Steady State Simulation of a Delayed Coker Unit

by

Norazah bt Abdul Rahman

Dissertation submitted in partial fulfilment of
the requirements for the
Bachelor of Engineering (Hons)
(Chemical Engineering)

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CERTIFICATION OF APPROVAL

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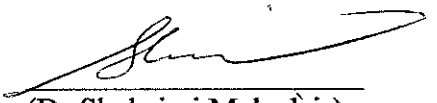
Approved by,



(Dr Shuhaimi Mahadzir)

UNIVERSITI TEKNOLOGI PETRONAS

TRONOH, PERAK

January 2009

CERTIFICATION OF ORIGINALITY

This is to certify that I am responsible for the work submitted in this project, that the original work is my own except as specified in the references and acknowledgements, and that the original work contained herein have not been undertaken or done by unspecified sources or persons.

NORAZAH BT ABDUL RAHMAN

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ABSTRACT

Delayed coking is a form of thermal cracking used for processing "bottom of barrel" residuum. Thermal cracking produces a wide range of hydrocarbons along with coke. Products of the coking process include sour fuel gas, sour liquefied petroleum gas (distillate), naphtha, light coker gas oil (HCGO), and heavy coker gas oil (FZGO). Steady state simulation of a delayed coker unit is the main aim of this project. It is also to study on hypothetical components of each component in the coker residue mixture. Besides that, this project is to explore the iCON simulator since iCON is a new simulator by using refinery oil. Features in iCON will be explored as the fractionator unit is being modeled. The delayed coker unit is a thermal reaction which involves large number of heat and very complex reaction. This project also studied the behavior of the hypothetical components with distillation process.

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To all individuals that has helped me in any way, but whose name is not mentioned here, thank you all.

CHAPTER 1
INTRODUCTION

1.1 Background of Study

Delayed coking is one of the main processes for heavy oil processing in petroleum refining industry. In recent years, an increasing has been observed in the number of crude oil which yields an appreciable amount of vacuum residue. Therefore, further processing of vacuum residue become important. Delayed coking is a thermal process in which the vacuum residue from crude distillation is heated in a furnace then confined in a reaction zone or coke drum under proper operating conditions of temperature and pressure until the unvaporized portion of the furnace effluent is converted to vapor and coke.

Delayed coking is an endothermic reaction and involved large hydrocarbon molecule, with the furnace supplying the necessary heat for the coking reactions. The reactions in the delayed coking are complex. The initial phase, the feed is partially vaporized and cracked as it passes trough the furnace. In the next step, cracking of the vapor occurs as it passes through the drum. In the final step, successive cracking and polymerization of the liquid confined in the drum takes place at high temperatures, until the liquid is converted into vapor and coke. The delayed coking is essentially a high temperature process involving extensive use of direct heat to up-grade product. This coking process, as a combined process of the severe thermal cracking and condensation reaction, needs to consume a large amount of high-grade energy. Thus it has been very difficult task in optimizing and simulating this delayed coker unit since many variables and constraints should be considered.

Fresh feed is preheated through a heat exchange system by entering the bottom of the coker fractionator. Coking take place in a coker drum and cracking process occur inside it. The untreated vapor leaves the coker drum enter the coker fractionator where it is separated in the desired fraction such as wet gas, distillate , light coker gas oil and heavy coke gas oil. Fractionating of this heavy residue employs multi-stage distillation towers

[1]. Active research is being carried out in developing a simulation steady-state behavior of distillation column. In this research, iCON simulation has been used in employs lumping of the Coker's component mixture. Due to the large hydrocarbon molecule involved, the simulation requires an efficient iterative calculation. Large problems are created because of the large hydrocarbon molecule in the mixture need to be separated. It is believed that in the petroleum industry, a system which consists of large hydrocarbon molecule is uncommon since it will involve numerous numbers of hypothetical components. Due to this there is possibility that the simulation may become unreasonable.

1.2 Problem Statement

Reaction in the delayed coker unit is very complex and complicated in which involve large hydrocarbon molecule. It is thermally cracking reaction that is highly endothermic which consume large amount of heat. In iCON a lots of coker's component does not establish such as heavy coker gas oil and light coker gas oil. These components are called as hypothetical components. In order to run a simulation it is important to ensure that all the hypothetical components are conventional. Hypothetical need to be established before builds the fractionator model.

In addition coker's fractionator involve high operating temperature and pressure. It is significant to make sure the operating temperature and pressure is corresponding to the process. The coker's components mixture consists of various numbers of boiling points in which some of the components have small boiling temperature differences and some of them have large boiling temperature differences. Since iCON is a new developed simulating program, many of their features need to be explored. The study on the iCON is done by fractionating this component mixture. The analysis could be done by converging the coker's fractionator column. Good initial estimate must be accomplished which resulting in less time consuming in doing the iteration.

1.3 Objective and Scope of Study

Objective of this project is to run a lumping model of the components present in mixture which involving the fractionators' of delayed coker unit system. It is also to study on hypothetical components of each component in the coker residue mixture. Besides that, this project is to explore the iCON simulator since iCON is a new simulator by using refinery oil. Features in iCON will be explored as the fractionator unit is being modeled. Under the present flowsheet conditions, the changes are adjusted with the aid of iCON simulation technique.

The method involves hydrocarbon coker mixture of heavy vacuum residue and being analyzed with other hydrocarbon component such as heavy residue from atmospheric bottom. By lumping the coker's fractionator column, study on behavior of distillation column can be carried out as well. The iteration is running continuously until the model is converged. While running the iteration, the effectiveness and behaviour of the iCON environment is studied. The operating condition, stages, and reflux ratio are manipulated in order to obtain model for the coker unit.

In iCON usually chromatohraphic information is not available and therefore characterization using hypothetical components is needed [2]. The physical properties of the pseudo components are gathered in order to calculated thermophysical properties of the oil mixture as well as necessary thermodynamic equilibrium behavior for the calculation of separation unit.

1.4 The relevancy of the project

This project could be used as one of the trial and analysis on the iCON simulator. The large hydrocarbon is being used as a trial component. Refinery oil; coker heavy residue consumed a huge amount of heat and large hydrocarbon molecule which is uncommon to iCON simulator. By doing this project the behavior of iCON and distillation process

could be analyzed. Some conclusions and recommendations could be done on the iCON features and performance.

Furthermore, this project is also to study on the behavior of the components mixture of the heavy residue. This study also reflected to the hypothetical components that significant in modeling of the fractionator column. The effect of the hypothetical components could be analyzed in form of their significances towards convergence of the iCON.

This project could be used as one of the analysis on optimization and lumping of delayed coker unit. By doing this research more optimization can be studied and employed in the future. Since delayed coker unit consist more than one unit thus more area could be justify for future optimization.

CHAPTER 2
LITERATURE REVIEW AND THEORY

2.1 Delayed Coker Fractionator

The delayed coker in which the fresh feedstocks is preheated, coked, and fractionated, consists of one heater and two coke drums. The fresh feedstocks, which is heated up to about 350 °C via the heat exchange with the hot product flows and the flue-gas in convection section is introduced into the bottom of the fractionating column to quench the high-temperature superheated reaction vapors. The preheated feedstocks from the bottom of the column, together with the condensed heavier ends from the reaction vapors, is pumped into the radiation house of the coking heater and quickly heated to slightly below 500 °C. After partially vaporized in the heater tubes and passed through a 4-way switch valve, the feedstocks are then introduced into one of the two coke drums where the coking reactions are taken place. Water with high pressure is therefore injected into the heater tubes to minimize the coke deposition and to delay the coking reactions in the tubes. The superheated reaction vapors drawn out from the top of the coke drums are then back to the base of the fractionating column, and are further separated into various products according to their boiling points such as wet gas, naphtha, light gas oil (diesel) and heavy gas oil. The coke produced in the delayed coker is almost pure carbon containing some of the impurities of the feed such as sulfur and metals [3].

In typical delayed coker, as shown in figure 1, the charge is fed directly to the bottom of the fractionator, where it is mixed with fractionator bottom (recycle). The residuum (fresh feed) from the Hydrocracker Fractionation Unit enters the bottom section of the fractionator where material lighter than the desired cut point of the coker gas oil is flashed off and the remaining material combines with the recycle material condensed in the bottom of the fractionator to form the combined feed. This combined feed is then routed to the charge furnace where the liquid is heated to its incipient coking temperature to produce vapourization and mild cracking. Steam is injected into the furnace feed line

to prevent coke deposition in the furnace coils, increase tube velocity and reduce hydrocarbon partial pressure.

The vapour/liquid mixture then enters the bottom of the coke drum where the vapour experiences further cracking and the liquid experiences successive cracking and polymerization until it is completely converted to vapour and coke. The coke drum effluent vapour enters the fractionator where the hot vapours are quenched with wash oil. The condensed portion then forms the recycle stream and is recycled to the furnace for another pass through the coke drum. The condensed vapour is fractionated into gas, naphtha, jet fuel and gas oil.

The fractionator or combination distillation tower separates the coker overheads into gases, gasoline, diesel, heavy coker gas oil and recycle. An oversized fractionator can be used to maximize the amount of diesel product and minimize the heavy coker gas oil to the FCCU. Hot overhead vapors can cause coking in the lower section (wet).

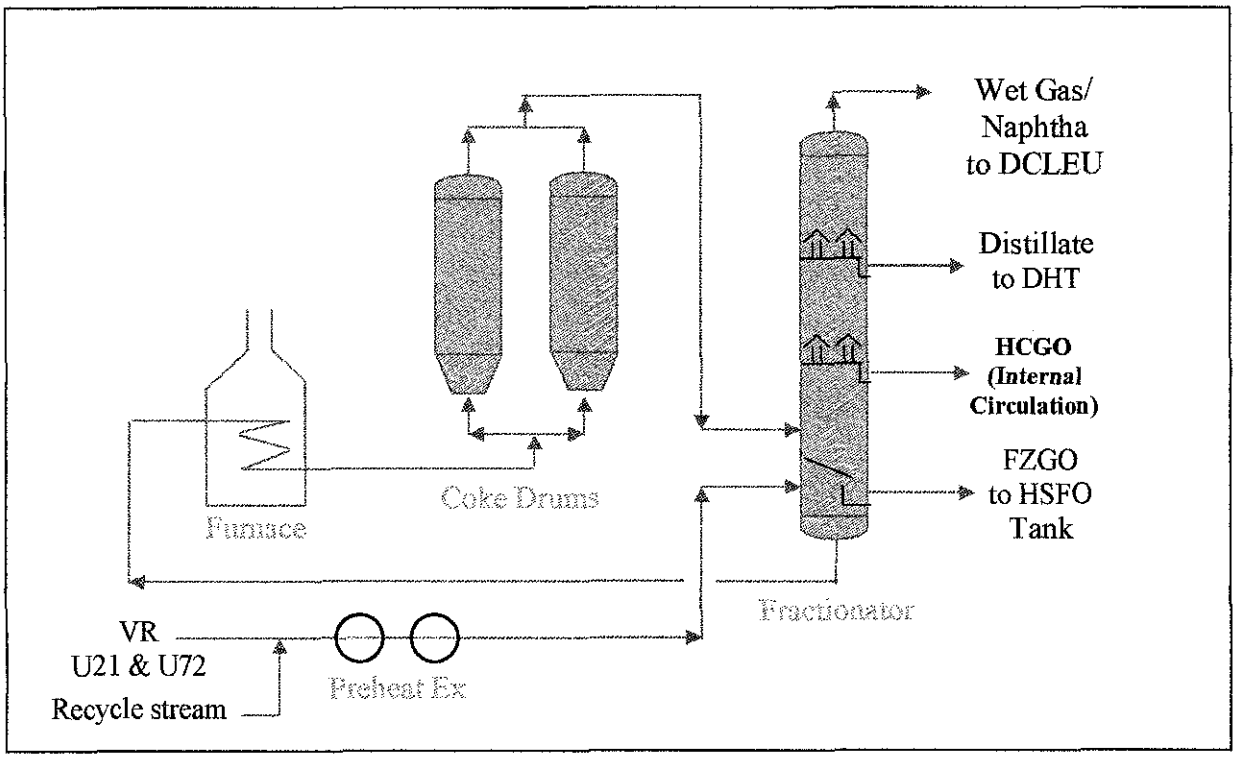

Figure 1: Overview of Delayed Coker Unit/Coking Section

The major amount of heat is removed in the heavy coker gas oil section by trapping out the oil and then extracting the heat with heat exchangers or steam boilers. The pressure in the fractionator and also the coke drums is controlled by the gas compressor at the top of the fractionator.

