

DEDICATION

This thesis is dedicated to the family, friends, colleagues, and instructors that were so invaluable throughout this process. Thank you for investing all that you have into me, my education, and this project.

ACKNOWLEDGEMENTS

I would like to thank everyone at AVEVA that was so helpful in the completion of this project; your support and knowledge were indispensable as we sought to master AVEVA Process Simulator and complete our objectives. Thank you to those that authored the project and took the time to oversee it; it was a great experience.

My professors and thesis committee were also quite helpful in this process. Your constant support, encouragement, and selfless volunteering of your time helped us tremendously in completing this project. Thank you to Adam Smith, Mike Gill, and David Carroll, as well as the rest of the Chemical Engineering faculty at Ole Miss.

My co-authors Erin Bridgman, John Marquez, and Ryan Schneider were an absolute delight to work with for the duration of this project. They proved to be excellent teammates and better friends. I am profoundly grateful for all the sacrifices you all made in working toward this common goal; it was a pleasure. Your contributions to this project cannot be overstated.

Lastly, I would also like to thank God for showing me grace in all the ways that He has, and for providing me with everything I might have needed to finish this project. To Christ be the Glory, now and always.

ABSTRACT

The following study was conducted to assess the economic and process feasibility of a dimethyl ether (DME) production process. A portion of the methanol produced by an existing facility would be used to produce DME through a dehydration reaction. DME production is being considered as a means to compensate for lost methanol revenue, as methanol prices have recently decreased and an excess supply is present on the open market.

The first milestone in the study was the construction of a working process model within AVEVA Process Simulation. This steady-state process model would provide information essential to further design and optimization efforts, and serve as a starting point from which process optimizations could be considered.

An additional process model was created, using the “base-case” as a template, in which a distillation column was economically optimized. In this optimized column model, the size and configuration of the distillation column were adjusted to minimize the Equivalent Annual Operating Cost (EAOC) of the unit.

Because the DME process was proposed as a temporary solution, it was necessary to cooperate with a Toller in renting process equipment. The Toller had a limited inventory, so availability was a consideration in the choosing of DME process equipment.

Upon delivering the optimized column model to management, the team was provided with a slightly adjusted process model. This model would serve as the basis for the creation of a

number of new models, using various equipment sizes from the Toller's inventory. An optimal set of equipment was subsequently chosen using the information provided by these models.

The team performed an economics analysis of the project using process information from the recommended optimized process model and cost estimation methods for chemical processes. While a number of estimates were made in this analysis, the economic metrics ultimately indicated that this project would be economically viable, with a net present value of $11.5 million and a conventional payback period of 2.6 years.

Process feasibility was also assessed in this study. It was concluded that the proposed process is feasible, as the three areas of concern identified were deemed acceptable in the presence of sufficient safety measures and process controls. This determination was made using a process conditions matrix and generalized experience within the chemical industry.

It was recommended that the company move forward with the dimethyl ether plant immediately. Additional recommendations to improve process economics were also made, which are discussed below.

PREFACE

The scenario explored in this thesis is that which was provided in the 2021 AVEVA Academic Competition, a competition in which students utilize AVEVA's Process Simulation Program to solve a chemical engineering design problem. The scenario provided in the problem statement of the competition is explored below, along with actions taken to by my team to reach a solution.

I completed this work with great help from three of my fellow Chemical Engineering students: Erin Bridgman, John Marquez, and Ryan Schneider. We worked under the supervision of the three professors that composed my thesis committee: Adam Smith, Mike Gill, and David Carroll.

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

LIST OF ABBREVIATIONS

DME	Dimethyl Ether
EAOC	Equivalent Annual Operating Cost

BACKGROUND

A methanol production facility produces 88,000 tonnes of methanol per year with two major clients. The first client receives 65,000 tonnes, while the second client receives 23,000 tonnes. Due to a recent economic downturn, the second client recently decided not to renew their methanol contract. This development left the plant with excess production of methanol, and no guaranteed buyer. This situation was exacerbated by a low market price of methanol, and an excess supply of methanol on the open market.

Management asked the team to evaluate three potential plans of action to offset the loss of the second client. The first option was to scale down methanol production, resulting in a yearly loss of $19,090,000 in methanol sales. This option may incur additional costs should the process equipment need modification or replacement to produce methanol at the reduced rate. The lower rate of production would also increase production cost per unit of methanol, leading to slimmer profit margins. The second option was to sell the methanol on the open market, which would yield a lower unit price compared to contract pricing. The third option was to convert the methanol to dimethyl ether (DME). DME has a lower supply and higher market demand than methanol, suggesting that it could serve as an attractive alternative to the methanol product. With this in mind, management tasked a group of engineers to explore the potential of a temporary DME facility that uses equipment rented from a Toller.

PROCESS OVERVIEW

Dimethyl ether is produced by the catalytic dehydration of methanol. The reaction is an equilibrium reaction, and no side reactions were considered. The Process Concept Diagram for the DME process can be seen below in Figure 1.

