A REPORT ON Modeling and Simulation of Crude Distillation Unit

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CERTIFICATE

This is to certify that the work contained in this dissertation titled "Modeling & Simulation of Crude Distillation Unit" has been carried out by Abhishek Sonu (R0404205002) under my supervision and has not been submitted elsewhere for a degree.

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Abstract

Simulation of distillation columns usually is carried out using an equilibrium model based on the assumption that thermodynamic and thermal equilibrium exists on each tray in the column. To deal with departures from thermodynamic equilibrium it is common practice to specify an efficiency that is the same for all components on all stages. However, component efficiencies are *not* equal in systems with more than two components and cannot be correlated, especially for nonideal systems. This makes the equilibrium model inappropriate for dynamic simulations where these efficiencies are subject to change and cannot be specified beforehand. Using constant efficiencies neglects the influence of the tray hydrodynamics on the mass transfer and the consequences for the column dynamics. Additionally, in certain column operations there is a departure from thermal equilibrium which cannot be modeled with the equilibrium model either.

In this report equilibrium as well as nonequilibrium model for the dynamic simulation of distillation columns is described. The nonequilibrium model incorporates the rigorous calculation of the mass and energy transfer rates and avoids the use of efficiencies. The influence of mass transfer correlations on column dynamics has been investigated. A new design mode is implemented that eliminates the need for a previously known column layout in order to do a nonequilibrium (dynamic) simulation, which enlarges the range of application of the model.



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Chapter 1 Introduction

To improve the economics, flexibility, operability, and safety of column-based separation processes, design that considers steady-state as well as dynamic behavior is desired. This requires a fundamentally sound model that is capable of accurately describing the separation process. A nonequilibrium model can provide accurate predictions of column performance without the need to fit experimental column data. My objective is to make use of Aspen Hysis which has an improved nonequilibrium model which allows both steady-state and dynamic behavior of columns to be studied.

Need for Dynamic Process Modeling

Current industry design practices feature separate process design and process control groups. Process design remains a largely sequential process with each process unit being sized individually with the emphasis on minimizing capital expenditures instead of an emphasis on overall performance. The resulting dynamic system of the connected equipment can be ill-behaved or difficult to control. Especially designs that minimize holdups may lead to severe control problems. Control engineers often are involved only after the design is already complete. This may lead to redesign of the control system or equipment, unnecessarily complicated control systems, and revenue lost due to delays or not meeting promised product deliveries.

In the process design phase, dynamic simulation is needed to determine dynamic responses to process disturbances such as surge tank levels, column sump levels, product flow rates, or product compositions required for sizing of the relevant equipment as well as for the selection and location of control sensors and actuators. Dynamic modeling can also identify whether the product goals are attainable and detect control schemes which are not adequate or appropriate for meeting given product specification goals. This is specifically of interest with the installation of advanced control algorithms. Initial tuning and optimization of control parameters can result in faster process start-up and control problems will be detected earlier when the process design can be altered without a considerable increase in costs. Controllability of alternative flowsheets can be investigated and auxiliary equipment for startup or shutdown can be located and designed. A side benefit from dynamic simulation is that process engineers will become familiar with the process dynamics and control issues (and control engineers with process design).

For process safety assessments, dynamic simulation can be used to check whether environmental constraints will be met during transitions or to test various emergencies shut down procedures without performing actual experiments. Start-up and shutdown of current processes can be optimized and while interfacing with the control structure dynamic simulation can facilitate online process optimization and operator training. It can also be a tool for testing the controls robustness, e.g., in relation to measurement errors or valve malfunctioning.

Clearly, dynamic simulation provides process and control engineers with a powerful tool to improve process design and production in various ways which can lead to improved revenues. However, dynamic process simulation is only yet starting to become of importance due to the fact that dynamic simulation programs and packages have suffered from a number



of deficiencies. Most engineers could not use them because they were geared for use by specialists; they were neither user-friendly nor interactive. Usually, even the more simple models required large computer resources. Simulators were not portable, fast, flexible, extendable, maintainable, or affordable.

Dynamic Column Simulation

The simulation of separation processes – in particular the simulation of distillation columns is an essential part of dynamic process simulators. Distillation is a high energy consumer in most chemical processes and the interactions between columns can be significant from the design as well as the operability point of view.

Simulation of separation processes by equilibrium stage calculations dates back to 1893 when Sorel published equations for simple, continuous, steady-state distillation. These equations included total and component material balances and a corresponding energy balance that could account for heat losses. Sorel's equations were not widely applied until 1921 when they were used in a graphical solution technique for binary systems by Ponchon (1921) and Savarit (1922), who employed an enthalpy concentration diagram. In 1925 a much simpler, but restricted, graphical technique was developed by McCabe and Thiele. The simplification was achieved by assuming constant molar overflow, eliminating the energy balance equations. Lewis and Matheson (1932) and Thiele and Geddes (1933) were the first to propose methods to solve the systems of equations in a tray by tray manner. Thiele and Geddes (1933) were also the first to solve each type of the MESH equations in turn (MESH is the acronym referring to the different types of equations: M=Material balance, E=Equilibrium, S=Summation, H=Heat balance).

After the introduction of the digital computer in the 1950's, the rate of development of algorithms and simulators has increased dramatically. Equilibrium stage calculations enabled engineers to design a column for the separation of simple mixtures without the need to build a pilot plant again and to scaleup. This had a big impact on the investment costs and design time for a new (part of a) plant. Several textbooks describe the development of the equilibrium stage simulation (see King, 1980; Henley and Seader, 1981; Holland, 1981). Seader (1985) has discussed some recent advances in numerical methods for application to mathematical modeling in process design. By the late 1970's, the steady-state equilibrium stage simulators had made substantial strides, particularly in the area of the physical property prediction. By the end of the 1970's some commercial simulators had gained a wide acceptance by chemical engineers working in process design.

Chemical engineers first used dynamic simulation to evaluate control and safety system designs in the mid 1950's. Only the simplest models could be used, even if just a single unit was to be simulated, since computers were, at that time, far too slow. For the dynamic simulation time derivatives of the holdups on the stages are required making the equations to solve differential ones. A model described by as many as fifty differential equations was then a large model. Early computer models and experiments that appeared in the literature were reported by Mah et al. (1962), Huckaba and coworkers (1963, 1965), Luyben et al. (1964), Waggoner and Holland (1965), Distefano (1968), and Howard (1970). Howard (1970) discussed a continuous distillation simulator and compared results with experiments. Constant molar holdups were assumed, derivative terms in the energy equation were



eliminated and a Runge-Kutta method was used for integration. Boston and Britt (1981) developed a commercial batch distillation simulator, based mainly on the model of Distefano. Gallun and Holland (1982) used Gear's method (1971a, b) to solve the equations involved in dynamic simulation. Holland and Liapis (1983) discuss the use of semi-implicit Runge-Kutta methods as well as the multi-step methods of Gear for the

Integration. Prokopakis and Seider (1983) simulated azeotropic distillation towers.

Gani et al. (1986, 1987a, b, 1989), Cameron (1988), and Ruiz (1988) proposed an extended model for the continuous dynamic simulation of distillation columns. They also discussed the optimization of the dynamic startup/shutdown operations (Gani et al., 1987a, b) and the hydraulics involved. Their model is, perhaps, the most comprehensive dynamic equilibrium stage model described in the literature. They neglected vapor holdup (which is much smaller than the liquid holdup) and assumed the equilibrium model for each stage in the column, using the Murphree plate efficiency. The equations are solved with an ODE solver which solves the algebraic equations through a procedural approach. Gani et al. (1987a, b) discuss startup/shutdown operations and the hydraulics involved which they reported to play a major role in these kinds of simulations. Cuille et al. (1986) simulated batch distillation with chemical reactions present. Holl et al. (1988) made a dynamic simulator called DIVA and Pantelides (1988) included dynamic simulation in SPEEDUP. Gani and Cameron (1989) proposed a general simulator for steady-state as well as dynamic simulation. The dynamic model could even help with steady-state calculations that are very hard to converge.

Several authors discuss the assumptions used in the dynamic simulation of separation columns that introduce errors. Ranzi et al. (1988) discussed the effects of the energy balances and the way they affect the simulation. They found that the energy balances must be evaluated completely in order to predict correct behavior. Choe and Luyben (1987) conclude that vapor holdups cannot be neglected (especially for columns operating at high pressures) and that column pressures should be calculated (especially for low pressure columns, where the tray pressure has a large influence on the tray temperature).