Coker feed is heated by a series of heat exchangers in the Feed and Preheat section. Vapors rising up the bubble tower from the flash zone are quenched by a series of pumparound cooling loops, the first of which is the flash zone gas oil (FZGO) circuit. It is followed by the HCGO and LCGO circuits, from which intermediate streams are drawn off for further processing. The HCGO stream splits into four streams. One stream serves as HCGO coke drum vapor quench. A second stream is sprayed into the flash zone section of Coker Fractionator to condense the heavy hydrocarbons in the coke drum vapor. Another stream is used as spray/ reflux to Coker Fractionator as HCGO pumparound. Coker distillate is splits into two streams. Coker distillate pumparound is pumped by the Coker Distillate Pumparound pumps where it split into two streams. One stream is routed Coker Fractionator and the other stream is routed to the Coker Gas Recovery Plant (GRP) to reboil the stripper tower. The Coker Fractionator overhead vapor is cooled in the Fractionator Overhead Condenser.

2.2 Fractional Distillation of Crude Oil

Crude oil or also called petroleum is a mixture of different hydrocarbons. Many useful products can be made by crude oil. But first the useful ones must be extracted from the crude oil and separated from one another. Different hydrocarbon chain lengths all have progressively higher boiling points, so they can all be separated by distillation. Each different chain length has a different property that makes it useful in a different way.

Fraction is made from different components of the crude oil. The fractions are separated by a process which called fractional distillation. This process is based on the principle that different substances boil at different temperatures. Crude oil can be fractioned into

naphtha (naphtha is made into petrol for cars, kerosene (kerosene is made into jet fuel) and residue. Residue can be processed further to get more useful products. The fractional distillation is worked by evaporate the mixture, and then cool it. As the mixture cools, the kerosene condenses first, and the naphtha condenses later.

The main equipment is a tall cylinder called a fractionator or fractional distillation column. Inside this column there are many trays, or horizontal plates, all located at different height. Each tray collects a different fraction when it cools to its own boiling point and condenses.

The residue is heated through coker furnace and some heat extracted form products such as FZGO at temperature around 170 degC, which makes most of the oil evaporate. The fluid then enters the column. As the vapour moves up through the fractionator, each fraction cools and condenses at a different temperature. As each fraction condenses, the liquid is collected in the trays. Products with higher boiling points condense on the lower trays in the column. Products with lower boiling points condense on the higher trays.

2.3 Components Mixture and Hypothetical Components

Historically, pseudo components are represented as components with defined normal boiling point, specific gravity and molecular weight. The normal boiling point is estimated from the TBP assay. The component specific gravities are usually based on the pseudo component normal boiling point and bulk gravity of the crude oil. Molecular weights are usually correlated using the normal boiling point and gravity. Sometimes, gravity curves and molecular weight curves are also available, and can be used for the estimation of the pseudo component gravity and/or molecular weight [2].

By means of boiling point, specific gravity, and molecular weight, the physical properties for the thermodynamic models can be determined. The most important physical properties necessary for the simulation of crude distillation systems are the critical temperature, critical pressure and acentric factor, and the ideal gas heat capacities [2].

Vapor-liquid equilibrium can be estimated using an equation of state such as Advanced_Peng-Robinson from the knowledge of critical properties and acentric factor, as well as the thermodynamic residuals for enthalpy, heat capacity, and entropy [2]. Enthalphies, heat capacites and entropies can be calculated by the knowledge of ideal gas heat capacities and from this point onwards any balance of interest in the component mixture can be constructed.

The concept of pseudo components is very useful but there are some limitations [2]:

- The three parameters - normal boiling point, specific gravity, and molecular weight - are not sufficient to define the chemical structure of the pseudo component.
- In actuality each pseudo component represents a series of actual chemicals with defined structures. These real components behave differently in mixing and reactions. Therefore pseudo components are not adequate to represent most of the refinery reactors.
- In addition, because of the lack of information related to chemical structure pseudo components are not adequate to assist in the calculation of many refinery tests such as octane numbers, pour point, cloud point, etc.
- The various refinery reaction processes alter the relationships of normal boiling point, gravity and molecular weight for pseudo components. All generalized correlations used in process simulators have a bias toward unreacted pseudo components. In particular, there is a bias toward paraffins due to the large amount of pure component data available for this class of chemicals and therefore their use in the construction of generalized correlations.

2.3.1 Wet Gas

Wet gas is a geological term for a mixture of hydrocarbons that contain a significant amount of liquid or condenseable compounds heavier than ethane. This compound may include propane or butane [4].

2.4 Simulation and optimization of large distillation column.

Wayburn and Seader (1983) proposed a mathematical model of great generality for distillation based (interlinked or noninterlinked) separation processes and a robust method of solution. While their procedure addressed the important issues of flexibility and robustness, it did not address the question of computational time with special emphasis on large distillation simulation problems [5].

In petroleum engineering, the usual way to reduce the size of the problem when simulating operations with distillation is to lump the components present in the oil mixture. Montel and Gouel (1984) proposed a lumping scheme based on the similarities of few properties of all the components [5]. Leibovici (1993) has established a procedure for the estimation of properties of the pseudo components which is accurate and completely consistent with the thermodynamic model to be used [5].

When a simulation of a flash operation is performed, the detailed results can be obtained by using a lumping procedure i.e, generating the detailed mixture results from the "lumped" solution [5]. It is obvious that employing a detailed description of the mixture in process optimization will lead to more accuracy and quality of the resulting solution.

2.5 Modelling the Coker's Fractionator by using iCON

The oil refining processes present many challenges to the simulation user. While the unit operations and thermodynamics are not generally difficult to model, the accurate characterization of petroleum as defined components or "pseudocomponents" can be difficult. All distillation problems can be approached with the following methodology. First, define the separation in terms of "key components" and, secondly, determine the "separation zones" in the column between adjacent products. Key components separate into the top and bottom products, with the light key recovered primarily in the overhead product and the heavy key recovered in the bottom product. The separation zone is the range of distributed components, which are components which appear in both products in significant quantities. Non-distributed components have little effect on the separation

Separation zones are very important for complex fractionators such as crude and vacuum units. There must be at least 6-8 significant pseudocomponents in these zones for the solution to be meaningful. Fewer pseudocomponents cause the separation to be discontinuous and

CHAPTER 3
METHODOLOGY

3.1 Research Methodology

Research of the delayed coker unit is to be done. A thorough search is made through the internet and from the libraries to collect all available information on the delayed coker unit context. The research is focusing on optimizing and modelling on delayed coker unit which involve fractionator column.

The actual operating data and some guidance are obtained from Petronas Penapisan Melaka. This actual operating data is needed in order to validate all the simulation data. Hypothetical data is needed to simulate streams with components that are not in available in the ICON component library such as FZGO, HCGO, and vacuum residue. After hypothethical components have been established, the project starts with simulating the streams in ICON simulation.

For the iCON simulation, some manuals are referred in obtaining more understanding on their features and functions. Trial and error is also applied since some of the features are not clear. The trial and error are based on the previous work. Comparison and analysis is performed based on the previous work.

3.2 Project Activities

3.2.1 Thermal data extraction

The procedure begins with gathering information of the components mixture from literature review and some reference books. This is done by obtaining the stream density, molecular weight and boiling temperature. All the data are plugged in the iCON environment to create hypothetical components. Some of the components are already established in the iCON components environment such as wet gas and naphtha. Wet gas is contained of ethane, methane, propane, propylene, butane, and butylenes. After all the

data have been inserted in the hypothethical component data, iCON will calculate all the remaining data such critical pressure, critical pressure, and enthalpy.

Below is the procedure to gather the initial data of the components mixture.

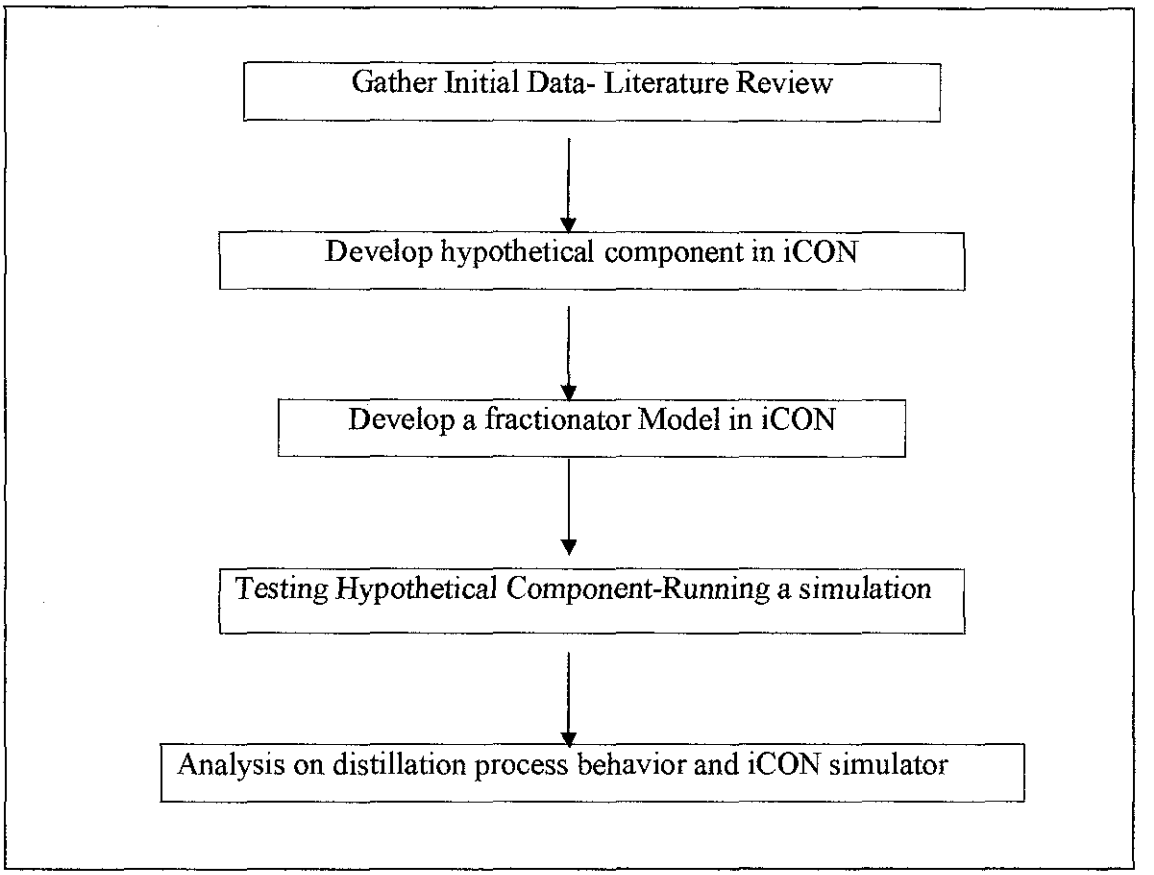

Figure 2 : Thermal Data Extraction Procedure

- Gather initial data. The initial data such as density, molecular weight, and boiling point are gathered for all components; wet gas, HCGO, FZGO and distillate.
- Develop hypothetical component- hypothetical is developed in the iCON environment by inserting the boiling point value or molecular weight/ density. In contra try, the wet gas stream was simulated in iCON without developing any hypothetical group. This is because the wet gas contain of hydrogen, ethane and all the light components until C5 components. All the components has been established in the iCON. For the hypothetical components, iCON will calculate others physical properties.

- Develop a fractionator Model in iCON – process stream is added and the design value of temperature, pressure, fraction of every component and flow rate are specified. After the process stream or feed stream is converged the fractionator tower is added to the iCON environment. Degree of freedom (DOF) of this tower need to be fulfilled first. The degree of freedom shows how many specifications are available before the system of equations representing the tower material and energy balance relationships can be solved. Then, specified the tower pressure and number of stages. The iteration can be started.
- Testing hypothetical components – hypothetical is based on the assumption value. Thus the validation of the hypothetical components needs to be performed to ensure the iteration of the tower run smoothly.
- Analysis on the simulation and distillation behavior – analysis can be made after all the components are valid and iteration started. The distillation behavior is different with the change in specified value such as changing the flow rate.