The DME process consists of three major process "blocks." The major "blocks" of the process are the methanol feed preparation, methanol dehydration reaction, and DME separation. Figure 2 below shows the Block Flow Diagram for the process.

The 23,000 tonnes of methanol from the existing plant are fed to the DME production facility. The methanol, received from the existing facility as a liquid, and must be vaporized before being sent to the reactor. The gas-phase reaction occurs in a catalytic packed bed reactor with a single-pass conversion of 81%. The remaining methanol, along with water and DME, are then sent to a distillation column, where the DME product is separated from water and methanol. The methanol and water mixture is sent to the existing methanol facility, where the methanol and water are separated. The separated methanol is then sent back into the DME process. While this recycle to the existing methanol facility was outside the scope of this analysis, consideration of this recycle would show an increased overall conversion and yield and more favorable process economics.

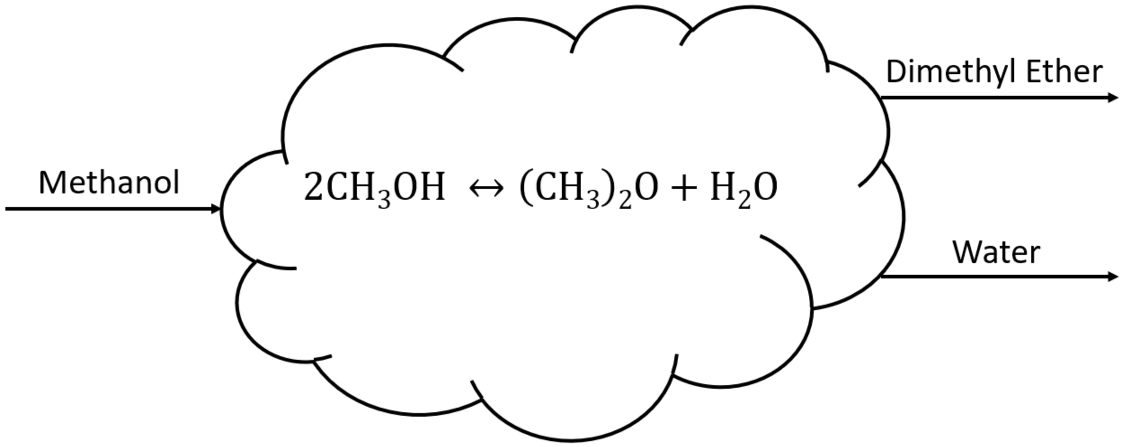

Figure 1: Process Concept Diagram

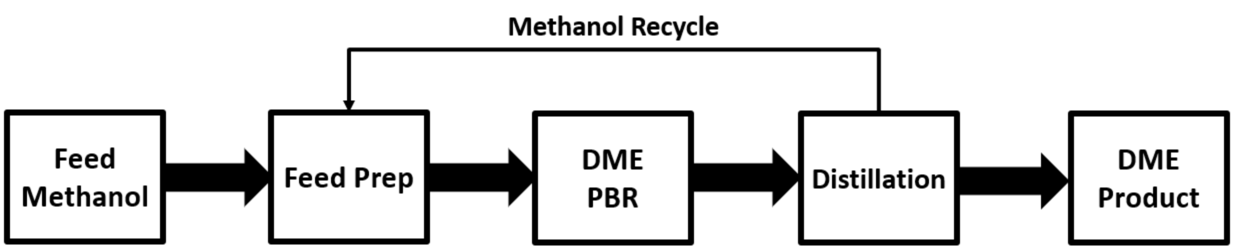

Figure 2: Block Flow Diagram

BASE CASE DESCRIPTION

As a preliminary investigation in the optimization of the DME process, the Process Flow Diagram, provided in Appendix A, was simulated in AVEVA's Process Simulator. The feed mixture is composed of 99.82 mole % methanol and 0.18 mole % water. Literature shows that most known equations of states for methanol-water mixtures cannot be applied with a high degree of accuracy. Therefore, to simulate the process, the UNIFAC equation of state was utilized to estimate interaction parameters, based on experimental data.

This "base-case" simulation modeled the synthesis of DME to achieve an 81% single-pass conversion of methanol, with a 99.5 weight % purity of DME product. This simulation utilized a catalytic packed bed reactor and a trayed distillation column.

Several constraints were inherent to the "base-case" design, specifically for the packed bed reactor and distillation column. The reactor is constrained by conversion, size, and pressure drop specifications. The length to diameter ratio for the reactor must be in the range of 3:1 to 8:1, and pressure drop across the reactor should not exceed 50 kPa. To achieve the necessary conversion (81%) and limit pressure drop to less than 50 kPa, the height and diameter of the reactor were found to be 7.11 meters and 1.35 meters, respectively.

Additionally, the distillation column must meet DME product purity specifications. The distillation column was constrained by a maximum flooding limit of 80%, with a recommended minimum flooding limit of 30%. The column was simulated in AVEVA to meet these provided requirements, which resulted in a distillation column 8.85 meters in height, 1 meter in diameter, with 15 valve trays.