Nonequilibrium column models

Although the equilibrium model has been the basis of the dynamic simulation of distillation columns, its shortcomings are well known. The model is based on the assumption that both phases leaving a stage are in thermal and thermodynamic equilibrium. In practice equilibrium is rarely attained since mass and heat transfer are actually *rate* processes that are driven by the gradients in chemical potential and temperature.

The traditional method of coping with finite rates of mass transfer in stagewise processes has been through the concept of stage efficiency. There are various definitions of stage efficiencies, but the most popular is the Murphree (1925) component vapor efficiency:

$$\xi = \frac{y_{ij} - y_{i,j-1}}{y_{ij}^* - y_{i,j-1}} \tag{1.1}$$



This stage efficiency *reflects* the ratio of actual mass transfer over the mass transfer that would be accomplished by an equilibrium stage. For lack of other information, the stage efficiency is taken to be the same for all components, obtained from some empirical correlation depending on the components in the mixture.

For a binary system both component efficiencies are equal, but unfortunately this is not the case in systems with more than two components. Diffusional interaction phenomena (for example reverse diffusion or osmotic diffusion; have proven that mass transfer can occur against a gradient or in absence of a gradient (Toor, 1964). If a component diffuses against its gradient the component efficiency will be negative (since the direction of mass transfer is the opposite of that what the equilibrium model would predict), and, if it diffuses without a gradient, the components efficiency will be infinite (since the equilibrium model predicts no mass transfer). Because diffusional interactions influence the fluxes differently for each component, component efficiencies in mixtures with three or more components do not have to be equal. In fact they can vary over a range from $-\infty$ to $+\infty$. This surprising result has been confirmed by experiment (Krishna et al., 1977). For ideal and moderately ideal systems the component efficiencies are only a weak function of the composition, in contrast to nonideal systems where the opposite is true. Consequently, in the distillation of nonideal systems the concentration transients could cause large component efficiency changes that might significantly alter the simulation. Therefore, any good model must be based on diffusion calculations that include diffusional interactions. However, dynamic simulators based on the equilibrium model use Murphree efficiencies which are assumed constant and equal for all components.

Efficiencies also depend on the type of operation, as they differ in distillation and absorption operations for the same mixture at hand. Plate hydraulics (including weeping and entrainment) influence the flows on a tray. Different vapor and liquid flows result into various flow regimes of the two phases on the tray (such as spray, emulsion, or bubble flow) which each have there own transfer properties (and thus, efficiencies). Thus, transients in the tray hydraulics imply possible changes in the component efficiencies but with a constant efficiency model such effects are totally neglected.

Another assumption of the equilibrium model, thermal equilibrium, forces the liquid and vapor leaving a stage to have the same temperature. In reality, heat transfer between the two phases is limited and the separate phases have their own temperatures. The assumption of thermal equilibrium makes it difficult to model the dynamics of sections in a column that are purposely used for heat transfer, or columns where feeds are normally subcooled or superheated (such as extractive distillation or strippers/absorbers).

To eliminate the problems discussed above we need to construct a new dynamic column model which does not employ overall thermodynamic and thermal equilibrium assumptions! A nonequilibrium model was developed by Krishnamurthy and Taylor (1985a-d, 1986) (see, also, Sivasubramanian et al., 1987; Powers et al., 1988; Lao et al., 1989, 1994; Taylor and Krishna, 1993) for steady-state simulation of separation processes. The nonequilibrium model splits the stage material and energy balances into balances for each phase, adding rate equations for the calculation of mass and energy interphase transfer rates. The mass transfer rates are computed through matrix routines directly from fundamental diffusion equations and mass transfer correlations. A second generation model was developed by Taylor et al.



(1994) which incorporated the pressure as a variable. Taylor et al. (1992) have demonstrated application of the nonequilibrium model to industrial column operations. Since the nonequilibrium model avoids

The use of tray efficiencies and includes the column hydraulics (which are very important in dynamic column simulation) it is suitable as a basis for developing a better dynamic column model.

As very few unsteady-state column data is available, dynamic simulations of columns or linked columns provide an ideal opportunity to study and analyze the dynamic behavior when no other model is available. However, it also makes it difficult to validate the results of a dynamic simulator other than by checking general trends.

Objectives

The purpose of this work has been to construct and implement dynamic equilibrium and nonequilibrium models into a dynamic column simulator. Requirements of the simulator were:

- Portable implementation
- Easily switch between Steady-State (SS) mode and Dynamic State (DS) mode
- Contain an extensive collection of models for handling the dynamics of many different kinds of trays
- Contain a variety of models for multicomponent diffusion coefficients, mass transfer coefficients, thermodynamic properties, and physical properties. Accurate models for these properties are needed in order to use the nonequilibrium model.
- Numerically robust as well as efficient in terms of computer time and storage
- Easy to use (interactive)
- Flexible and extendible
- Graphical output

The simulator was to be used to investigate influences of different holdup models, tray layout parameters, mass transfer coefficient and diffusion models on open loop simulations. Optional were the inclusion of controllers (closed loop simulations) and the operation outside normal operation to study startup and shutdown operations. Since dynamic experimental measurements are virtually absent, no comparison with data is carried out.



Chapter 2 Steady-State Simulation of Nonequilibrium Columns

The steady-state nonequilibrium model and its equations are introduced. Extra specifications required by the nonequilibrium model in comparison with the equilibrium model are identified. A new *design mode* which enables the simultaneous design of the column layout and column simulation is explained. This design mode enables the use of the nonequilibrium model in flowsheet design calculations

The Nonequilibrium Model

A second generation nonequilibrium model was developed by Taylor and coworkers and is described in detail by Taylor et al. (1994). It can be used to simulate trayed columns as well as packed columns. Packed columns are simulated with stages representing a discrete integration over the packed bed. The more stages are used the better the integration, and the more accurate the results will be. A schematic diagram of a nonequilibrium stage is shown in Figure 2.1. This stage may represent one (or more than one) tray in a trayed column or a section of packing in a packed column. The vertical wavy line in the middle of the diagram represents the interface between the two phases which may be vapor and liquid (distillation), gas and liquid (absorption) or two liquids (extraction).

Figure 2.1 also serves to introduce the notation used in writing down the equations that model the behavior of this nonequilibrium stage. The flow rates of vapor and liquid phases leaving the j-th stage are denoted by Vj and Lj respectively. The mole fractions in these streams are $y_{i,j}$ and $x_{i,j}$. The $N_{i,j}$ are the molar fluxes of species i on stage j. When multiplied by the area available for interphase mass transfer we obtain the rates of interphase mass transfer. The temperatures of the vapor and liquid phases are not assumed to be equal and we must allow for heat transfer as well as mass transfer across the interface.



Figure 2. 1: Schematic diagram of a nonequilibrium stage (Taylor and Krishna, 1993).

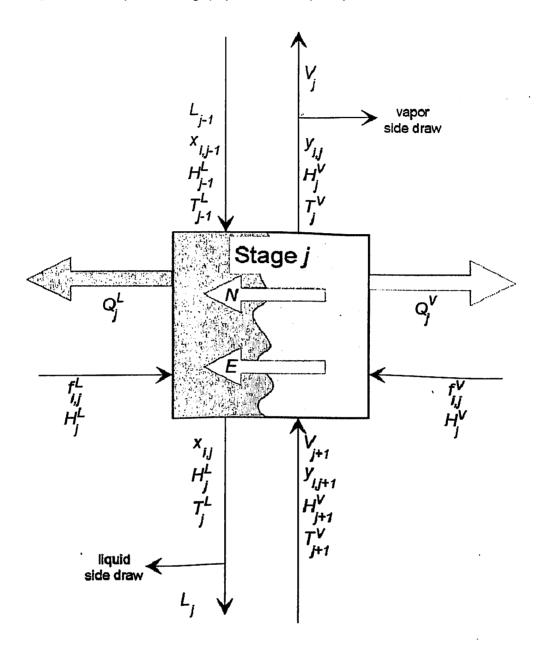


Figure 2.1: Schematic diagram of a nonequilibrium stage (Taylor and Krishna, 1993).