3.2.2 Develop the fractionator Model using iCON

Next step is developing the process stream and analyzing iCON and distillation behavior. The process stream is illustrated as figure 3 below. For initial stage all the operating data are inserted using the PPM's operating data. The design value of temperature, pressure, number of stages, and flow rate are specified. The lumping is running contiously until the model converges. By converging the model, it shows that the hypothetical components are all correct.

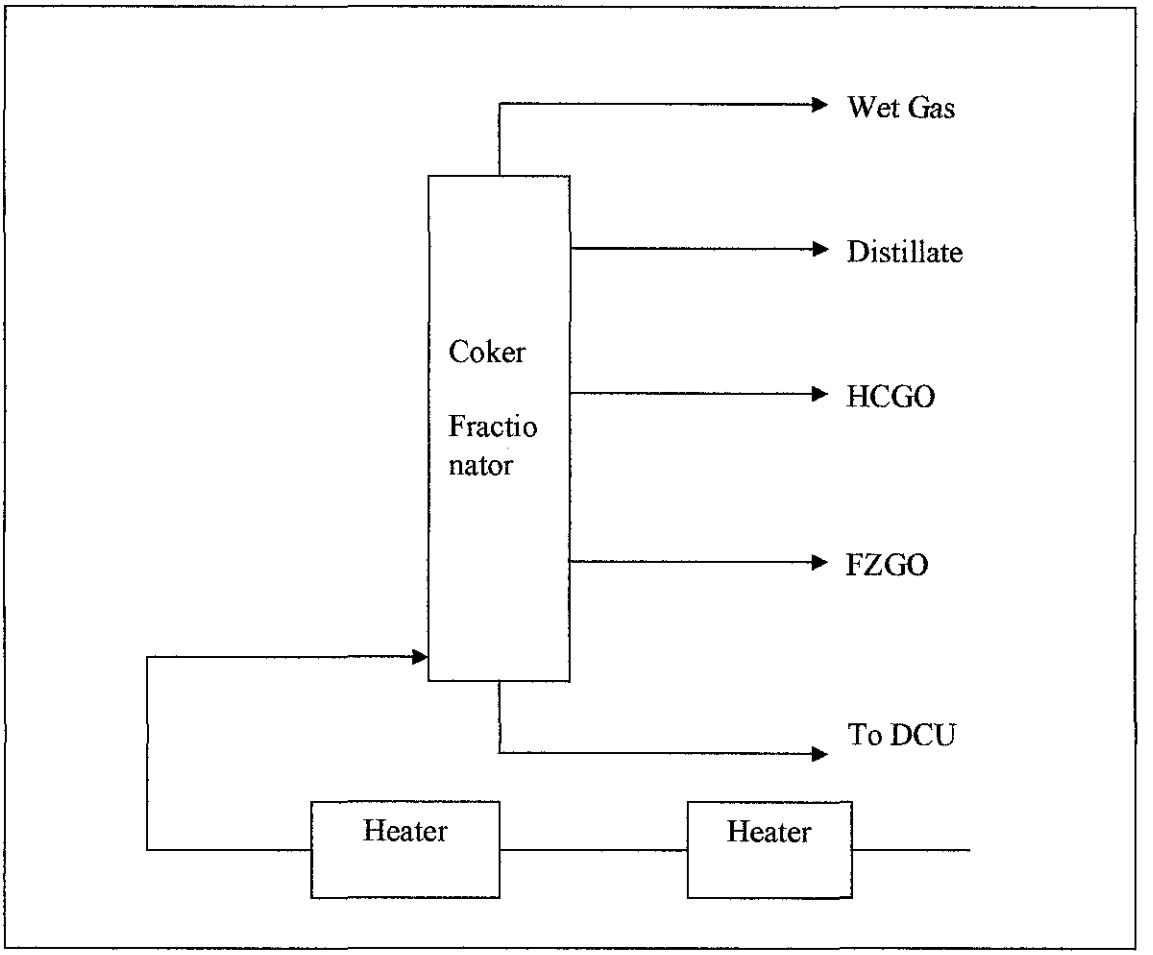

Figure 3: Model of Process Stream created in the ICON

3.3 Key Milestone

Table below is the suggested key milestone for this whole project.

Table 1: Suggested Key Milestone

Detail/Month	Aug	Sept	Oct	Nov	Dec	Jan	Feb	Mac	Apr
Data gathering/Literature Review									
Create Hypothetical Component									
Create Process Stream									
Simulated Creep Test Run									
Process Modification									

3.4 Project Tools

This project work needs a few items to be executed. For the engineering calculations and result analysis, Microsoft Excel application will be used. ICON simulation software is used as Process Unit simulation which is the main contribution in this whole project.

CHAPTER 4
RESULT AND DISCUSSION

4.1 First Analysis on Hypothetical Component and iCON

At the first analysis the following data is used. Hypothetical component are typically hydrocarbon compound with carbon number greater than 30. The resulted hypothetical groups represent four streams in the DCU one of them is the unit fresh feed while the other three are fractionator products which will go to some further treatment. The density need to be defined first since it is the data needed in the ICON simulation. The molecular weight estimation has been done as tabulated in the table below.

Table 2: Molecular Weight Estimation

Vacuum Residue			
Proposed Formula	No of 'C' atom	No of 'H2' atom	MW
$C_{(n)}H_{(2n+2)}$	50	102	702
	60	122	842
	70	142	982
	80	162	1122
Flash Zone Gas Oil			
Proposed Formula	No of 'C' atom	No of 'H2' atom	MW
$C_{(n)}H_{(2n+2)}$	40	82	562
	45	92	632
	55	112	772
	65	132	912
Heavy Coker Gas Oil			
Proposed Formula	No of 'C' atom	No of 'H2' atom	MW
$C_{(n)}H_{(2n+2)}$	30	62	422
	25	52	352
	15	32	212
	10	22	142

Appendix 1, 2 and 3 are the hypothetical components that have been created in the iCON simulator. The hypothetical components are created by plugged in the molecular weight value only. Below is the composition data of components mixture given by Petronas Penapisan Melaka.

Table 3: Composition of Component Mixture of Delayed Coker Residue

Component	mol %	mw
Methane	0.2103	16.04246
Ethane	8.62E-02	30.06904
Propane	5.20E-02	44.09562
n-Butane	2.05E-02	58.1222
Propylene	1.32E-02	42.07974
FZGO	0.100155	318.4057
HCGO	3.70E-08	278.0657
Distillate	0.150699	222.7989
Hydrogen	0.230433	2.01588
Ethylene	8.51E-03	28.05316
Isobutane	5.31E-03	58.1222
1-Buthene	1.53E-02	56.10632
Naphthalene	0.107415	128.1705

Data from a Penapisan Melaka are used at stage 49 at operating pressure 780 kpa and having one feed entering the column on stage 40 and two products is considered. Feed enter at 191 kgmole/hr. unfortunately iCON could not calculate this operating condition. The result is shown in appendix 1- 8.

4.2 Testing the Hypothetical Components

From the data above iCON could not performed the iteration. The trial and error is performed to study the source of error. Other hypothetical components are developed with the different process. The data is listed below.

Table 4: Tapis Blend of Light End Unit

TAPIS

Comp.	SPGR	BP (F)	Cut %	Fraction	Flow KBD
H20	0.9990	100.0000	0.1000	0.0010	0.0200
NaCl	2.1600	2669.0000	0.1000	0.0010	0.0200
LN+	0.5610	45.1000	1.7040	0.0170	0.3401
LVN	0.6571	146.5000	8.2360	0.0822	1.6439
HVN	0.7533	302.6000	18.4600	0.1842	3.6846
KERO	0.7931	444.5000	13.6000	0.1357	2.7146
ADO	0.8248	604.0000	40.4000	0.4032	8.0639
LSWR	0.8542	967.6000	17.6000	0.1756	3.5130
		Total	*100.2000*	*1.0000*	**20.0000**

The above data, LN+, LVN, HVN, Kero, ADO and LSWR are the hypothetical components. The data is used to iterate the tower of light end unit. iCON can calculated and converged the tower as shown in figure below.

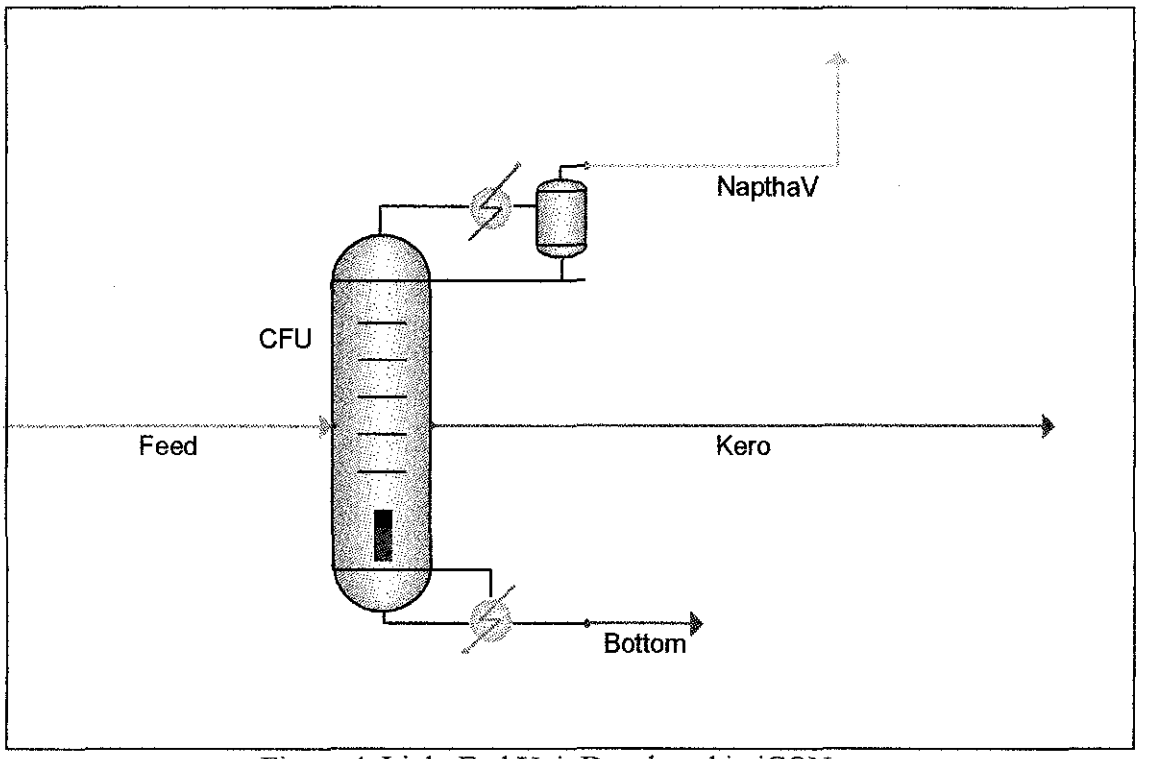

Figure 4: Light End Unit Developed in iCON

Thus, from the above simulation, it is clear that the first hypothetical component is cause of the failure in converge the distillation tower. The iteration is taken a few minute to converge the tower.

4.3 Discussion on Hypothetical Component

Separation zones are very important for complex fractionators such as crude and vacuum units. There must be at least 3 significant pseudocomponents properties in these zones for the solution to be meaningful. Fewer information of the pseudocomponents cause the separation to be discontinuous which the problem comes from the hypothetical component itself.

A small change in the column operation may predict a large change in the product flows. This does not correspond to the actual operation in which the crude oil is a continuum of components. Note that the results for these types of columns depend largely on the separation zone components. Non-distributed components have little effect. The critical separation zone for crude and vacuum columns is the flash zone. TBP distillations cannot reach the cutpoint temperature between the heavy gas oil and wet gas components.

For complex fractionators such as crude, vacuum, FCC main fractionators, etc, the definition of the pseudocomponents is much more important than the number of theoretical trays used in the model. To define the column feed is to define the column products for such columns. The number of theoretical trays in the model has little influence on the separations. The separations and resultant product distillations depend mostly on the product rates. Thus, to lower the end point of a product, the rate of the next lower product must be increased. The plant operation behaves in this fashion also.