BASE CASE ASSESSMENT

The base-case simulation of the DME production process achieved the required single-pass conversion in the catalytic reactor, as well as DME product specifications. Thus, the production of DME is a feasible alternative to scaling-down the existing methanol facility, or holding out for new long-term methanol contracts.

Without consideration of construction and auxiliary costs, the DME plant has the potential to be lucrative. Shown below in Table 1, the annual profit of the preliminary DME process design is $2.1 million. The DME is sold at $0.83/kg, and the process generates approximately $11.2 million in annual revenue.

The design team highly recommends proceeding with the project. The preliminary design shows the process to be economically viable, before implementing any optimization strategies. Although the proposed process consists of rented equipment, the "base-case" design uses cost estimation methods from Turton (1) to estimate the purchase cost of process equipment, as rental prices were not available to the design team at this stage in the project.

Table 1: "Base-Case" Annual Economics

Total Purchase Cost of Equipment:	$	(787,186)
Total Utility Cost of Base Case:	$	(649,058)
Cost of Operating Labor:	$	(715,000)
Cost of Raw Materials/Feed:	$	(6,899,904)
Product Revenue	$	11,166,126
Profit:	$	2,114,978

DESIGN OPTIMIZATION

The optimization of the DME process was conducted in two stages. First, the minimization of operating cost, specifically through minimizing the equivalent annual operating cost (EAOC) of the distillation tower. Second, the optimization of DME production through increasing reactor single-pass conversion and DME separation in the column, utilizing available Toller equipment.

PART I: MINIMIZATION OF EAOC

Following the simulation of the “base-case” process, the first optimization was performed on the distillation column.

Equivalent annual operating cost (EAOC) is a metric used to perform economic comparison and evaluations. EAOC is a function of operating costs, purchase costs, present worth factor, material factor, and pressure factor. As recommended by management, a present worth factor of seven was used in this project. The material factor is one, which assumes the material of construction for process equipment is carbon steel. The pressure factor is a function of defined operating pressures for equipment, and is different for each piece of process equipment. Thus, the minimization of EAOC is performed by minimizing total utility cost and equipment operating costs.

The purchase cost of the distillation column is the summation of the costs associated with the tower, trays, reboiler and condenser. Therefore, to minimize the EAOC, the volume of the column and number of trays used must be minimized. Within AVEVA’s optimization set manager, an objective function was defined to minimize EAOC through changing variables of tower height and diameter, as well as total utility cost. The optimization function converged, resulting in a tower that met product specifications; the tower was 7 meters in height and 1 meter in diameter.

After determining the smallest volume required for the distillation column and minimizing utility cost, the minimum number of trays needed to meet product specifications was found. It was discovered that only seven valve trays were required to achieve the desired separation within the column, reducing purchase cost of trays by over 50%.

As seen in Table 2, the strategies of reducing tower volume, trays, and utility cost employed in the first stage of optimization resulted in an EAOC of $100,930. The optimized dimethyl ether column will produce a 99.82 mole % product using seven valve trays, while remaining within the specifications provided in the AVEVA Problem Statement. The optimized column reduced the EAOC by 45.1% from the “base-case.”

Table 2: Column Optimization Parameters

Table 1: Column Optimization Parameters				
	Number of Trays	**Height (m)**	**Diameter (m)**	**EAOC ($/yr)**
Base Case	15	8.9	1.0	$ (183,977)
Optimized Tower	7	7.0	1.0	$ (100,930)

PART II: OPTIMIZATION OF DME PRODUCTION

In order to maximize profit from the process, it is important to produce the largest amount of product possible while mitigating the costs of production. The product must still meet all specifications, such as purity and phase. In the optimization process, the pieces of equipment that have the largest effects on production capacity are the reactor and distillation column. In the reactor, a greater conversion results in more product produced per unit of reactant fed. Once a high conversion is achieved, it is important to be able to separate the largest practical amount of the desired product from other process components. A highly efficient separation process will accomplish this at a minimum cost.

Following optimization of the reactor and distillation column, the heat exchangers within the process were examined. The rental cost and operating cost of the exchangers contribute significantly to project economics; these costs were minimized by using appropriately sized exchangers within the process. The available reactors, columns, and heat exchangers in the Toller's inventory are shown below in Tables 3 and 4.