If Figure 2.1 represents a single tray then the term ϕ_j^L is the fractional liquid entrainment defined as the ratio of the moles of liquid entrained in the vapor phase in stage j to the moles of downflowing liquid from stage j. Similarly, ϕ_j^V is the ratio of vapor entrained in the liquid leaving stage j (carried down to the tray below under the downcomer) to the interstage vapor flow. For packed columns, this term represents axial dispersion. Weeping in tray columns



may be accounted for with a similar term. The component (M)aterial balance equations for each phase may be written as follows:

$$M_{ij}^{V} \equiv (1 + r_{j}^{V} + \phi_{j}^{V})V_{j}y_{ij} - V_{j+1}y_{i,j+1} - \phi_{j+1}^{V}V_{j-1}y_{i,j+1} - f_{ij}^{V} - \sum_{\nu=1}^{n} G_{ij\nu}^{V} + N_{ij}$$

$$= 0 \qquad i = 1, 2, \dots, c$$

$$(2.1)$$

$$M_{ij}^{L} \equiv (1 + r_{j}^{L} + \phi_{j}^{L})L_{j}x_{ij} - L_{j-1}x_{i,j-1} - \phi_{j+1}^{L}L_{j+1}x_{i,j+1} - f_{ij}^{L} - \sum_{\nu=1}^{n} G_{ij\nu}^{L} - N_{ij}$$

$$= 0 \qquad i = 1, 2, \dots, c$$
(2.2)

where G_{ijv} is the interlinked flow rate for component i from stage ν to stage j, n is the number of total stages (trays or sections of packing), r_j is the sidestreams flowratio, and f_{ij} is the component feed rate.

The last terms in Equations (2.1) and (2.2) are the mass transfer rates (in kmols/s). At the V/L interface we have continuity of mass and, thus, the mass transfer rates in both phases must be equal. Mass transfer from the "V" phase to the "L" phase is defined as positive.

The total material balances for the two phases are obtained by summing Equations (2.1) and (2.2) over the component index i.

$$M_{t_{j}}^{V} \equiv (1 + r_{j}^{V} + \phi_{j}^{V})V_{j} - V_{j+1} - \phi_{j-1}^{V}V_{j-1} - F_{j}^{V} - \sum_{i=1}^{c} \sum_{\nu=1}^{n} G_{ij\nu}^{V} + N_{tj}$$

$$= 0$$
(2.3)

$$M_{t_{j}}^{L} \equiv (1 + r_{j}^{L} + \phi_{j}^{L})L_{j} - L_{j-1} - \phi_{j+1}^{L}L_{j+1} - F_{j}^{L} - \sum_{i=1}^{c} \sum_{\nu=1}^{n} G_{ij\nu}^{L} - N_{tj}$$

$$= 0$$
(2.4)

 F_j denotes the total feed flow rate for stage j, $F_j = \sum_{i=1}^{c} f_{ij}$.

Here total flow rates and mole fractions are used as independent variables and total as well as component material balances are included in the set of independent model equations. In the nonequilibrium model of Krishnamurthy and Taylor (1985) component flow rates were treated as variables.

The nonequilibrium model uses two sets of (R)ate equations for each stage:

$$R_{ij}^{V} \equiv N_{ij} - N_{ij}^{V} = 0 i = 1, 2 ..., c - 1$$
 (2.5)

$$R_{ij}^{L} \equiv N_{ij} - N_{ij}^{L} = 0 i = 1, 2 ..., c - 1$$
 (2.6)

Where N ij is the mass transfer rate of component i on stage j. The mass transfer rate in each phase is computed from a diffusive and a convective contribution with



$$N_{ij}^{V} = J_{ij}^{V} a_{j}^{I} + y_{ij} N_{tj} (2.7)$$

$$N_{ij}^{L} = J_{ij}^{L} a_{j}^{I} + x_{ij} N_{tj} (2.8)$$

where $a_j^{\ l}$ is the total interfacial area for stage j and N_{ij} is the total rate on stage j. The diffusion fluxes J are given by (in matrix form):

$$(J^{V}) = c_{t}^{V}[k^{V}](\overline{y^{V} - y^{I}})$$
 (2.9)

$$(J^{L}) = c_{t}^{L}[k^{L}](\overline{x^{I} - x^{L}})$$
 (2.10)

The matrices of mass transfer coefficients, [k], are calculated from

$$[k^P] = [R^P]^{-1}[\Gamma^P]$$
 (2.11)

Where Γ^P is a matrix of thermodynamic factors for phase P. For systems where an activity coefficient model is used for the phase equilibrium properties the thermodynamic factor matrix (order c-1) is defined by

$$\Gamma_{ij} = \delta_{ij} + x_i \left(\frac{\partial \ln \gamma_i}{\partial x_j}\right)_{T,P,x_k,k \neq j=1...c-1}$$
(2.12)

If an equation of state is used γi is replaced by φi. Expressions for the composition derivatives of ln γi are given by Taylor and Kooijman (1991). The rate matrix R (order c-1) is a matrix of mass transfer resistances calculated from the following formulae:

$$R_{ii}^{P} = \frac{z_{i}}{k_{ic}^{P}} + \sum_{k=1}^{c} \frac{z_{k}}{k_{ik}^{P}}$$
 (2.13)

$$R_{ij}^{P} = -z_{i} \left(\frac{1}{k_{ij}^{P}} - \frac{1}{k_{ic}^{P}} \right)$$
 (2.14)

where $k_{ij}^{\ P}$ are binary pair mass transfer coefficients for phase P. Mass transfer coefficients, k_{ij} are computed from empirical models (Taylor and Krishna, 1993) and multicomponent diffusion coefficients evaluated from an interpolation formula (Kooijman and Taylor, 1991). Equations (2.13) and (2.14) are suggested by the Maxwell-Stefan equations that describe mass transfer in multicomponent systems (see Taylor and Krishna, 1993). The matrix of thermodynamic factors appears because the fundamental driving force for mass transfer is the chemical potential gradient and not the mole fraction or concentration gradient. This matrix is calculated from an appropriate thermodynamic model.

The binary mass transfer coefficients are estimated from empirical correlations as functions of column internal type as well as design, operational parameters, and physical properties including the binary pair Maxwell-Stefan diffusion coefficients. Thus, the mass transfer coefficient models form the basis of the nonequilibrium model and it is possible to change the



behavior of a column by selecting a different mass transfer coefficient correlation. Table 2.1 gives a summary of the correlations per type of column internals which are currently supported by our steady-state nonequilibrium model; they are described in detail by Taylor and Krishna (1993).

Note that there are c-1 times c-1 binary pair Maxwell-Stefan diffusion coefficients, but only c-1 times c-1 elements in the [R^P] and [k^P] matrices and, therefore, only c-1 rate equations per phase. This is the result of the fact that diffusion calculations only yield relative transfer rates. We will need an extra equation that will "bootstrap" the mass transfer rates: the energy balance for the interface. Note also that, in this model, the flux correction on the mass transfer coefficients has been neglected.

The (E)nergy balance equations on stage j are written for each phase as follows:

$$E_{j}^{V} \equiv (1 + r_{j}^{V} + \phi_{j}^{V})V_{j}H_{j}^{V} - V_{j+1}H_{j+1}^{V} - \phi_{j-1}^{V}V_{j-1}H_{j-1}^{V} - F_{j}^{V}H_{j}^{VF} - \sum_{\nu=1}^{n} G_{j\nu}^{V}H_{j\nu}^{V} + Q_{j}^{V} + e_{j}^{V} = 0$$

$$(2.15)$$

$$E_{j}^{L} \equiv (1 + r_{j}^{L} + \phi_{j}^{L})L_{j}H_{j}^{L} - L_{j-1}H_{j-1}^{L} - \phi_{j+1}^{L}L_{j+1}H_{j+1}^{L} - F_{j}^{L}H_{j}^{LF} - \sum_{\nu=1}^{n} G_{j\nu}^{L}H_{j\nu}^{L} + Q_{j}^{L} - e_{j}^{L} = 0$$
(2.16)

The last term in the left-hand-side of Equations (2.15) and (2.16), e_j, represents the energy transfer rates for the vapor and liquid phase which are defined by

$$e_j^V = a_j^I h^V (T^V - T^I) + \sum_{i=1}^c N_{ij}^V \bar{H}_{ij}^V$$
 (2.17)

$$e_j^L = a_j^I h^L (T^I - T^L) + \sum_{i=1}^c N_{ij}^L \bar{H}_{ij}^L$$
 (2.18)

Where H_{ij} are the partial molar enthalpies of component i for stage j. The continuity of the energy fluxes across the V/L interface which gives the interface energy balance:

$$E_j^I \equiv e_j^V - e_j^L = 0 (2.19)$$

Where H^V and H^L are the vapor and liquid heat transfer coefficients respectively, and $^{I}T^{V}$, T^{I} , and T^{L} the vapor, interface, and liquid temperatures. For the calculation of the vapor heat transfer coefficients the Chilton-Colburn analogy between mass and heat transfer is used:

Le =
$$\frac{\lambda}{DC_p\rho} = \frac{Sc}{Pr}$$
 (2.20)

$$h^V = k_0 C_n L e^{2/3} (2.21)$$



For the calculation of the liquid heat transfer coefficients a penetration model is used:

$$h^L = k\rho C_p \sqrt{\mathsf{Le}} \tag{2.22}$$

Where k is the average mass transfer coefficient and D the average diffusion coefficient.