4.4 Second Analysis on Hypothetical Component and iCON

Second iteration is continuing since first analysis could not converge by iCON. The hypothetical component is reformulated.

Below are the tabulated data for the new hypothetical component.

i) Heavy Coker Gas Oil

- Vapor Pressure.9 psia at 100 °C; < 1 psia (Reid Method)
- Vapor Density(Air = 1): Not available
- Boiling Range : 220 - 620 °C
- Specific Gravity: 0.98 at 20 deg C

ii) Light Coker Gas Oil

- Vapor Pressure <1 psia (Reid Method)
- Vapor Density (Air = 1): Not available
- Boiling Range : 200 - 450 °C
- Specific Gravity 0.90 at 20 deg C

The feed is entering at 100 kgmole/hr at 10 stages. The outlet stream consists of distillate, wet gas, HCGO and FGGO. By using these new data, iCON can calculate it and converged. The result is shown in appendix 9 - 10.

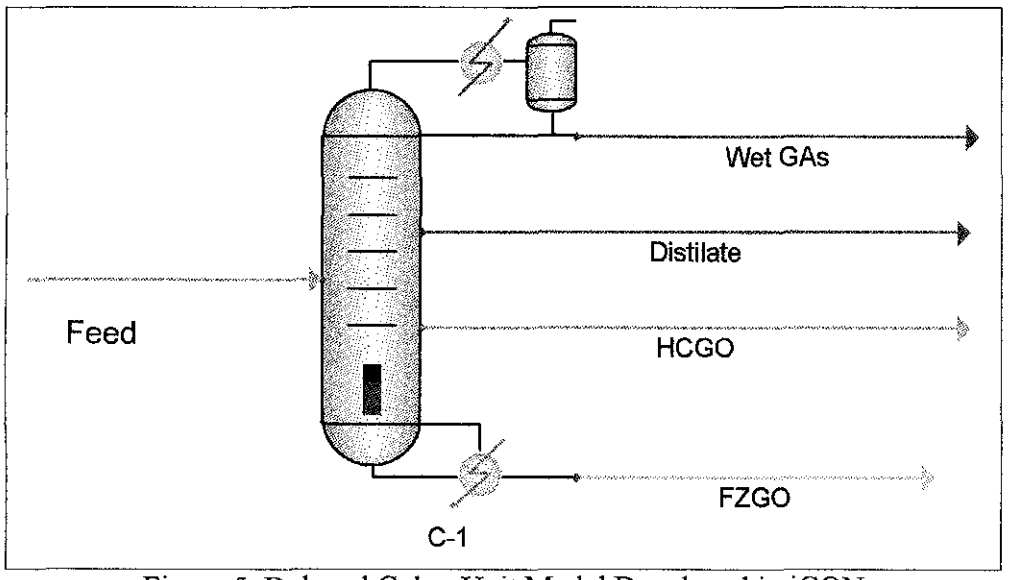

Figure 5: Delayed Coker Unit Model Developed in iCON

4.5 Discussion on Second Analysis

Given the complex nature of the feed to fractionator, and the fact that the desired products from coker frctionator are distributions of hydrocarbons between certain distillation ranges, when designing or simulating distillation towers it is important to be able to estimate the yields of the different products based on the feed basic properties.

From the 2 analysis, it is important to ensure that the hypothetical components are correctly assumed. Since the delayed coker unit is very complex unit and involve large hydrocarbon molecule, it is difficult to model the fractionation unit. The process data must be matched. The tower pressure and the inlet value must be correctly matched in order to converge the tower. Otherwise, the simulation will take a long time to converge even could not be solved.

This model is validating with the Petronas Penapisan Melaka's model. Unfortunately the process stream's value given by PPM could not be used in the iCON simulator. This is because they use different simulator. Hence all the value could not be used in developing the delayed coker unit model using iCON.

CHAPTER 5
CONCLUSION AND RECOMMENDATIONS

In conclusion, hypothetical components must be correctly assumed in order to do a lumping procedure on distillation unit. It is proved that normal boiling point, specific gravity, and molecular weight are not sufficient to define the chemical structure of the pseudo/hypothetical component. It is because of the lack of information related to chemical structure pseudo components are not adequate to assist in the calculation of the distillation unit. The behavior of the distillation unit can be observed trough the component involved in the distillation process and operating condition itself.

Given the complex nature of the feed to fractionator, and the fact that the desired products from coker frctionator are distributions of hydrocarbons between certain distillation ranges, when designing or simulating distillation towers it is important to be able to estimate the yields of the different products based on the feed basic properties. The complex reaction of delayed coker unit need correct process data in simulating it.

After some analysis and evaluation, existing delayed coker unit reveal that there exist potentials for energy-use improvement for the unit optimization. Some improvements can be formulated in future on delayed coker unit after done more research and analysis. Since the delayed coker unit involved large amount of heat, it is important to consider energy optimization on this unit. Maybe analysis effect on heat exchanger network can be analyzed and some improvement can also be made on the network. However, for future work delayed coker unit maybe can be improved further by looking at the other section of the unit such as furnace duty and coker unit itself in order to optimize the production.

REFERENCES

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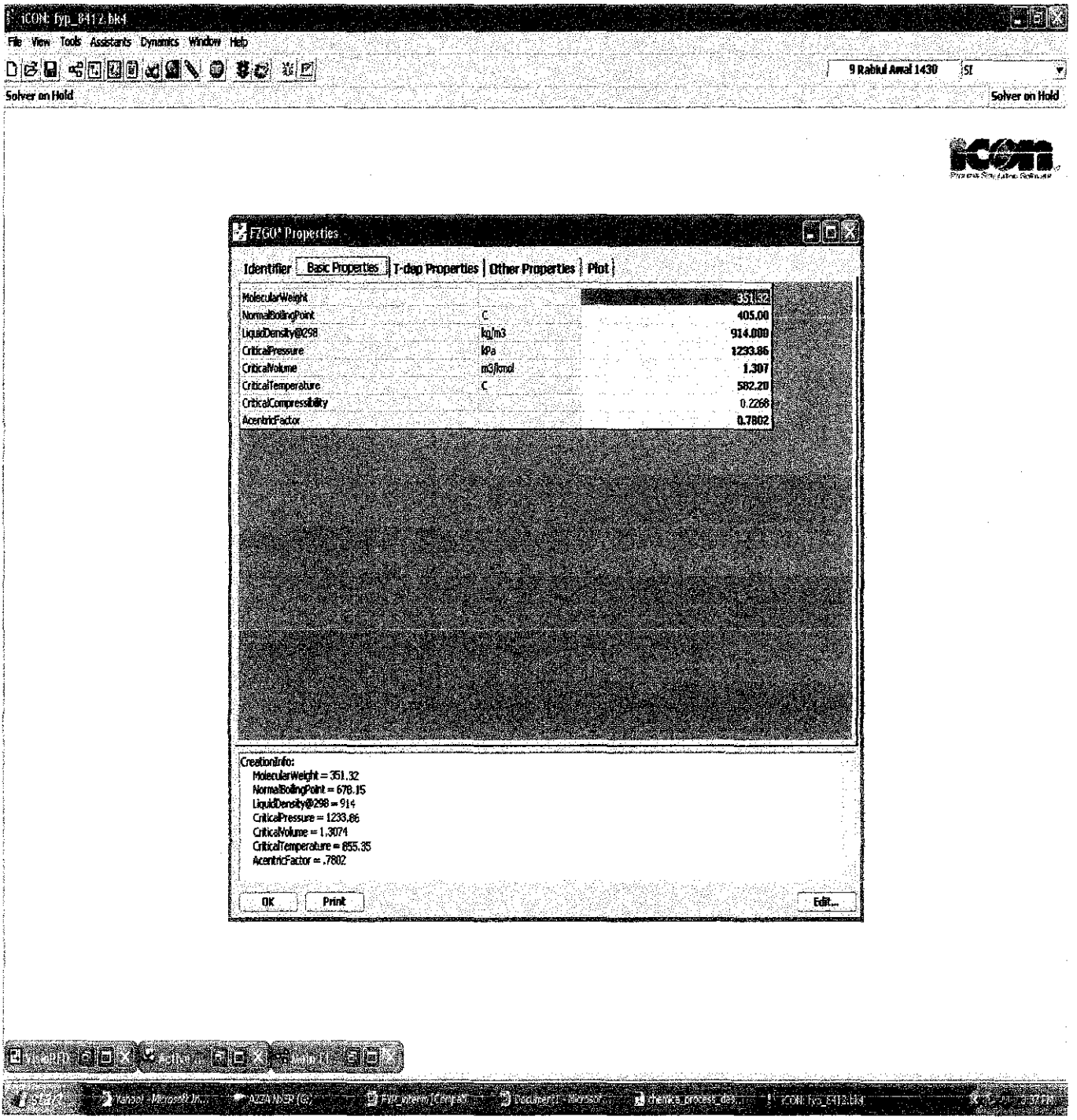
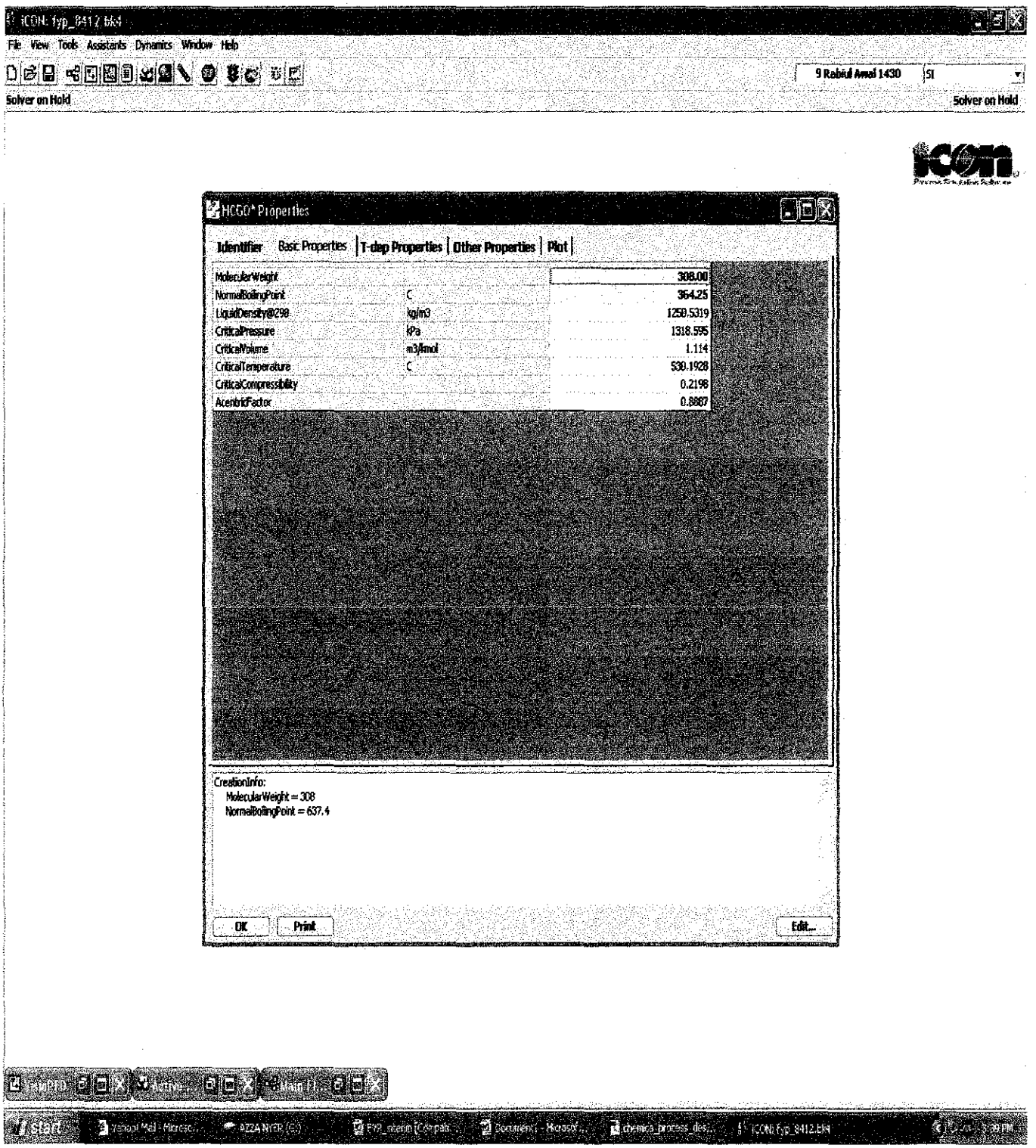
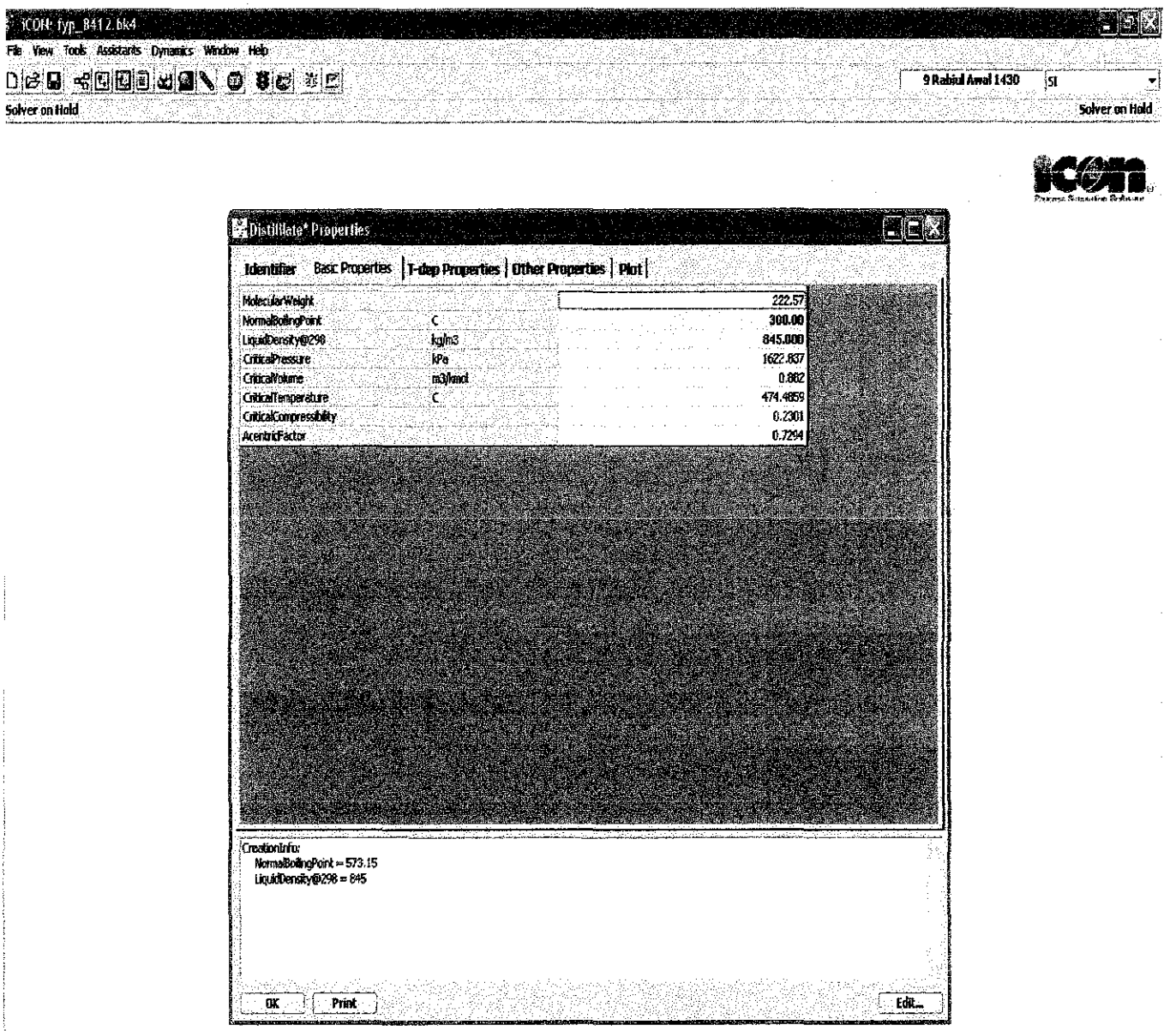
Appendix 1