Table 3: Available Reactors and Columns from Toller

Equipment	Length (m)	Diameter (m)	Max Op Temp (°C)	Max Op Pressure (bar)	Catalyst Volume* (m^3)	Rental Cost ($/mo)**
Reactor A	5	1	400	12	3.93	10.0 k
Reactor B	4	0.8	400	12	2.00	6.31k
Reactor C	7	1.4	400	12	10.74	20.3 k
	Length *** (m)	Diameter**** (m)	Max Op Temp (°C)	Max Op Pressure (bar)	No. of Valve Trays*****	Rental Cost ($/mo.)
Column A	9	0.5	300	11	20	5.8 k
Column B	10	0.6	300	7	24	7.9 k
Column C	10	0.8	300	15	24	11.9 k

Table 4: Available Heat Exchangers from Toller

	Area (m^2)	Max - Tube P(bar)/T(°C)	Max - Shell P(bar)/T(°C)	Configuration Shell-pass	Tube-pass	Rental Cost ($/mo.)
Exchanger A	125	15/150	15/150	1	2	5.9 k
Exchanger B	90	15/300	50/300	1	2	4.5 k
Exchanger C	60	15/150	15/400	1	2	3.7 k
Exchanger D	40	20/300	15/180	1	1	3.4 k
Exchanger E	180	20/300	50/300	1	2	6.7 k
Exchanger F	100	15/150	15/150	2	4	6.1 k
Exchanger G	20	20/300	15/180	1	1	1.1 k
Exchanger H	150	50/300	15/300	1	1	6.1 k

The reactor was the first piece of equipment investigated. Three possible reactors were provided from a Toller with sizing, as well as maximum value for temperature, pressure, and catalyst volume. The temperature and pressure rating of the reactors were sufficient, and each reactor could be completely filled with catalyst if necessary. A void fraction of 0.25 was used for all reactor configurations, to simplify comparison. The first reactor simulated, reactor A, obtained an 84% single-pass conversion and a sufficiently low pressure drop. Reactor C produced an 84% single pass conversion, again with an acceptable pressure drop. At the specified reactor temperature and pressure, reactor B could not meet process requirements. Even when fully packed with catalyst, reactor B could not provide adequate single-pass conversion. For this reason, it was determined that reactor B was not a suitable piece of equipment at the specified process conditions. The savings from renting reactor B, the smallest piece of equipment, would not justify the lower conversion and higher pressure drop. The largest reactor, reactor C, had the highest rental cost, but could not achieve a higher single-pass conversion than the smaller, less expensive reactor A. Thus, reactor A was determined to be the optimal reactor. The optimal catalyst volume fraction was then found to be 0.22, using an iterative solution to deliver improved economics.

An optimal distillation column was then chosen. The DME production rate using column A was found to be significantly higher than that of the "base-case," as can be seen in Table 5. The implementation of column C resulted in a DME production rate similar to that of column A, but column C was larger and had a higher rental cost. Column B was restricted by a relatively low maximum allowable pressure. To operate below this value, it was necessary to decrease the saturation temperature of the overhead significantly. This change resulted in a temperature differential between the process stream and the utility stream was insufficient; the overhead could not be condensed with cooling water. To condense the overhead, a refrigerated water utility was necessary. Refrigerated water is more expensive than cooling water, and increased the cost of operating the tower more than 300%. This drastic increase in operating costs proved column B to be economically undesirable. Column A was thus chosen as the optimal tower.

Table 5: Design Progression

	Single-Pass Conversion	DME Product (kmol/hr)
Base-Case Simulation	0.81	33.9
Optimized Column Simulation	0.81	33.9
Optimized Process Simulation (Using Toller Equipment)	0.84	41.7

Once the optimal reactor and distillation column were chosen, the heat exchangers were considered. The size of each heat exchanger is directly related to rental cost; larger exchangers had higher rental rates. The smallest heat exchangers that met process needs were utilized, by finding those most similar to the sizes of heat exchangers in the "base-case." Once each heat exchanger had been chosen, the flowrates of relevant utility streams were manipulated to reduce utility costs while maintaining process specifications. The heat exchanger configuration for the optimized case can be seen below, in Table 6.

Table 6: Optimized Equipment Configuration

Optimized Equipment Set		
Equipment on PFD	Toller Equipment	Yearly Rental Cost ($)
R-101	Reactor A	$ 120,000.00
T-101	Column A	$ 69,600.00
E-101	Exchanger A	$ 70,800.00
E-102	Exchanger B	$ 54,000.00
E-103	Exchanger C	$ 44,400.00
E-104	Exchanger H	$ 73,200.00
E-105	Exchanger G	$ 13,200.00
	Yearly Cost	$ (445,200.00)
	Overall Yearly Cost	$ (691,585.56)

RECOMMENDED OPTIMIZED DESIGN

The recommended optimized design for the DME process utilizes the equipment configuration shown in Table 6. A screenshot of the AVEVA simulation and the associated stream table for the optimized design are provided in Appendix B and C, respectively. The utility usage and other relevant values for the chosen heat exchangers can be seen below, in Table 7.

The economic viability of the "base-case" design, optimized distillation design, and recommended optimized design were assessed using a net present value analysis for each instance. Each income statement considered a 12-year plant life, a 3% inflation rate, and a 12% minimum acceptable rate of return. Additionally, construction costs were estimated to be $1 million, with plant construction beginning in January 2022 and lasting for two years. The income statement for the recommended optimized design is provided in Appendix D. As reflected below in Table 8, the recommended optimized design generates $13.7 million annually from the sale of the DME, with an annual equivalent of $1.9 million and net present value of $11.5 million. The proposed optimized design is projected to result in a conventional capital investment payback period of approximately 6 months after plant start-up. Although additional changes to the process may prove necessary as the project progresses, the recommended optimized design case appears promising.