In the nonequilibrium model of Krishnamurthy and Taylor (1985) the pressure was taken to be specified on all stages, as is normally done in equilibrium model simulations. However, column pressure drop is a function of tray (or packing) type as well as the column design and column operating conditions, information that is required for or available during the solution of the nonequilibrium model equations. It was, therefore, quite straightforward to add a **hydraulic** equation to the set of independent equations for each stage and to make the pressure of each stage (tray or packed section) an unknown variable. The stage is assumed to be at mechanical equilibrium so, $p_j^V = p_j^L = p_j$.

Phase (E)quilibrium is assumed to exist only at the interface with the mole fractions in both phases related by:

$$Q_{ij}^{I} \equiv K_{ij}x_{ij}^{I} - y_{ij}^{I} = 0 \qquad i = 1, 2, ..., c$$
 (2.29)

Where K_{ij} is the equilibrium ratio for component i on stage j. The K_{ij} are evaluated at the (calculated) temperature, pressure, and mole fractions at the interface.

The mole fractions must (S) um to unity in each phase:

$$S_j^V \equiv \sum_{i=1}^c y_{ij} - 1 = 0 (2.30)$$

$$S_j^L \equiv \sum_{i=1}^c x_{ij} - 1 = 0 {(2.31)}$$

as well as at the interface:

$$S_j^{VI} \equiv \sum_{i=1}^c y_{ij}^I - 1 = 0 \tag{2.32}$$

$$S_j^{LI} \equiv \sum_{i=1}^c x_{ij}^I - 1 = 0 \tag{2.33}$$

Table 2.3 lists the type and number of equations for the nonequilibrium model. The model consists of 5c+6 equations and variables, where c is the number of components. The equations are solved simultaneously using inside out method (see appendix).



Nonequilibrium and equilibrium models require many similar specifications. Feed flows and their thermal condition must be specified for both models, as must the column configuration (number of stages, feed and sidestream locations etc.). Additional specifications that are the same for both simulation models include the specification of, for example, reflux ratios or bottom product flow rates if the column is equipped with a condenser and/or a reboiler. The specification of the pressure on each stage is necessary if the pressure drop is not computed; if it is, only the top stage pressure needs be specified (the pressure of all other stages being determined from the pressure drop equations that are part of the model described in the appendix).

Table 2. 1: Nonequilibrium model equations type and number

Table 2.3: Nonequilibrium model equations type and number

Equation	Number
Material balances	2c + 2
Energy balances	3
transfer Rate equations	2c-2
Summations equations	2
Hydraulic equation	1
interface eQuilibrium relations	С
Total MERSHQ	5c + 6

If we solve the nonequilibrium model with Newton's method, we also require initial guesses for all the variables. This is done with an initial guess routine normally used for equilibrium stage simulation which uses a bottoms flowrate and reflux ratio specification and solves the column using the idealWilson K-value model. Temperatures of the vapor, interface, and liquid are then initialized as being equal to the temperature from this guess. Mass and energy transfer rates are initialized as zero and the interface mole fractions are set equal to the bulk mole fractions which are also provided by the initial guess. Pressure drops are initially assumed to be zero.

A nonequilibrium simulation needs the following extra specifications (in comparison with an equilibrium model):

- The column internals type and the layout
- Mass transfer coefficient model
- Flow model for both phases
- Entrainment and weeping models
- Pressure drop model
- Physical properties models



Table 2. 2: Physical Property Needs of Equilibrium and Nonequilibrium Models

Property	EQ model	NEQ model	Used for
K values	Yes	Yes	Driving forces
Enthalpy	Yes	Yes	Energy balances
Activity coefficient	Yes	Yes	K values, enthalpies
Fugacity coefficients	Yes	Yes	K values, enthalpies
Vapor pressure	Yes	Yes	K values
Heat capacity	Yes	Yes	Enthalpies, heat-transfer coefficient
Mass-transfer coefficients		Yes	Mass-transfer rate equations
Heat-transfer coefficients		Yes	Energy-transfer rate equation
Density		Yes	Mass-transfer coefficients
Diffusion coefficients		Yes	Mass-transfer coefficients
Viscosity		Yes	Mass-transfer coefficients
Surface tension		Yes	Mass-transfer coefficients
Thermal conductivity		Yes	Heat-transfer coefficients

For the estimation of transport properties the nonequilibrium model requires the evaluation of many more physical properties (such as densities, viscosities, diffusivities, heat capacities, thermal conductivities, surface tension) which the equilibrium model does not need.

In addition, a nonequilibrium simulation cannot proceed without some knowledge of the column type and the internals layout in order to determine mass transfer coefficients, interfacial area, and pressure drop. Tray type and mechanical layout data, for example, is needed in order to calculate the mass transfer coefficients for each tray. For packed columns the packing type, size and material must be known. Column layout is specified per section of the column, where a section is represented by one or more trays (or packed bed). Standard tray or packing, layout and data can be stored on-line in libraries to be easily accessible. For designers this restriction resulted in first simulating the column with an equilibrium model, rating the column and only then using a nonequilibrium model. If the flows in the column changed - due to different specifications or a change in the feeds - the column had to be recalculated with the equilibrium model and re-rated. Therefore, the nonequilibrium model could not be used in the design of flowsheets, where changes in the flows required a continuous re-rating of the column.

The Design Mode

For each type of internal a specific design mode routine needs to be written that will determine the column layout given a set of flow rates and physical properties on the tray or in the section of packing. However, similarities in tray- and packing-design allow combination for similar types of internals. Liquid-liquid extractors require completely different design methods, even if a similar internals layout is used as in distillation (this is due to the smaller difference in the properties of the contacting phases). Since the layout must be adapted for changes in both flow rates and properties the column layout is input as well (this facilitates the user to make specifications that the design mode will not change as far is possible).



The initial layout is determined after the flows are known from the initial guess. Each stage in the column is designed separately and independently of adjacent stages. Then, after each iteration (that is, an update of the flows) the same design routine is called for re-design. Since the flowrates are also dependent on the layout (to a smaller degree than the layout is dependent on the flow rates) it is important that the design routine is only executed if the flowrates have changed more than by a certain fraction (which can be specified). After convergence has been attained, the internals design is rationalized, making the design for each stage in a column section the same. Then the simulation is restarted with the previously converged answer as starting point. The design method provides a complete design of any trayed or packed section in the column. In this manner trayed and packed sections can be freely mixed in a column simulation/design.

Different design methods can be employed:

- Fraction of flooding; this is the standard design method for trays; we have employed a modified version of the method published by Barnicki and Davis (1989).
- Pressure drop; this is the usual design method for packed columns, but is very useful as well for tray design with pressure drop constraints.

The methods generate a column-design that might not be optimal from an engineer's viewpoint. They must be seen as starting points for the actual design layouts. Also, the design does not include constructional calculations to determine tray support constructions or thicknesses of trays or the column. Design mode is automatically triggered if the column diameter is not specified. Other layout parameters can be specified but they may be changed by the design routine. Each of these methods behaves differently and they are discussed in more detail below. An additional and very important de-rating factor is the system factor (SF). It represents the uncertaincy in design correlations with regard to phenomena which are currently still not properly modelled, such as foaming.