iCON: fyp_0412.bk4
File View Tools Assistants Dynamics Window Help
9 Rabiul Awal 1430
Solver on Hold
Solver on Hold
FZGO* Properties
Identifier
Basic Properties
T-dep Properties
Other Properties
Plot
MolecularWeight 351.32
NormalBoilingPoint C 405.00
LiquidDensity@298 kg/m3 914.000
CriticalPressure kPa 1233.86
CriticalVolume m3/kmol 1.307
CriticalTemperature C 582.20
CriticalCompressibility 0.2268
AcentricFactor 0.7802
CreationInfo:
MolecularWeight = 351.32
NormalBoilingPoint = 678.15
LiquidDensity@298 = 914
CriticalPressure = 1233.86
CriticalVolume = 1.3074
CriticalTemperature = 855.35
AcentricFactor = .7802
OK
Print
Edit...

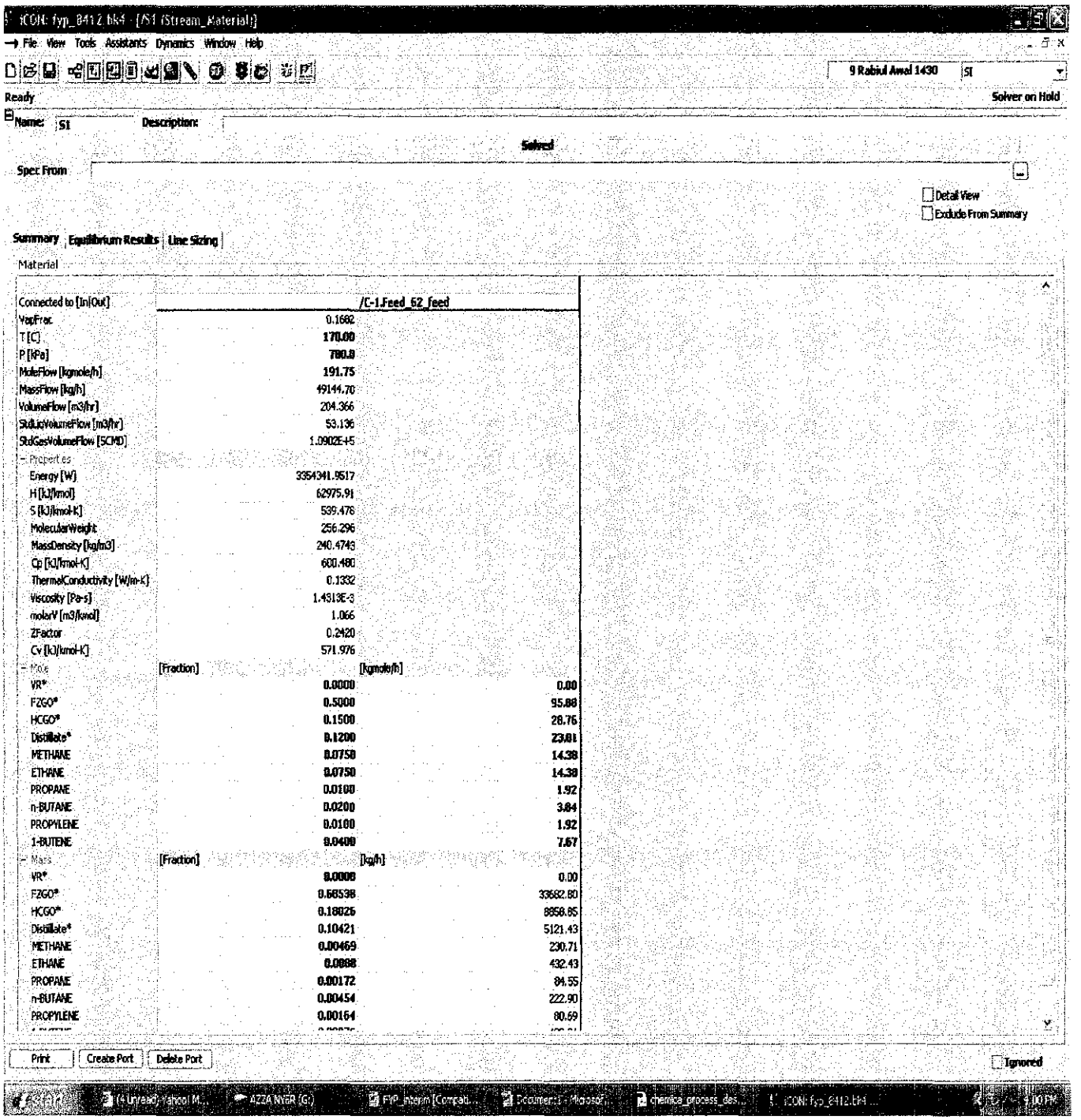
Appendix 2


ICON: fyp_8412.bk4
File View Tools Assistants Dynamics Window Help
9 Rabiul Awal 1430
SI
Solver on Hold
Solver on Hold
HCGO* Properties
Identifier
Basic Properties
T-dep Properties
Other Properties
Plot
MolecularWeight 308.00
NormalBoilingPoint C 364.25
LiquidDensity@298 kg/m3 1258.5319
CriticalPressure kPa 1318.595
CriticalVolume m3/kmol 1.114
CriticalTemperature C 530.1928
CriticalCompressibility 0.2198
AcentricFactor 0.8887
CreationInfo:
MolecularWeight = 308
NormalBoilingPoint = 637.4
OK
Print
Edit...

Appendix 3


ICON: fyp_8412.bk4
File View Tools Assistants Dynamics Window Help
9 Rabiul Awal 1430
SI
Solver on Hold
Solver on Hold
Distillate* Properties
Identifier
Basic Properties
T-dep Properties
Other Properties
Plot
MolecularWeight 222.57
NormalBoilingPoint C 300.00
LiquidDensity@298 kg/m3 845.000
CriticalPressure kPa 1622.837
CriticalVolume m3/kmol 0.882
CriticalTemperature C 474.4859
CriticalCompressibility 0.2301
AcentricFactor 0.7294
CreationInfo:
NormalBoilingPoint = 573.15
LiquidDensity@298 = 845
OK
Print
Edit...

Appendix 4



ICON: fyp_0412.bk4 - [/S1 (Stream_Material)]

File View Tools Assistants Dynamics Window Help

9 Rabiul Awal 1430 | S1

Ready | Solver on Hold

Name: S1 | Description:

Solved

Spec From

Detail View | Exclude From Summary

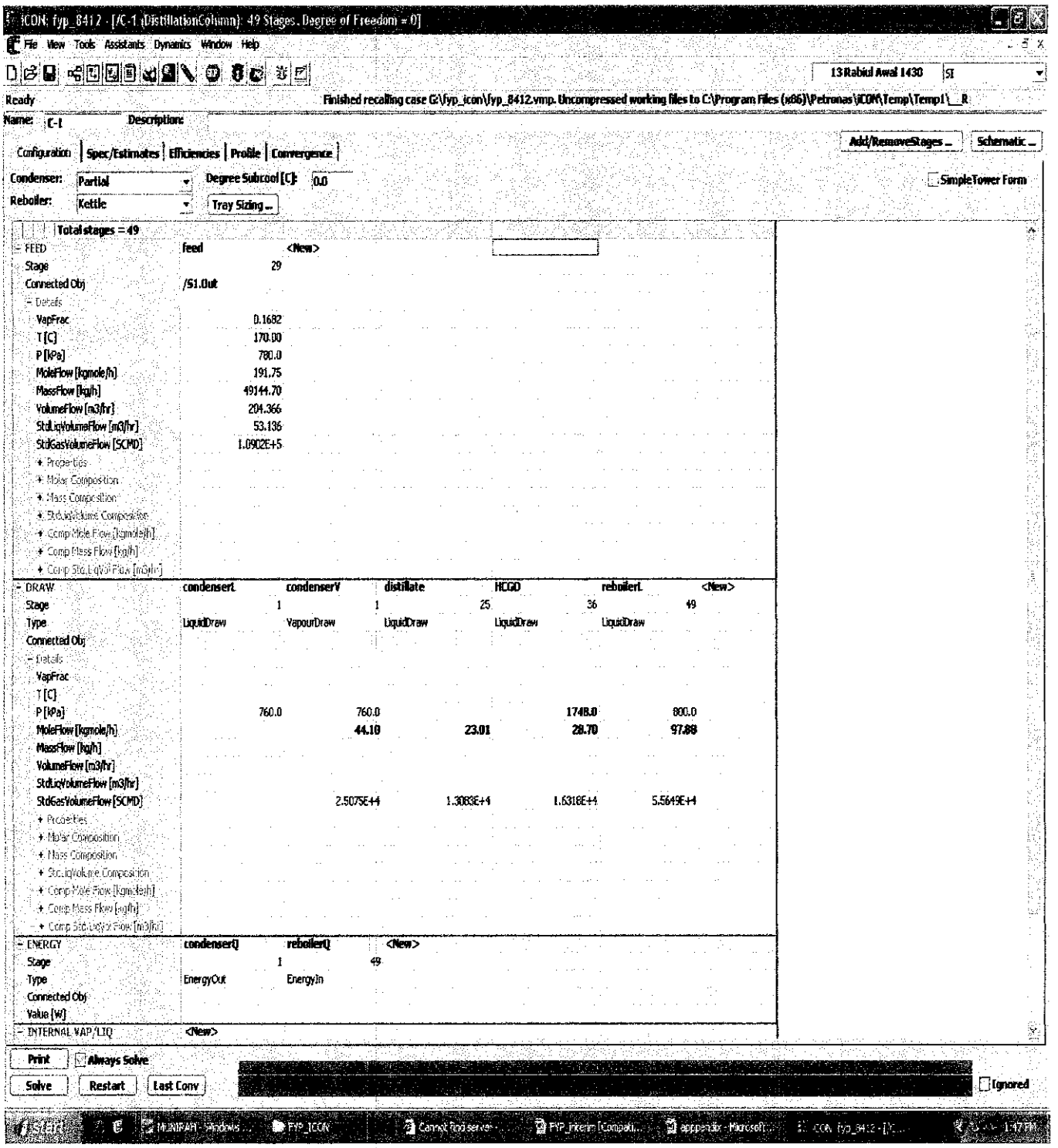
Summary | Equilibrium Results | Line Sizing

Material

Connected to [In|Out]	/C-1.Feed_62_feed	
VapFrac	0.1682	
T [C]	**170.00**	
P [kPa]	**780.0**	
MoleFlow [kgmole/h]	**191.75**	
MassFlow [kg/h]	49144.70	
VolumeFlow [m3/hr]	204.366	
StdLiqVolumeFlow [m3/hr]	53.136	
StdGasVolumeFlow [SCMD]	1.0902E+5	
- Properties		
Energy [W]	3354341.9517	
H [kJ/kmol]	62975.91	
S [kJ/kmol-K]	539.476	
MolecularWeight	256.296	
MassDensity [kg/m3]	240.4743	
Cp [kJ/kmol-K]	600.480	
ThermalConductivity [W/m-K]	0.1332	
Viscosity [Pa-s]	1.4913E-3	
molarV [m3/kmol]	1.066	
ZFactor	0.2420	
Cv [kJ/kmol-K]	571.976	
- Mole	[Fraction]	[kgmole/h]
VR*	**0.0000**	**0.00**
FZGO*	**0.5000**	**95.88**
HCGO*	**0.1500**	**28.76**
Distillate*	**0.1200**	**23.01**
METHANE	**0.0750**	**14.38**
ETHANE	**0.0750**	**14.38**
PROPANE	**0.0100**	**1.92**
n-BUTANE	**0.0200**	**3.84**
PROPYLENE	**0.0100**	**1.92**
1-BUTENE	**0.0400**	**7.67**
- Mass	[Fraction]	[kg/h]
VR*	**0.0000**	0.00
FZGO*	**0.68538**	33682.80
HCGO*	**0.18026**	8858.85
Distillate*	**0.10421**	5121.43
METHANE	**0.00469**	230.71
ETHANE	**0.0088**	432.43
PROPANE	**0.00172**	84.55
n-BUTANE	**0.00454**	222.90
PROPYLENE	**0.00164**	80.59

Print | Create Port | Delete Port | Ignored

Appendix 5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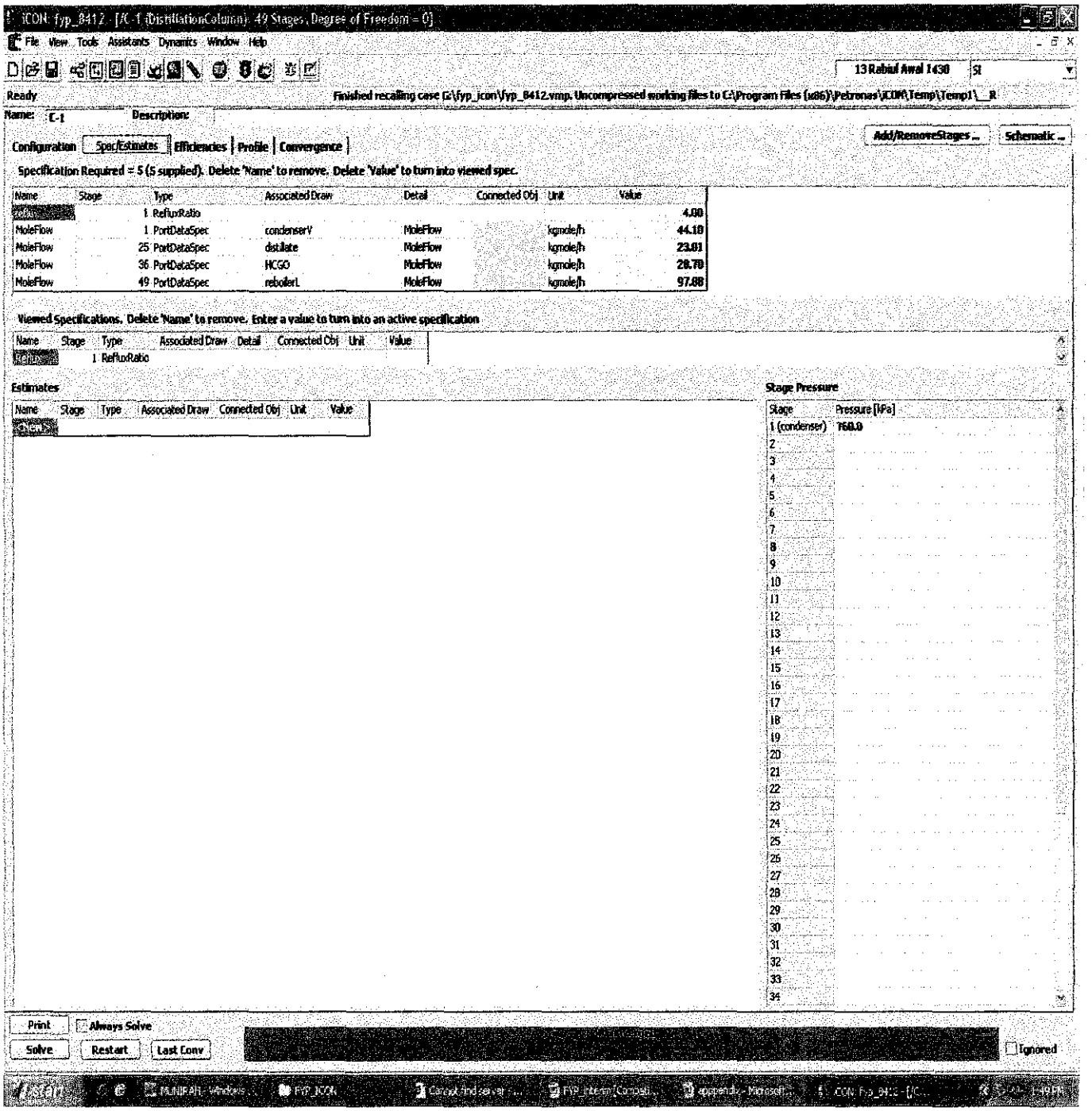
File View Tools Assistants Dynamics Window Help
13 Rabiul Awal 1430
SI
Ready
Finished recalling case G:\fyp_icon\fyp_8412.vmp. Uncompressed working files to C:\Program Files (x86)\Petronas\ICON\Temp\Temp1__R
Name: C-1 Description:
Configuration | Spec/Estimates | Efficiencies | Profile | Convergence
Add/RemoveStages ... Schematic ...
Condenser: Partial Degree Subcool [C]: 0.0
SimpleTower Form
Reboiler: Kettle Tray Sizing ...
Total stages = 49
FEED feed <New>
Stage 29
Connected Obj /S1.Out
Details
VapFrac 0.1682
T [C] 170.00
P [kPa] 780.0
MoleFlow [kgmole/h] 191.75
MassFlow [kg/h] 49144.70
VolumeFlow [m3/hr] 204.366
StdLiqVolumeFlow [m3/hr] 53.136
StdGasVolumeFlow [SCMD] 1.09O2E+5
Properties
Molar Composition
Mass Composition
Std Liq Volume Composition
Comp Mole Flow [kgmole/h]
Comp Mass Flow [kg/h]
Comp Std Liq Vol Flow [m3/hr]
DRAW condenserL condenserV distillate HCGD reboilerL <New>
Stage 1 1 25 36 49
Type LiquidDraw VapourDraw LiquidDraw LiquidDraw LiquidDraw
Connected Obj
Details
VapFrac
T [C]
P [kPa] 760.0 760.0 1748.0 800.0
MoleFlow [kgmole/h] 44.10 23.01 28.70 97.88
MassFlow [kg/h]
VolumeFlow [m3/hr]
StdLiqVolumeFlow [m3/hr]
StdGasVolumeFlow [SCMD] 2.5075E+4 1.3083E+4 1.6318E+4 5.5649E+4
Properties
Molar Composition
Mass Composition
Std Liq Volume Composition
Comp Mole Flow [kgmole/h]
Comp Mass Flow [kg/h]
Comp Std Liq Vol Flow [m3/hr]
ENERGY condenserQ reboilerQ <New>
Stage 1 49
Type EnergyOut EnergyIn
Connected Obj
Value [W]
INTERNAL VAP/LIQ <New>
Print Always Solve
Solve Restart Last Conv
Ignored