Table 7: Heat Exchanger Utility Information (Optimized Design)

		E101A	E101B	E102	E103A	E103B	E104	E105
Utility Type		LPS	LPS	HPS	CW	CW	CW	MPS
Process Side		Shell	Shell	Tube	Shell	Shell	Shell	Tube
Duty	kW	388	850	188	420	582	565	286
W_{shell}	kg/h	3150	3150	402	3150	3150	5218	508
W_{tube}	kg/h	657	1439	3150	6240	8322	41276	618

Table 8: Economic Progression of DME Process

Table 5: Economic Progression				
Economic Summary:	**Using a 12% MARR**			
	Net Present Value ($)	**Annual Equivalent ($)**	**Conventional Payback Period (yr)**	**Revenue ($/yr)**
Base-Case Simulation	$ (3,886,000)	$ (627,000)	11.7	$ 11,166,000
Optimized Column Simulation	$ (3,641,000)	$ (588,000)	11.6	$ 11,166,000
Optimized Process Simulation (Using Toller Equipment)	$ 11,474,000	$ 1,852,000	2.6	$ 13,731,000

APPROVAL CONSIDERATIONS

The construction of the DME production facility is spurred by the poor prospects of finding new long-term methanol contracts in the near future. The company is actively searching for new methanol contracts; if one is obtained, then the methanol available to be used in DME production may decrease dramatically. This would require a scale-down of the DME process or the purchase of methanol from an outside source. Should the process be scaled down, process equipment would have to be resized, again through cooperation with the Toller. Should methanol be purchased from an outside source, factors such as pricing and availability would warrant additional consideration.

At this stage in the project, the DME is to be sold on the open market, without any production contracts. With no DME contracts, some instability in the selling price of DME is to be expected. Without any assurances or guarantees for the future of the DME market, the economics of the project could vary widely in shifting market conditions.

The influence of catalyst cost and replacement were not considered, as they laid outside the scope of this analysis. The catalytic packed-bed reactor will operate 360 days per year, and will require catalyst replacement. Replacement of the catalyst at regular intervals could prove costly, and may negatively impact project economics. As the project progresses, the catalyst would need to be examined more closely. Specifically, factors such as catalyst pricing and availability should be considered.

PROCESS SAFETY

Due to the nature of the process at hand, there are a few notable process safety concerns that require consideration. As shown in the Process Conditions Matrix in Table 9, the reactor temperature and pressure and the pressure in the tower were noted as potential areas of concern. The reactor operates at a high temperature and pressure of 375 ℃ and 10.75 bar. These conditions were necessary to ensure the reactor feed remained in the vapor phase. The reactor conditions also favored high reaction rates within the unit. The high pressure in the distillation tower (10.06 bar) is necessary in maintaining vapor-liquid equilibrium throughout the tower. Safe and controlled operation of the process would require adequate pressure relief and controls for temperature and pressure.

Flammability concerns are present, as both methanol and DME are highly flammable. In order to protect against potential fires and explosions, it is essential that there is proper ventilation throughout the facility to remove any vapor that may be present. In addition, ignition sources should be limited when at all possible.

Table 9: Process Conditions Matrix for DME Process

Process Conditions Matrix for the Dimethyl Ether PFD										
	Reactors and Separators (Table 6.1-6.3)					Other Equipment Table 6.4				
Equipment	High Temp	Low Temp	High Pressure	Low Pressure	Non. Stoich Feed	Comp.	Exch.	Htr.	Valve.	Mix
R-101	X		X							
T-101			X							
E-101										
E-102										
E-103										
E-104										
E-105										
P-101 A/B										
P-102 A/B										
PCV on Tower Feed										
PCV on Reflux										

ENVIRONMENTAL CONCERNS

Methanol and DME are both toxic chemicals, and warrant environmental concerns. Therefore, wastewater treatment is imperative to inherently safer design. Due to the toxicity of both components, the wastewater must be treated thoroughly to prevent any potential contamination of groundwater. Additionally, it is important to have emergency procedures in place, in the event of a loss of containment.

RECOMMENDATIONS AND CONCLUSIONS

As stated earlier in our initial recommendation, the design team recommends that management proceed with the proposed DME process. While the "base-case" process had sufficiently attractive economic metrics, the economics of the process improved significantly as optimizations were pursued. The favorable economics of the optimized process suggest that this proposal would be an excellent way to make up for lost revenue, while maintaining the possibility of procuring future methanol contracts. With the approval of management, the design team could proceed with the project by beginning to coordinate contractors and suppliers for upcoming plant construction. The design team could then more accurately gauge the time and investment needed to fully implement the project.