Table 2. 3: Tray layout data

General (sieve) tray layout data:			
Column diameter	Active area		
Number of flow passes	Total hole area		
Tray spacing	Downcomer area		
Liquid flow path length	Weir length		
Hole diameter	Weir height		
Hole pitch	Deck thickness		
Downcomer clearance			
Additional data for bubble caps:			
Cap diameter	Slot area		
Slot height	Riser area		
Skirt clearance	Annual area		
Additional data for valves:			
Closed Loss K	Open Loss K		
Eddy Loss C	Ratio Valve Legs		
Valve Density	Valve Thickness		
Fraction Heavy Valves	tion Heavy Valves Heavy Valve Thickness		

Tray layout parameters that specify a complete design (for the calculation of mass transfer coefficients and pressure drops) are shown in Table 2.4. For packings only the column diameter and bed height are design parameters, other parameters are fixed with the selection of the type of packing (such as void fraction, nominal packing diameter, etc.). The packed bed height must be specified since it determines the desired separation and the capacity.



Chapter 3 Thermodynamic data and model

Reliable thermodynamic data are essential for the accurate design or analysis of distillation columns. Failure of equipment to perform at specified levels is often attributable, at least in part, to the lack of such data. This subsection summarizes and presents examples of phase equilibrium data currently available to the designer. The thermodynamic concepts used are presented in the subsection

Phase equilibrium data

To determine the actual state of a mixture defined by its components and two intensive variables (usually pressure and temperature), a unique set of conditions and equations defining equilibrium is required. Consider a closed, multi-component and multi-phase system whose phases are in thermal, mechanical, and mass transfer equilibrium. The general conditions necessary for thermodynamic equilibrium between heterogeneous phases are established (for all *i*):

$$T^1 = T^2 = ... = T^*$$
 Thermal Equilibrium - no heat flux across phases

$$P^1 = P^2 = ... = P^x$$
 Mechanical Equilibrium - no phase displacement

$${\mu_i}^1={\mu_i}^2=...={\mu_i}^x$$
 Mass Transfer Equilibrium - no mass transfer for component i between phases

For mixtures containing more than two species, an additional degree of freedom is available for each additional component. Thus, for a four component system, the equilibrium vapor and liquid compositions are fixed only if the pressure, temperature, and mole fractions of two components are set. The K values are widely used in multicomponent distillation calculations, and the ratio of the K values of two species, called the relative volatility,

$$\alpha_{ij} = Ki/Kj$$

When using equations of state to represent the vapour and liquid behaviour, you have:

$$f_i^{p} = \phi_i^{p} y_i P$$

$$f_i^L = \phi_i^L x_i P$$



and therefore:

$$K_i = \frac{\phi_i^L}{\phi_i^V}$$

Activity coefficient based models can easily be expressed in this format:

$$f_i^L = \phi_i^L x_i P = \gamma_i x_i f_i^{ref}$$

and therefore:

$$\phi_i^L = \frac{\gamma_i f_i^{\textit{ref}}}{P}$$

So far, the equality of fugacities on the phases for each individual component has been used as the criteria for phase equilibria. Although the equality of fugacities is a necessary criterion, it is not sufficient to ensure that the system is at equilibrium. A necessary and sufficient criterion for thermodynamic equilibrium is that the fugacities of the individual components are the same and the Gibbs Free Energy of the system is at its minimum.

Mathematically:

$$f_i^I = f_i^{II} = f_i^{III} \dots$$

and

Gsystem = minimum.

Thermodynamic Calculation Model

Peng-Robinson Equation of State: The Peng Robinson (1976) equation of state (EOS) is a modification of the RK equation to better represent VLE calculations. The densities for the liquid phase in the SRK did not accurately represent the experimental values due to a high universal critical compressibility factor of 0.3333.

The PR is a modification of the RK equation of state which corresponds to a lower critical compressibility of about 0.307 thus representing the VLE of natural gas systems accurately. The PR equation is represented by:



$$P = \frac{RT}{V-b} - \frac{\alpha}{V(V+b) + b(V-b)}$$

where:

$$a = a_c \alpha$$

$$a_c = 0.45724 \frac{R^2 T_c^2}{P_c}$$

$$b = 0.077480 \frac{RT_c}{P_c}$$

The functional dependency of the "a" term is shown in the following relation.

$$\sqrt{\alpha} = 1 + \kappa (1 - T_r^{0.5})$$

$$\kappa = 0.37464 + 1.5422 \omega - 0.26992 \omega^2$$

The accuracy of the PR and SRK equations of state are approximately the same. However, the PR EOS represents the density of the liquid phase more accurately due to the lower critical compressibility factor.

These equations were originally developed for pure components. To apply the PR EOS to mixtures, mixing rules are required for the "a" and "b" terms.

A quick reference of calculation methods is shown in the table below for the HysysPR EOS.

Table 3.1

Calculation Method	Applicable Phase	Property Class Name
Z Factor	Vapour and Liquid	COTH_HYSYS_ZFactor Class
Molar Volume	Vapour and Liquid	COTH_HYSYS_Volume Class
Enthalpy	Vapour and Liquid	COTH_HYSYS_PREnthalpy Class
Entropy	Vapour and Liquid	COTH_HYSYS_Entropy Class
Isobaric heat capacity	Vapour and Liquid	COTH_HYSYS_Cp Class
Fugacity coefficient calculation	Vapour and Liquid	COTH_HYSYS_LnFugacityCoeff Class
Fugacity calculation	Vapour and Liquid	COTH_HYSYS_LnFugacity Class
Isochoric heat capacity	Vapour and Liquid	COTH_HYSYS_Cv Class

The compressibility factor, Z, is calculated as the root for the following equation:



$$Z^{3} - (1 - B)Z^{2} + Z(A - 3B^{2} - 2B) - (AB - B^{2} - B^{3}) = 0$$

$$A = \frac{aP}{R^{2}T^{2}}$$

$$B = \frac{bP}{RT}$$

The following relation calculates the fugacity coefficient.

$$\ln \phi_i = -\ln(V - b) + \frac{\bar{b}}{V - b} + \frac{a}{2\sqrt{2}b} \ln \left(\frac{V + b(1 + \sqrt{2})}{V + b(1 - \sqrt{2})} \right) \left(-1 + \frac{\bar{a}}{a} + \frac{\bar{b}}{b} \right)$$
$$\bar{a} = \frac{\partial n^2 a}{\partial n}$$
$$\bar{b} = \frac{\partial nb}{\partial n}$$



Chapter 4 Dynamic Model Development

Dynamic nonequilibrium models for tray columns are developed. A full model with four holdup terms describing both froth and downcomer, as well as a two holdup model with only froth holdup terms, are introduced. The nonequilibrium models will be compared to corresponding conventional equilibrium models which are also described in this chapter. Finally, issues concerning the implementation and the integration of dynamic column models are discussed.

Nonequilibrium Model Assumptions

A schematic diagram of a general tray in a column is provided in Figure 4.1. Of central importance is the zone where vapor and liquid phases are brought into contact with each other in order to promote mass and energy transfer between the phases. A tray can operate in different flow regimes: spray, froth, emulsion, bubbling liquid, or foam. Here we will generally refer to the dispersion on the tray as the froth, although we do not limit our model to that regime. Above the froth is an area for vapor disengagement, to separate the phases to let them move countercurrently in the column. Similarly we have a downcomer for liquid disengagement. These disengagement areas are essential to the operation of a trayed column and certainly play a role in its performance. What differentiates the dynamic model from the steady-state model (as described by Taylor et al., 1994) is the use of holdup terms. For steady-state simulation holdup calculations are not required, however, in the dynamic model they represent the basic differential equations. For the general tray a number of distinct holdups can be identified:

- the liquid in the froth on a tray,
- the vapor dispersed in the froth on a tray,
- the liquid in the downcomer below a tray,
- the vapor above the froth/downcomer on a tray.



Figure 4. 1: Schematic diagram of a general tray

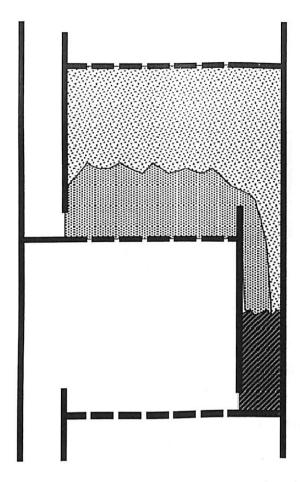


Figure 4.1: Schematic diagram of a general tray.