Appendix 6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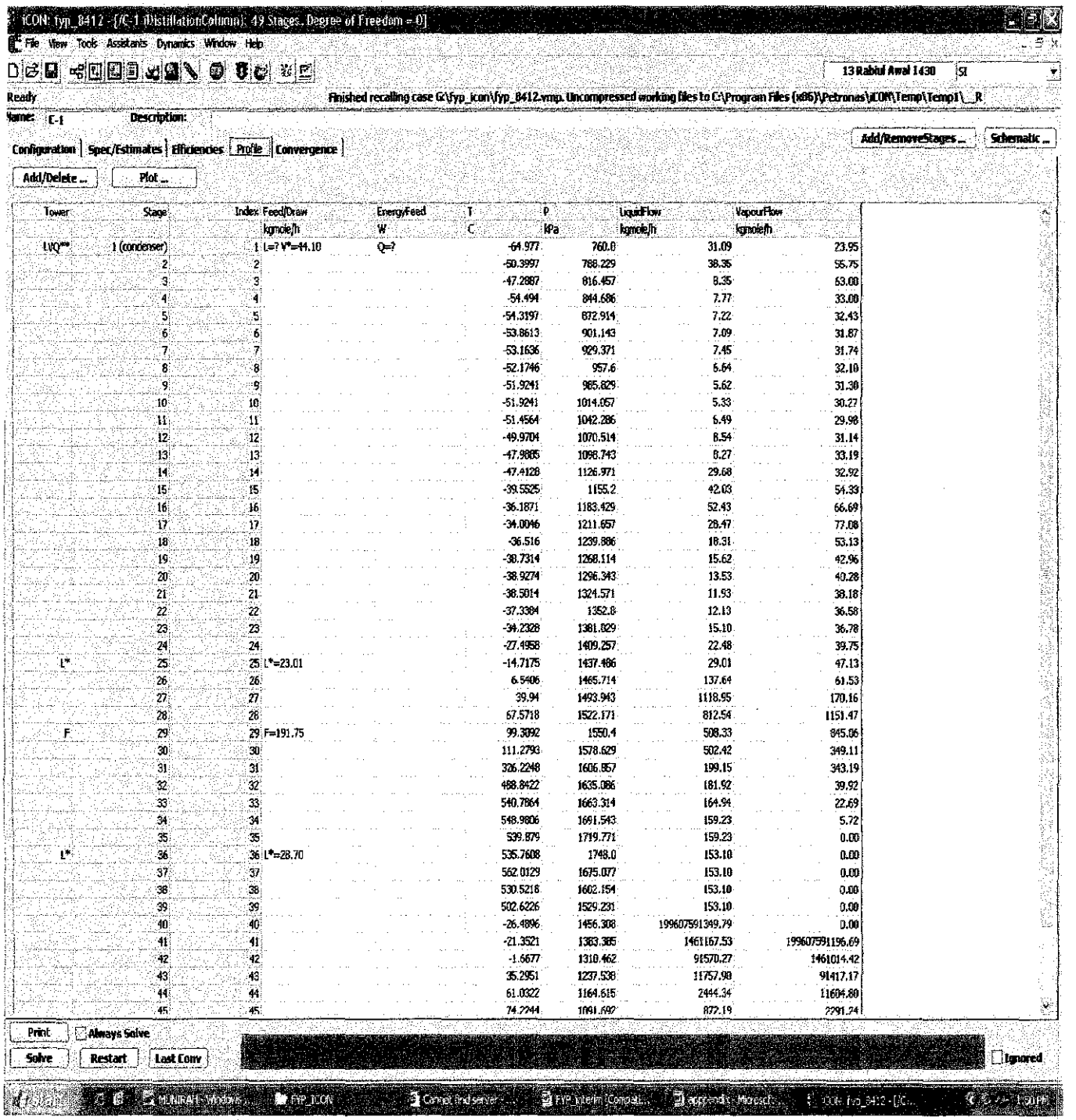
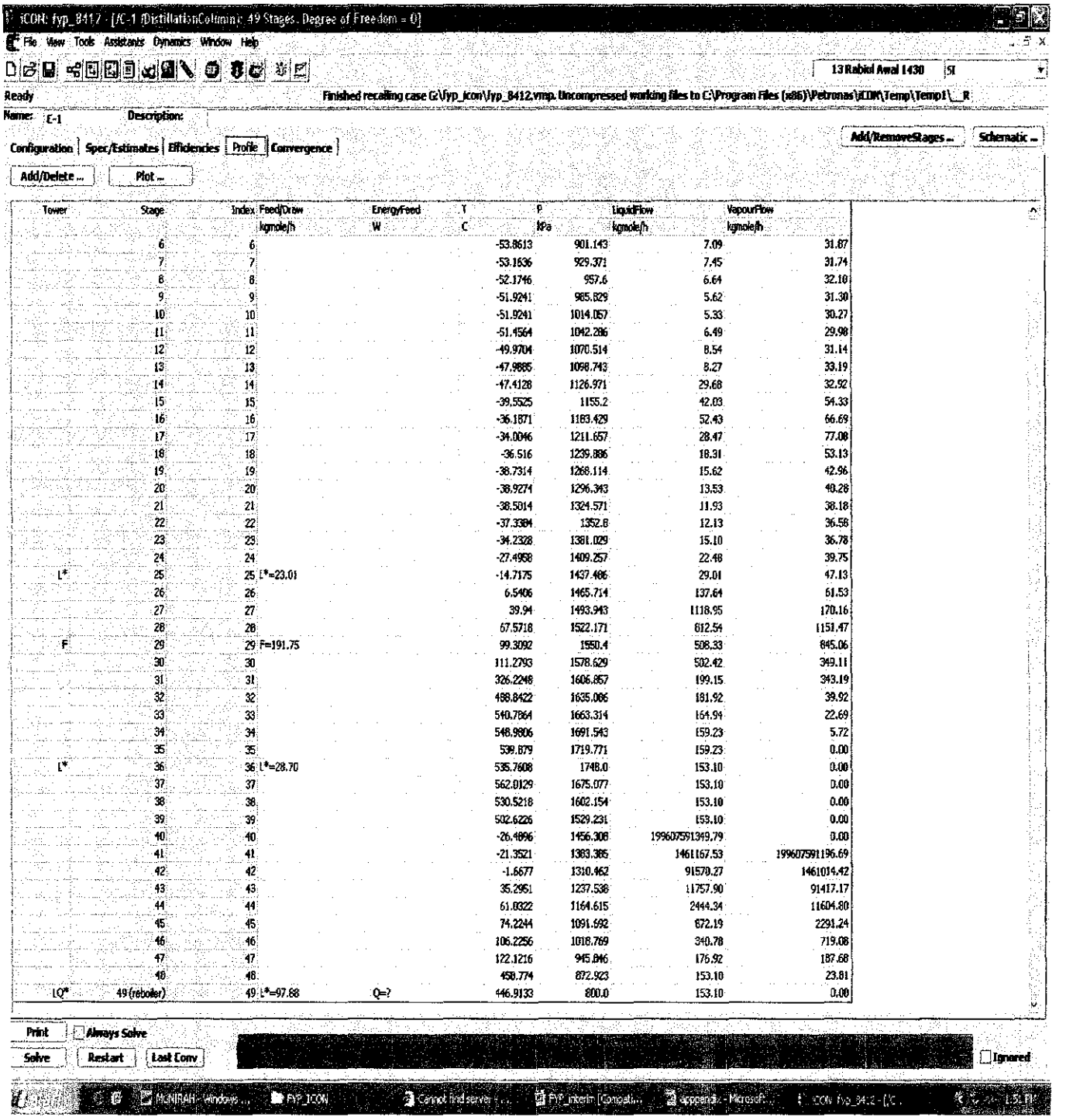
File View Tools Assistants Dynamics Window Help
13 Rabiul Awal 1430
Ready
Finished recalling case c:\fyp_icon\fyp_8412.vmp. Uncompressed working files to C:\Program Files (x86)\Petronas\ICON\Temp\Temp1__R
Name: C-1
Description:
Configuration
Spec/Estimates
Efficiencies
Profile
Convergence
Add/RemoveStages ...
Schematic ...
Specification Required = 5 (5 supplied). Delete 'Name' to remove. Delete 'Value' to turn into viewed spec.
Name Stage Type Associated Draw Detail Connected Obj Unit Value
RefluxRatio 1 RefluxRatio 4.00
MoleFlow 1 PortDataSpec condenserV MoleFlow kgmole/h 44.10
MoleFlow 25 PortDataSpec distillate MoleFlow kgmole/h 23.01
MoleFlow 36 PortDataSpec HCGO MoleFlow kgmole/h 28.79
MoleFlow 49 PortDataSpec reboilerL MoleFlow kgmole/h 97.88
Viewed Specifications. Delete 'Name' to remove. Enter a value to turn into an active specification
Name Stage Type Associated Draw Detail Connected Obj Unit Value
RefluxRatio 1 RefluxRatio
Estimates
Name Stage Type Associated Draw Connected Obj Unit Value
Stage Pressure
Stage Pressure [kPa]
1 (condenser) 760.0
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20
21
22
23
24
25
26
27
28
29
30
31
32
33
34
Print
Always Solve
Solve
Restart
Last Conv
Ignored

Appendix 7



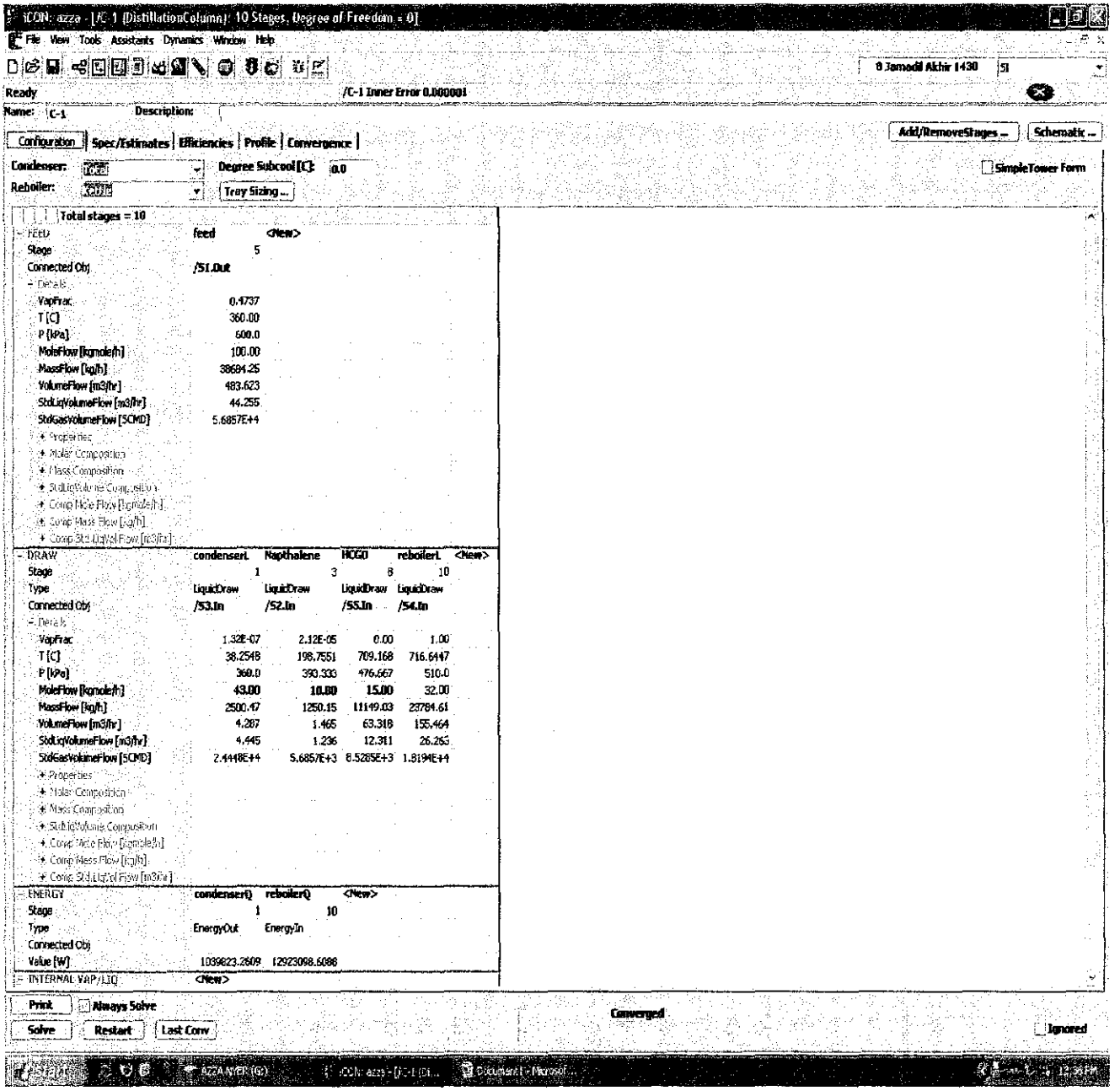
Tower	Stage	Index	Feed/Draw	EnergyFeed	T	P	LiquidFlow	VapourFlow
			kgmole/h	W	C	kPa	kgmole/h	kgmole/h
LVQ**	1 (condenser)	1	L=? V*=44.10	Q=?	-64.977	760.0	31.09	23.95
	2	2			-50.3997	788.229	38.35	55.75
	3	3			-47.2887	816.457	8.35	63.08
	4	4			-54.494	844.686	7.77	33.08
	5	5			-54.3197	872.914	7.22	32.43
	6	6			-53.8613	901.143	7.09	31.87
	7	7			-53.1636	929.371	7.45	31.74
	8	8			-52.1746	957.6	6.64	32.10
	9	9			-51.9241	985.829	5.62	31.30
	10	10			-51.9241	1014.057	5.33	30.27
	11	11			-51.4564	1042.286	6.49	29.98
	12	12			-49.9704	1070.514	8.54	31.14
	13	13			-47.9885	1098.743	8.27	33.19
	14	14			-47.4128	1126.971	29.68	32.92
	15	15			-39.5525	1155.2	42.03	54.33
	16	16			-36.1871	1183.429	52.43	66.69
	17	17			-34.0046	1211.657	28.47	77.08
	18	18			-36.516	1239.886	18.31	53.13
	19	19			-38.7314	1268.114	15.62	42.96
	20	20			-38.9274	1296.343	13.53	40.28
	21	21			-38.5014	1324.571	11.93	38.18
	22	22			-37.3384	1352.8	12.13	36.58
	23	23			-34.2328	1381.029	15.10	36.78
	24	24			-27.4958	1409.257	22.48	39.75
L*	25	25	L*=23.01		-14.7175	1437.486	29.01	47.13
	26	26			6.5406	1465.714	137.64	61.53
	27	27			39.94	1493.943	1118.95	170.16
	28	28			67.5718	1522.171	812.54	1151.47
F	29	29	F=191.75		99.3092	1550.4	508.33	845.06
	30	30			111.2793	1578.629	502.42	349.11
	31	31			326.2248	1606.857	199.15	343.19
	32	32			488.8422	1635.086	181.92	39.92
	33	33			540.7864	1663.314	164.94	22.69
	34	34			548.9806	1691.543	159.23	5.72
	35	35			539.879	1719.771	159.23	0.00
L*	36	36	L*=28.70		535.7608	1748.0	153.10	0.00
	37	37			562.0129	1675.077	153.10	0.00
	38	38			530.5218	1602.154	153.10	0.00
	39	39			502.6226	1529.231	153.10	0.00
	40	40			-26.4896	1456.308	199607591349.79	0.00
	41	41			-21.3521	1383.385	1461167.53	199607591196.69
	42	42			-1.6677	1310.462	91570.27	1461014.42
	43	43			35.2951	1237.538	11757.90	91417.17
	44	44			61.0322	1164.615	2444.34	11604.80
	45	45			74.2244	1091.692	872.19	2291.24