The design team proposed two additional recommendations that could improve the economics of the process. The first of these recommendations is the outright purchase of process equipment, rather than a yearly rental of equipment. While purchasing the equipment might impact the adaptability of the process to changing economic conditions, it would present an improvement in project economics. With a sufficiently long project life, the outright purchase of equipment would improve the net present value of the project at the end of its lifespan, by removing the annual equipment rental costs. Should management elect not to purchase the equipment, a long-term rental contract with the Toller should be investigated.

The second of these recommendations is that long term DME production contracts be pursued, rather than selling on the open market. As is the case with methanol, contract prices for

DME may very well exceed open market prices. Production contracts also provide stability in the selling price of the product.

LIST OF REFERENCES

1. Turton, Richard. *Analysis, Synthesis and Design of Chemical Processes*. Prentice Hall, 2018.
2. "Once-Through Dimethyl Ether Process." AVEVA Competition Problem Statement

Appendix A

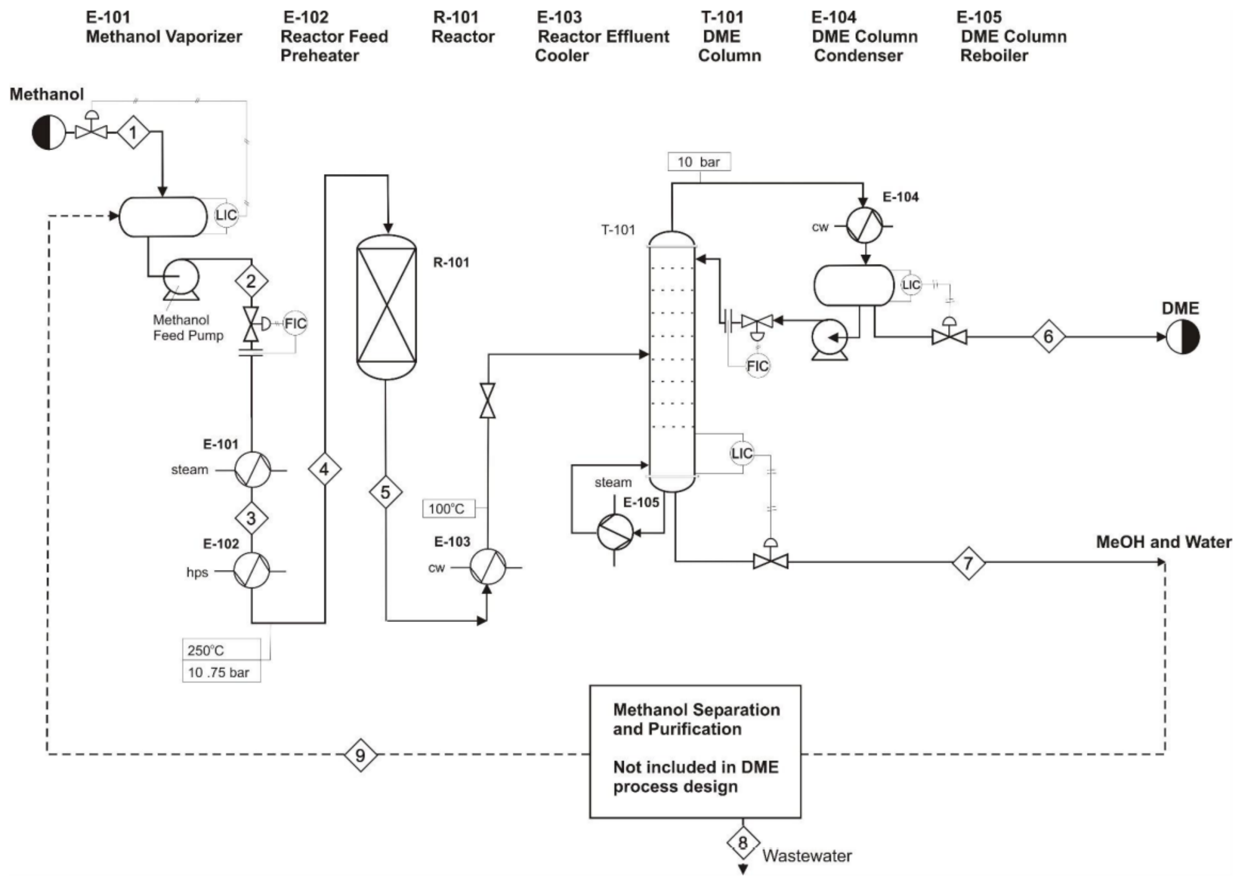

Figure 3: Preliminary Process Flow Diagram

Appendix B

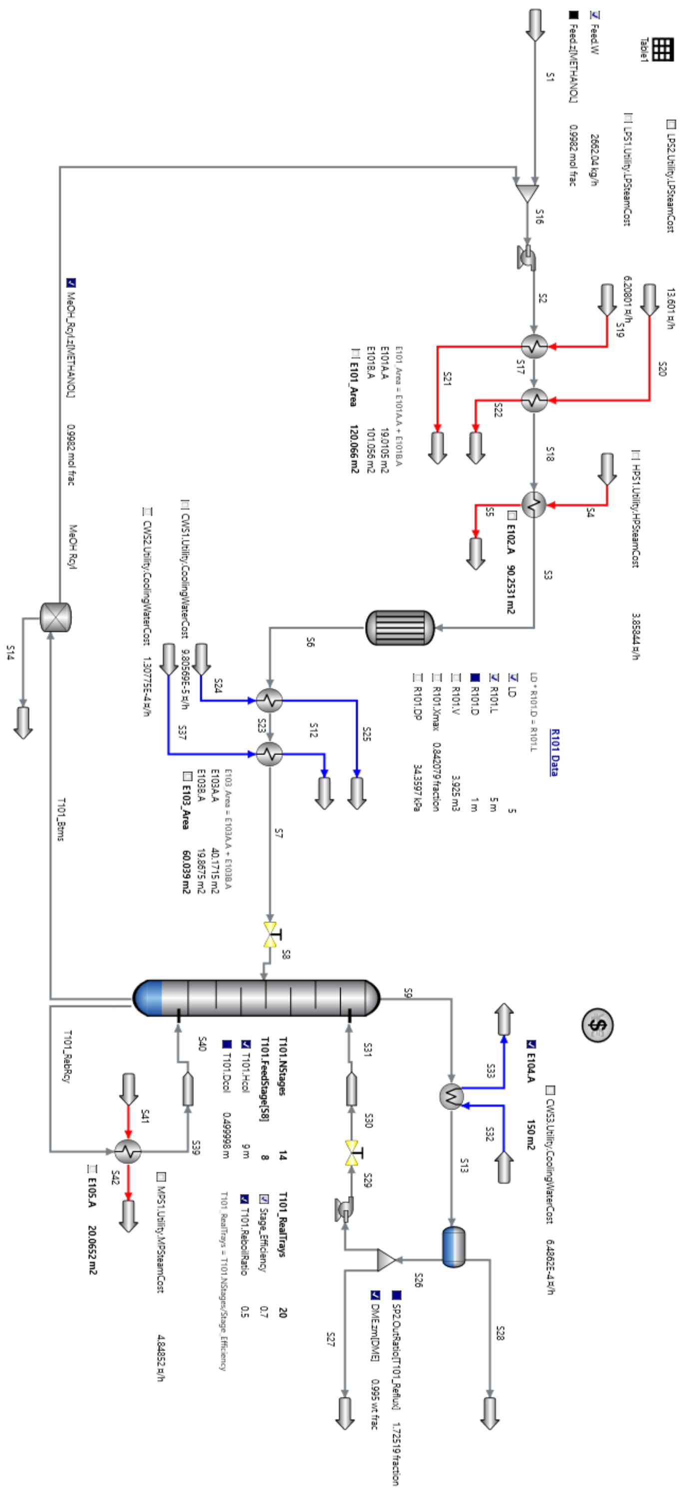

Figure 4: Screenshot of Optimized Solution

Appendix C: Optimized Design Stream Table

Table 10: Optimized Design Stream Table

Name	Units(SI)	MeOH_Rc	S1	S2	S3	S4	S5	S6	S7	S8	S9	S12	S13	S14	S16	S17	S18	S19
W	kg/h	488.24	2662	3150.3	3150.3	401.5	401.5	3150.3	3150.3	3150.3	5217.9	8322.1	5217.9	747.36	3150.3	3150.3	3150.3	656.93
F	kmol/h	15.249	83.145	98.394	98.394	-	-	98.394	98.394	98.394	113.53	-	113.53	41.485	98.394	98.394	98.394	-
T	C	30.262	30	30.241	250	256.28	256.14	375.27	100	100.37	46.151	90.204	43.806	65.394	30.041	139.46	139.32	150
P	kPa	101.3	101.3	1115	1075	4414.8	4404.8	1040.6	1030.6	1050	1000	470	990	1006.6	101.3	1110	1105	476.1
z[H2O]	mol frac	0.0018	0.0018	0.0018	0.0018	-	-	0.4221	0.4221	0.4221	0.0004	-	0.0004	1	0.0018	0.0018	0.0018	-
z[METHANOL]	mol frac	0.9982	0.9982	0.9982	0.9982	-	-	0.1576	0.1576	0.1576	0.0069	-	0.0069	8E-17	0.9982	0.9982	0.9982	-
z[DME]	mol frac	0	0	0	0	-	-	0.4203	0.4203	0.4203	0.9926	-	0.9926	7E-11	0	0	0	-