The froth is modelled by two (or more, if multiple liquid phases are present) separate holdups. Figure 4.2 is a schematic diagram of these holdups and also shows the connecting flows between the different holdups. The following assumptions have been made in our dynamic nonequilibrium model:

The trays are in mechanical equilibrium

• Thermodynamic equilibrium is assumed **only** at the interface between vapor and liquid phases on the tray. This is standard practice in the analysis of interphase mass transfer processes.

Mass transfer occurs only between vapor and liquid on the tray, dictated by the

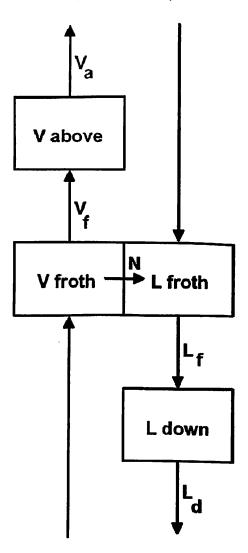
transfer resistance in each phase

Condenser and reboiler operate at equilibrium.

The dynamic model developed here uses all four holdups terms and avoids simplifications often made in other dynamic models such as constant holdups, neglecting energy derivatives, neglecting vapor



Figure 4. 2: Schematic diagram of the holdups and connecting flows.



holdups, and constant (tray/component) efficiencies. To reduce the number of model equations the holdup terms for the vapor above the froth and in the downcomer can be lumped into the froth holdups or ignored (if it is desired to do so).

Nonequilibrium Model Equations

Component molar holdup terms are denoted with U_{ij}^{Pw} where P indicates the holdup phase type (V or L), w the place in the model (f for froth, d for downcomer, and a for above the froth), i the component, and j the plate number. Similarly, total molar holdups are denoted with U_{tj}^{Pw} and energy holdups with E_j^{Pw} .



Vapor and liquid holdup compositions are computed from

$$y_{ij}^w = \frac{U_{ij}^{Vw}}{U_{ij}^{Vw}} \tag{4.1}$$

$$x_{ij}^{w} = \frac{U_{ij}^{Lw}}{U_{ij}^{Lw}} \tag{4.2}$$

The interstage liquid and vapor flows on plate j are denoted with $L_j^{\ W}$ and $V_j^{\ W}$ where w indicates the holdup from which the flows originate (f, d, or a). Component molar feed flows are denoted similarly to molar holdups as $F_{ij}^{\ PW}$. The mass transfer rates through the interface are positive from vapor to liquid and denoted by $N_{ij}^{\ W}$.

For a general stage (one not at the top or bottom of the column) the component molar balances over the four different holdups are:

$$\frac{dU_{ij}^{Vf}}{dt} = y_{i,j+1}^a V_{j+1}^a + F_{ij}^{Vf} - y_{ij}^f V_j^f - N_{ij}^f$$
 (4.3)

$$\frac{dU_{ij}^{Lf}}{dt} = x_{i,j-1}^d L_{j-1}^d + F_{ij}^{Lf} - x_{ij}^f L_j^f + N_{ij}^f$$
 (4.4)

$$\frac{dU_{ij}^{Va}}{dt} = y_{ij}^{f} V_{j}^{f} + F_{ij}^{Va} - y_{ij}^{a} V_{j}^{a}$$
 (4.5)

$$\frac{dU_{ij}^{Ld}}{dt} = x_{ij}^{f} L_{j}^{f} + F_{ij}^{Ld} - x_{ij}^{d} L_{j}^{d}$$
 (4.6)

The component molar holdups must sum to the total molar holdups:

$$0 = \sum_{i=1}^{c} U_{ij}^{Vf} - U_{ij}^{Vf}$$
 (4.7)

$$0 = \sum_{i=1}^{c} U_{ij}^{Lf} - U_{ij}^{Lf}$$
 (4.8)

$$0 = \sum_{i=1}^{c} U_{ij}^{Va} - U_{ij}^{Va}$$
 (4.9)

$$0 = \sum_{i=1}^{c} U_{ij}^{Ld} - U_{ij}^{Ld}$$
 (4.10)

The energy balances for each holdup are:

$$\frac{dE_{j}^{Vf}}{dt} = \frac{E_{j+1}^{Va}}{U_{t,j+1}^{Va}} V_{j+1}^{c} + \sum_{i=1}^{c} H_{ij}^{FVf} F_{ij}^{Vf} - \frac{E_{j}^{Vf}}{U_{tj}^{Vf}} V_{j}^{f} - \epsilon_{j}^{Vf} + Q_{j}^{Vf}$$
(4.11)



$$\frac{dE_{j}^{Lf}}{dt} = \frac{E_{j-1}^{Lf}}{U_{Lj-1}^{Lg}} L_{j-1}^{d} + \sum_{i=1}^{c} H_{ij}^{FLf} F_{ij}^{Lf} = \frac{E_{j}^{Lf}}{U_{ij}^{Lf}} L_{j}^{f} + \epsilon_{j}^{Lf} + Q_{j}^{Lf}$$
(4.12)

$$\frac{dE_{j}^{V,a}}{dt} = \frac{E_{j}^{V,f}}{U_{ij}^{V,f}} V_{j}^{f} + \sum_{i=1}^{c} H_{ij}^{FV,a} F_{ij}^{V,a} - \frac{E_{j}^{V,a}}{U_{i}^{V,a}} V_{i}^{x} + Q_{j}^{V,a}$$
(4.13)

$$\frac{dE_{j}^{Ld}}{dt} = \frac{U_{j}^{Lf}}{U_{ij}^{Lf}}L_{j}^{f} + \sum_{i=1}^{r} H_{ij}^{FLd} F_{ij}^{Ld} + \frac{E_{j}^{Ld}}{U_{ij}^{Ld}} L_{j}^{d} + Q_{j}^{Ld}$$
(4.14)

Where \square is the energy transport to/from the interface (see below). $H_{ij}^{\ FPw}$ is the partial molar enthalpy of component i in the feed to the specified holdup and Q_j^{Pw} is the heat input into the specified holdup. The energy holdups E_j^{Pw} are related to the component molar holdups and the component enthalpies $(H_{ij}^{\ Pw})$ by

$$0 = \sum_{i=1}^{r} (H_{ij}^{Vf} U_{ij}^{Vf}) - E_{j}^{Vf}$$
 (4.15)

$$0 = \sum_{i=1}^{c} (H_{ij}^{Lf} U_{ij}^{Lf}) - E_{j}^{Lf}$$
 (4.16)

$$0 = \sum_{i=1}^{r} (H_{ij}^{Va} U_{ij}^{Va}) - E_{j}^{Va}$$
 (4.17)

$$0 = \sum_{i=1}^{r} (H_{ij}^{Ld} U_{ij}^{Ld}) - E_{j}^{Ld}$$
 (4.18)

Enthalpies are functions of the holdup temperature, T_j^{Pw} , pressure, pj, and holdup molar compositions. The energy fluxes from the vapor to the interface and from the interface to the liquid on plate are:

$$\epsilon_{j}^{Vf} = \sum_{i=1}^{l} N_{ij}^{Vf} H_{ij}^{Vf} + h_{j}^{Vf} a_{j} (T_{j}^{Vf} - T_{j}^{I})$$
 (4.19)

$$\epsilon_{j}^{Lf} = \sum_{i=1}^{t} N_{ij}^{Lf} H_{ij}^{Lf} - h_{j}^{Lf} a_{j} (T_{j}^{I} - T_{j}^{Lf})$$
 (4.20)

Where $T_j^{\ I}$ is the temperature of the interface on plate j. The energy balance over the interface equates these energy fluxes:

$$0 = \epsilon_i^{Vf} - \epsilon_i^{Lf} \tag{4.21}$$

The interface compositions x_{ij}^{lf} and $y_{ij}^{lf must}$ sum to unity,

$$0 = \sum_{i=1}^{n} y_{ij}^{IJ} - 1 \tag{4.22}$$

$$0 = \sum_{i=1}^{c} x_{ij}^{II} - 1 \tag{4.23}$$



and obey the equilibrium relations (i = 1......c) as well:

$$0 = K_{ij}^{If} x_{ij}^{If} - y_{ij}^{If} \tag{4.24}$$

The mass transfer rates N_{ij} from the vapor to the interface are equal to the mass transfer rates from the interface to the liquid. They are computed with the following rate equations:

$$(0) = (N_j) - N_{tj}(y_j^{Vf}) - c_j^{Vf} a_j [R_j^{Vf}]^{-1} [\Gamma_j^{Vf}] ((y_j^{Vf}) - (y_j^{If}))$$

$$(4.25)$$

$$(0) = (N_j) - N_{tj}(x_j^{Lf}) - c_j^{Lf} a_j [R_j^{Lf}]^{-1} [\Gamma_j^{Lf}] ((x_j^{If}) - (x_j^{Lf}))$$
(4.26)

Where N_{tj} is the total mass transfer rate on plate j which equals to the sum of all the component mass transfer rates N_{ij} . c^{vf} and c^{Lf} are the molar concentrations of the vapor and liquid phase of the froth. Note that only c-1 fluxes, with c being the number of components, are independent and we will obtain 2(c-1) rate equations. Also note that the rate equations are in matrix/vector form. The rate matrix Γ is defined by (2.13, 2.14) and the thermodynamic factor matrix by (2.12).