Appendix 8

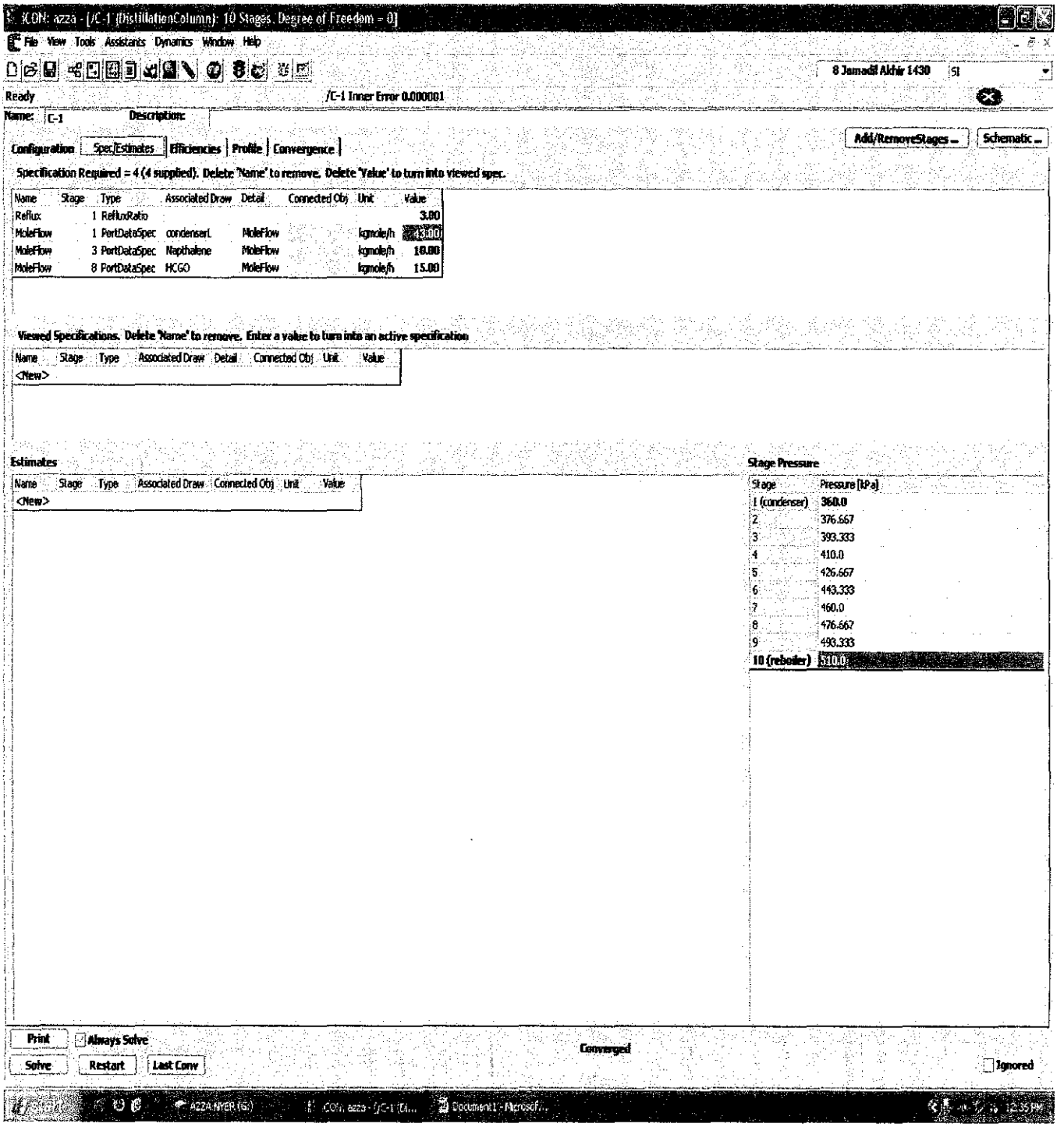


Tower	Stage	Index	Feed/Draw	EnergyFeed	T	P	LiquidFlow	VapourFlow
			kgmole/h	W	C	kPa	kgmole/h	kgmole/h
	6	6			-53.8613	901.143	7.09	31.87
	7	7			-53.1636	929.371	7.45	31.74
	8	8			-52.1746	957.6	6.64	32.10
	9	9			-51.9241	985.829	5.62	31.30
	10	10			-51.9241	1014.057	5.33	30.27
	11	11			-51.4564	1042.286	6.49	29.98
	12	12			-49.9704	1070.514	8.54	31.14
	13	13			-47.9885	1098.743	8.27	33.19
	14	14			-47.4128	1126.971	29.68	32.92
	15	15			-39.5525	1155.2	42.03	54.33
	16	16			-36.1871	1183.429	52.43	66.69
	17	17			-34.0046	1211.657	28.47	77.08
	18	18			-36.516	1239.886	18.31	53.13
	19	19			-38.7314	1268.114	15.62	42.96
	20	20			-38.9274	1296.343	13.53	40.28
	21	21			-38.5014	1324.571	11.93	38.18
	22	22			-37.3384	1352.8	12.13	36.56
	23	23			-34.2328	1381.029	15.10	36.78
	24	24			-27.4958	1409.257	22.48	39.75
L*	25	25	L*=23.01		-14.7175	1437.486	29.01	47.13
	26	26			6.5406	1465.714	137.64	61.53
	27	27			39.94	1493.943	1118.95	170.16
	28	28			67.5718	1522.171	612.54	1151.47
F	29	29	F=191.75		99.3092	1550.4	508.33	845.06
	30	30			111.2793	1578.629	502.42	349.11
	31	31			326.2248	1606.857	199.15	343.19
	32	32			488.8422	1635.086	181.92	39.92
	33	33			540.7864	1663.314	164.94	22.69
	34	34			548.9806	1691.543	159.23	5.72
	35	35			539.879	1719.771	159.23	0.00
L*	36	36	L*=28.70		535.7608	1748.0	153.10	0.00
	37	37			562.0129	1675.077	153.10	0.00
	38	38			530.5218	1602.154	153.10	0.00
	39	39			502.6226	1529.231	153.10	0.00
	40	40			-26.4896	1456.308	199607591349.79	0.00
	41	41			-21.3521	1383.385	1461167.53	199607591196.69
	42	42			-1.6677	1310.462	91570.27	1461014.42
	43	43			35.2951	1237.538	11757.90	91417.17
	44	44			61.8322	1164.615	2444.34	11604.80
	45	45			74.2244	1091.692	672.19	2291.24
	46	46			106.2256	1018.769	340.78	719.08
	47	47			122.1216	945.846	176.92	187.68
	48	48			458.774	872.923	153.10	23.81
LQ*	49 (reboiler)	49	L*=97.88	Q=?	446.9133	800.0	153.10	0.00

Appendix 9


ICON: azza - [/C-1 (DistillationColumn): 10 Stages, Degree of Freedom = 0]
File View Tools Assistants Dynamics Window Help
8 Jamadil Akhir 1430
Ready
/C-1 Inner Error 0.000001
Name: C-1 Description:
Configuration | Spec/Estimates | Efficiencies | Profile | Convergence
Add/RemoveStages ... | Schematic ...
Condenser: Total
Reboiler: Partial
Degree Subcool [C]: 0.0
Tray Sizing ...
SimpleTower Form
Total stages = 10
FEED feed <New>
Stage 5
Connected Obj /51.Out
Details
VapFrac 0.4737
T [C] 360.00
P [kPa] 600.0
MoleFlow [kgmole/h] 100.00
MassFlow [kg/h] 38684.25
VolumeFlow [m3/hr] 483.623
StdLiqVolumeFlow [m3/hr] 44.255
StdGasVolumeFlow [SCMD] 5.6857E+4
Properties
Molar Composition
Mass Composition
StdLiqVolume Composition
Comp Mole Flow [kgmole/h]
Comp Mass Flow [kg/h]
Comp Std Liq Vol Flow [m3/hr]
DRAW condenserL Napthalene HCGO reboilerL <New>
Stage 1 3 8 10
Type LiquidDraw LiquidDraw LiquidDraw LiquidDraw
Connected Obj /53.In /52.In /55.In /54.In
Details
VapFrac 1.32E-07 2.12E-05 0.00 1.00
T [C] 38.2548 198.7551 709.168 716.6447
P [kPa] 360.0 393.333 476.667 510.0
MoleFlow [kgmole/h] 43.00 10.80 15.00 32.00
MassFlow [kg/h] 2500.47 1250.15 11149.03 23784.61
VolumeFlow [m3/hr] 4.287 1.465 63.318 155.464
StdLiqVolumeFlow [m3/hr] 4.445 1.236 12.311 26.263
StdGasVolumeFlow [SCMD] 2.4418E+4 5.685/E+3 8.5285E+3 1.8194E+4
Properties
Molar Composition
Mass Composition
StdLiqVolume Composition
Comp Mole Flow [kgmole/h]
Comp Mass Flow [kg/h]
Comp Std Liq Vol Flow [m3/hr]
ENERGY condenserQ reboilerQ <New>
Stage 1 10
Type EnergyOut EnergyIn
Connected Obj
Value [W] 1039823.2609 12923098.6086
INTERNAL VAP/LIQ <New>
Print
Always Solve
Solve
Restart
Last Conv
Converged
Ignored

Appendix 10


ICON: azza - [/C-1 (DistillationColumn): 10 Stages, Degree of Freedom = 0]
File View Tools Assistants Dynamics Window Help
8 Jamadil Akhir 1430
Ready
/C-1 Inner Error 0.000001
Name: C-1
Description:
Configuration
Spec/Estimates
Efficiencies
Profile
Convergence
Add/RemoveStages...
Schematic...
Specification Required = 4 (4 supplied). Delete 'Name' to remove. Delete 'Value' to turn into viewed spec.
Name Stage Type Associated Draw Detail Connected Obj Unit Value
Reflux 1 RefluxRatio 3.00
MoleFlow 1 PortDataSpec condenserL MoleFlow kgmole/h 3300
MoleFlow 3 PortDataSpec Napthalene MoleFlow kgmole/h 18.00
MoleFlow 8 PortDataSpec HCGO MoleFlow kgmole/h 15.00
Viewed Specifications. Delete 'Name' to remove. Enter a value to turn into an active specification
Name Stage Type Associated Draw Detail Connected Obj Unit Value
<New>
Estimates
Name Stage Type Associated Draw Connected Obj Unit Value
<New>
Stage Pressure
Stage Pressure [kPa]
1 (condenser) 360.0
2 376.667
3 393.333
4 410.0
5 426.667
6 443.333
7 460.0
8 476.667
9 493.333
10 (reboiler) 510.0
Print
Always Solve
Converged
Solve
Restart
Last Conv
Ignored