Name	Units(SI)	S20	S21	S22	S23	S24	S25	S26	S27	S28	S29	S30	S31	S32	S33	S37	S39	S40
W	kg/h	1439.3	656.93	1439.3	3150.3	6240.1	6240.1	5217.9	1914.7	0	3303.2	3303.2	3303.2	41276	41276	8322.1	617.8	617.8
F	kmol/h	-	-	-	98.394	-	-	113.53	41.66	0	71.872	71.872	71.872	-	-	-	28.367	28.367
T	C	149.22	147.59	146.77	147.9	30	88	43.806	43.806	43.806	44.475	44.47	44.47	30	41.798	30	168.17	168.17
P	kPa	466.23	446.1	436.23	1035.6	500	470	990	990	990	1690	1640	1640	300	270	500	976.56	976.56
z[H2O]	mol frac	-	-	-	0.4221	-	-	0.0004	0.0004	0.0001	0.0004	0.0004	0.0004	-	-	-	0.7317	0.7317
z[METHANOL]	mol frac	-	-	-	0.1576	-	-	0.0069	0.0069	0.0009	0.0069	0.0069	0.0069	-	-	-	0.2683	0.2683
z[DME]	mol frac	-	-	-	0.4203	-	-	0.9926	0.9926	0.999	0.9926	0.9926	0.9926	-	-	-	5E-11	5E-11

Name	Units(SI)	S41	T101_Btms	T101_RebRcy	T101_Reflux
W	kg/h	508.23	1235.6	617.8	3303.2
F	kmol/h	-	56.734	28.367	71.872
T	C	177.4	156.12	156.12	43.806
P	kPa	944.08	1006.6	1006.6	990
z[H2O]	mol frac	-	0.7317	0.7317	0.0004
z[METHANOL]	mol frac	-	0.2683	0.2683	0.0069
z[DME]	mol frac	-	5E-11	5E-11	0.9926

Appendix D

INCOME STATEMENT

		-3	-2	-1		1	2	3	4	5	6	7	8	9	10	11
End of Year		-2	-1	0	1	2	3	4	5	6	7	8	9	10	11	12
Income Statement																
Revenue		3%			$13,730,539	$14,142,455	$14,566,729	$15,003,731	$15,453,843	$15,917,458	$16,394,982	$16,886,831	$17,393,436	$17,915,239	$18,452,696	$19,006,277
Expenses	Inflation															
Materials (C_{RM})	0%				(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)	(6,899,904)
Catalyst (C_{Cat})	0%			-			-			-			-			
		0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
Labor (C_{OL})	3%				(715,000)	(736,450)	(758,544)	(781,300)	(804,739)	(828,881)	(853,747)	(879,359)	(905,740)	(932,912)	(960,899)	(989,726)
Utilities (C_{UT})	0%				(246,386)	(246,386)	(246,386)	(246,386)	(246,386)	(246,386)	(246,386)	(246,386)	(246,386)	(246,386)	(246,386)	(246,386)
Waste Treatment (C_{WT})	0%				-	-	-	-	-	-	-	-	-	-	-	-
Other (C_{OT})					(3,157,988)	(3,195,097)	(3,233,320)	(3,272,688)	(3,313,237)	(3,355,003)	(3,398,021)	(3,442,330)	(3,487,969)	(3,534,976)	(3,583,394)	(3,633,265)
Depreciation																
Land Depreciation					-	-	-	-	-	-	-	-	-	-	-	-
Land (BV)				$0	-	-	-	-	-	-	-	-	-	-	-	-
Building Depreciation					(24,573)	(25,641)	(25,641)	(25,641)	(25,641)	(25,641)	(25,641)	(25,641)	(25,641)	(25,641)	(25,641)	(24,573)
Bldg BV				$1,000,000	975,427	949,786	924,145	898,504	872,863	847,222	821,581	795,940	770,299	744,658	719,017	694,444
		MACRS Bldg			2.46%	2.56%	2.56%	2.56%	2.56%	2.56%	2.56%	2.56%	2.56%	2.56%	2.56%	2.46%
Equipment Depreciation					(220,218)	(377,407)	(269,532)	(192,479)	(137,617)	(137,463)	(137,617)	(68,732)	-	-	-	-
Equipment BV				$1,541,066	1,320,848	943,441	673,909	481,430	343,813	206,350	68,733					
		MACRS Equp			14.29%	24.49%	17.49%	12.49%	8.93%	8.92%	8.93%	4.46%				
Taxable Income / (Loss)			-	-	2,466,470	2,661,570	3,133,402	3,585,333	4,026,319	4,424,180	4,833,666	5,324,479	5,827,796	6,275,420	6,736,472	7,212,423
Income Taxes					(690,612)	(745,240)	(877,353)	(1,003,893)	(1,127,369)	(1,238,770)	(1,353,426)	(1,490,854)	(1,631,783)	(1,757,118)	(1,886,212)	(2,019,478)
Net Income / (Loss)				-	**1,775,859**	**1,916,331**	**2,256,049**	**2,581,440**	**2,898,949**	**3,185,410**	**3,480,239**	**3,833,625**	**4,196,013**	**4,518,302**	**4,850,260**	**5,192,945**

Figure 5: Income Statement for Optimized Design