The pressure, p_j , is computed from the tray pressure drop and the pressure of the tray above. The

pressure at the top of the column is specified (p_{spec}):

$$0 = p_1 - p_{spec} \tag{4.27}$$

$$0 = p_j - p_{j-1} - \Delta p_{j-1} \tag{4.28}$$

The interholdup flow rates are determined through calculation of the total molar holdups. The total molar holdups can be computed from the height of the froth, h^f_j , the clear liquid height, h^{cl}_j , the tray spacing, h^{ts}_j , and the liquid height in the downcomer, h^d_j , of plate j:

$$0 = (h_j^f - h_j^{cl}) A_j^f c_j^{Vf} - U_{tj}^{Vf}$$
 (4.29)

$$0 = h_j^{cl} A_j^f c_j^{Lf} - U_{tj}^{Lf}$$
 (4.30)

$$0 = \{(h_j^{ts} - h_j^f)A_j^b + (2h_j^{ts} - h_j^d)A_j^d\}c_j^{Va} - U_{tj}^{Va}$$
(4.31)

$$0 = h_j^d A_j^d c_j^{Ld} - U_{tj}^{Ld} (4.32)$$

The liquid heights are computed by empirical correlations or theoretical relations (see Appendix A). Note that each total holdup must be a function of the relevant flow rate (e.g., U_{ij}^{Ld} should be a function of, L_j^d etc.) to prevent higher index systems (more on this topic below). Since this is not the case for (4.31) we can replace it with

$$0 = V_j^f - V_j^a \tag{4.33}$$

To use a constant molar vapor holdup above the froth (usually the change in $U^{Va}_{\ \ ij}$ is small). This assumption maintains the index of the system at one (instead of two) but violates the physical constraint of a fixed volume between the trays. For the correct dynamic simulation it



is important that the liquid height correlations behave correctly besides being accurate (which is not required for steady-state simulation).

Alternatively, instead of computing the total holdups from liquid heights, which are computed from empirical correlations, the liquid and vapor flows could be computed directly from empirical or theoretical relations. Models describing the holdup or flow need to be accurate and have the correct behavior. Since the number of models which describe liquid heights and froth densities is much larger in comparison to models describing the flows, holdups are being calculated.

The variables and equations, as well as their number, type, and association, are summarized in Tables 4.1. The association is non-trivial as it determines the index of the resulting system of equations. Each variable must be represented (either directly or indirectly) in its associated equation to prevent the generation of a higher index DAE system. That is why the flow variables are matched up with the calculation of the molar holdups from the liquid heights (or directly from the calculation of the flows if that alternative is chosen). That is also why the mass transfer rates are somewhat strangely paired with the equilibrium (4.24), mass transfer rate (4.25, 4.26), and interface composition summation (4.22, 4.23) equations (together with the interface compositions).

The total number of equations is 7c+18 per general stage where c represents the number of components in the system. Out of these, 4c+4 equations are ordinary differential equations while the rest are algebraic equations. The feed flows, heat inputs, top and condenser pressures and product streams are functions of time. If they are constants we are solving a steady-state

(SS) process, where all differential terms are set to zero. If they change over time we switch to dynamic simulation (DS) where we solve the resulting Differential-Algebraic system of equations until steadystate is reached (or until the variable changes are less then some specified small fraction). Of course, only during a steady-state simulation can we activate the design-mode which simultaneously corrects the column design to handle the process flows at hand. The resulting design can then be directly used for the dynamic simulation.

A simplification of this full tray model results from ignoring the vapor above the froth and the liquid in the downcomer. Equations 4.5, 4.6, 4.9, 4.10, 4.13, 4.14, 4.17, 4.18, 4.31, 4.32 are omitted from this model which has variables. The neglected downcomer and vapor holdup could be optionally lumped into the liquid and vapor holdup equations (4.29, 4.30). As mentioned before, the simplified model can (optionally) lump the downcomer liquid and/or vapor above the froth with the liquid or vapor froth holdup, respectively. Also, all holdups can be calculated at steady-state and kept constant during the dynamic simulation. However, lumping holdups, or keeping them constant, are not good representations of the real behavior of trays.

Reboiler and Condenser

Distillation columns also have various types of condensers and reboilers that usually have a significantly larger holdup than the holdup on any tray to give the column operational stability. It is these larger holdups that lead to differences in the transient behavior of various



variables and, therefore, have a large effect on the column behavior. They also cause the system of equations to be very stiff.

Table 4. 1: Variables and equations for the full dynamic model

Variable(s)	Equation(s)	Number	Differential
$[U_{ij}^{Vf}, U_{ij}^{Lf}, U_{ij}^{Va}, U_{ij}^{Ld}]$	(4.3), (4.4), (4.5), (4.6)	4c	+
$\left(E_j^{Vf}, E_j^{Lf}, E_j^{Va}, E_j^{Ld}\right)$	(4.11), (4.12), (4.13), (4.14)	4	+
$U_{tj}^{Vf}, U_{tj}^{Lf}, U_{tj}^{Va}, U_{tj}^{Ld}$	(4.7), (4.8), (4.9), (4.10)	4	
$ig ig V^f_j,L^f_j,V^a_j,L^d_j$	(4.29), (4.30), (4.31), (4.32)	4	
$T_j^{Vf}, T_j^{Lf}, T_j^{Va}, T_j^{Ld}$	(4.15), (4.16), (4.17), (4.18)	4	
T_j^T	(4.21)	1	-
N_{ij}	(4.25), (4.26)	с	
x_{ij}^{If}	(4.26), (4.23), (4.24)	c	·
y_{ij}^{If}	(4.24), (4.22)	c	
$\stackrel{\sim}{P_J}$	(4.27), (4.28)	1	

The reboiler is modelled as a liquid holdup in the bottom of the column followed by a partial (equilibrium) reboiler. The holdup component molar balances (c) are:

$$\frac{dU_{ib}^L}{dt} = x_{in}L_n^a - x_{ib}L_b \tag{4.34}$$

Where the liquid mole fraction is computed by:

$$x_{ib} = \frac{U_{ib}^{L}}{U_{ib}^{L}} \tag{4.35}$$

and the tray above the reboiler is tray n. The total holdup is computed by summing the component holdups:

$$0 = \sum_{i=1}^{c} U_{ib}^{L} - U_{tb}^{L} \tag{4.36}$$

Assuming a constant molar holdup we write the total molar balance (for dynamic state):

$$0 = L_n^d - L_b (4.37)$$

We could also assume a more realistic constant volumetric holdup for the reboiler. At steady-state this equation is replaced by a direct specification of the molar (or possibly volumetric) holdup in the reboiler:



$$0 = U_{tb}^L - U_{tb,spec}^L (4.38)$$

The energy holdup and energy relation are:

$$\frac{dE_b}{dt} = \frac{E_n^{Ld}}{U_n^{Ld}} L_n^d - \frac{E_b}{U_{tb}^L} L_b \tag{4.39}$$

Equilibrium models

The nonequilibrium models developed above will be compared with the corresponding results from two equilibrium models which use specified tray efficiency, (assumed constant over the integration interval). The first model (EQL) neglects the vapor holdup and only the liquid holdup in the froth is included. This holdup can be computed by (4.30) or held constant (computed at steady-state or user specified). The set of equations for this model are:

$$\frac{dU_{ij}}{dt} = y_{i,j+1}V_{j+1} + x_{i,j-1}L_{j-1} + F_{ij} - y_{ij}V_j - x_{ij}L_j$$
(4.62)

Table 4. 2: Variables and equations for the EQL model

Variable(s)	Equation(s)	Number	Differential
U_{ij}	(4.62)	c	+
U_{ij}	(4.63)	1	_
E_{i}	(4.65)	, 1	+
y_{ij}	(4.67)	С	- .
V_{j}	(4.68)	1	
L_{i}^{j}	(4.64)	1	_
T_{i}^{j}	(4.66)	1	_
P_{j}^{\prime}	(4.69)	1	



$$0 = \sum_{j=1}^{c} U_{ij} - U_{tj} \tag{4.63}$$

$$0 = h_j^{el} A_j^f c_j^{Lf} - U_{lj} (4.64)$$

$$\frac{dE_{j}}{dt} = (\sum_{i=1}^{c} g_{i,j+1} H_{i,j+1}^{V}) V_{j+1} + (\frac{E_{j-1}}{U_{i,j-1}}) L_{j-1} + \sum_{i=1}^{c} H_{ij}^{T} F_{ij} + Q_{j}$$

$$-(\sum_{i=1}^{c} y_{ij} H_{ij}^{V}) V_{j} - (\frac{E_{j}}{U_{j}}) L_{j}$$
(4.65)

$$0 = \sum_{i=1}^{c} H_{ij}^{L} U_{ij} - E_{ij}$$
 (4.66)

$$0 = \xi_j K_{ij} x_{ij} - y_{ij} + (1 - \xi_j) y_{i,j+1}$$
 (4.67)

$$0 = \sum_{i=1}^{c} y_{ij} - 1 \tag{4.68}$$

$$0 = p_j - p_{j-1} - \Delta P_{j-1} \tag{4.69}$$

Where we have c+1 differential equations. The tray pressures are computed with the tray pressure drops (the pressure of the tray at the top of the column specified). Table 4.4 lists the 2c+6 equations and variables.

If the vapor holdup is not neglected we obtain the following set of equations (model EQLV):

$$\frac{dU_{ij}^{V}}{dt} + \frac{dU_{ij}^{L}}{dt} = y_{i,j+1}V_{j+1} + x_{i,j-1}L_{j-1} + F_{ij} - y_{ij}V_{j} - x_{ij}L_{j}$$
(4.70)

$$0 = \sum_{i=1}^{c} U_{ij}^{V} - U_{tj}^{V}$$
 (4.71)

$$0 = \sum_{i=1}^{c} U_{ij}^{L} - U_{tj}^{L}$$
 (4.72)



$$0 = (h_j^f - h_j^{el})A_j^f c_j^{Vf} - U_{lj}^V$$
(4.73)

$$0 = h_j^{cl} A_j^f c_j^{Lf} - U_{ij}^L (4.74)$$

$$\frac{dE_{j}^{V}}{dt} + \frac{dE_{j}^{L}}{dt} = \left(\frac{E_{j+1}^{V}}{U_{i,j+1}^{V}}\right)V_{j+1} + \left(\frac{E_{j-1}^{L}}{U_{i,j-1}^{L}}\right)L_{j-1} + \sum_{i=1}^{c} H_{ij}^{T}F_{ij} + Q_{j}$$

$$E^{V} = E^{L}$$

$$-(\frac{E_{j}^{V}}{U_{j}^{V}})V_{j} - (\frac{E_{j}^{L}}{U_{j}^{L}})L_{j}$$
(4.75)

$$0 = \sum_{i=1}^{c} H_{ij}^{V} U_{ij}^{V} - E_{ij}^{V}$$
(4.76)

$$0 = \sum_{i=1}^{c} H_{ij}^{L} U_{ij}^{L} - E_{ij}^{L}$$
(4.77)

$$0 = \xi_j K_{ij} x_{ij} - y_{ij} + (1 - \xi_j) y_{i,j+1}$$
 (4.78)

$$0 = p_j - p_{j+1} - \Delta P_{j-1} \tag{4.79}$$

(4.80)

with c+1 differential equations and a total of 2c+8 equations and variables. Again, neglected holdups could be lumped in as discussed previously.

Physical Property Models

So far only mathematical equations of the dynamic model have been discussed. However, the successful implementation of a column solver requires more than "just solving" the equations. A large and important part of a process simulator are the physical property models which supply the K-values, activity coefficients, binary diffusivities, densities, heat capacities, enthalpies, vapor pressures, viscosities, thermal conductivities, surface tensions, and binary mass transfer coefficients. A nonequilibrium model has a much higher demand for properties compared to an equilibrium model (Taylor et al., 1994). Property models also impose a problem specially associated with dynamic simulation. Often, different correlations are used over different state variable ranges. When a switch between different correlations occurs due to a change in a state variable (such as temperature, pressure, or composition) it causes a discontinuity in the simulation. For the sake of consistency, properties need to be continuous and differentiable at

or around any switching points. Depending on the solver used, proper handling of these discontinuities may require the physical property model/correlation switches to be signaled in some way. However, this is not (yet) done in the present implementation of the models described above. Rather, discontinuities of this kind are avoided as much as possible by using a single correlation for the whole integration. This has implications for the manner in which the model equations can be integrated.



Chapter 5 Simulation Result of Petroleum Mixture Distillation

This chapter contains simulation results of crude distillation unit which is developed with the help of Aspen Hysys software. UNIFAC and Antoine correlations were used to model the liquid thermodynamics and the Peng-Robinson equation of state for the vapor, including the excess enthalpies.

Although the principles of multicomponent distillation apply to petroleum, synthetic crude oil, and other complex mixtures, this subject warrants special consideration for the following reasons:

- 1. Such feedstocks are of exceedingly complex composition, consisting of, in the case of petroleum, many different types of hydrocarbons and perhaps of inorganic and other organic compounds. The number of carbon atoms in the components may range from 1 to more than 50, so that the compounds may exhibit atmospheric-pressure boiling points from -162°C (-259°F) to more than 538°C (1000°F). In a given boiling range, the number of different compounds that exhibit only small differences in volatility multiplies rapidly with increasing boiling point. For example, 16 of the 18 octane isomers boil within a range of only 12°C (22°F).
- 2. Products from the distillation of complex mixtures are in themselves complex mixtures. The character and yields of these products vary widely, depending upon the source of the feedstock. Even crude oils from the same locality may exhibit marked variations
- 3. The scale of petroleum-distillation operations is generally large, and as discussed in detail by Nelson (Petroleum Refinery Engineering, 4th ed., McGraw-Hill, New York, 1958) and Watkins (Petroleum Refinery Distillation, 2d ed., Gulf, Houston, 1979), such operations are common in several petroleum refinery processes including atmospheric distillation of crude oil, vacuum distillation of bottoms residuum obtained from atmospheric distillation, main fractionation of gaseous effluent from catalytic cracking of various petroleum fractions, and main fractionation of effluent from thermal coking of various petroleum fractions. These distillation operations are conducted in large pieces of equipment that can consume large quantities of energy. Therefore, optimization of design and operation is very important and frequently leads to a relatively complex equipment configuration.

Characterization of the Crude in Hypo-components

Due to the variety of molecules that compose petroleum, it is necessary to characterize it as a mixture of a limited number of pseudo-components; they are similar to the different fractions of crude and they have physical properties that represent an average of the properties of various components. In order to convert a distillation curve TBP into a series of pseudo-



components, it is necessary to define a component that has a lower boiling temperature than the initial boiling point of the curve and a higher boiling temperature than the final boiling point. Afterwards, this region is divided in equal areas with horizontal lines, according to the number of pseudo-components that want to be produced; the percentage of the weight of each pseudo-component is established by tracing vertical lines to the boiling point of each in a way that the areas that make up the vertical line with the temperature of the superior and the inferior pseudo-component are equal.

In general, TBP distillations are conducted in columns with 15 to 100 theoretical stages at reflux ratios of 5 or greater. Thus, the new ASTM D 2892 test method, which involves a column with 14 to 17 theoretical stages and a reflux ratio of 5, essentially meets the minimum requirements. Distillate may be collected at a constant or a variable rate. Operation may be at 101.3-kPa (760-torr) pressure or at a vacuum at the top of the column as low as 0.067 kPa (0.5 torr) for high-boiling fractions, with 1.3 kPa (10 torr) being common. Results from vacuum operation are extrapolated to 101.3 kPa (760 torr) by the vapor-pressure correlation of Maxwell and Bonner [Ind. Eng. Chem., 49, 1187 (1957)], which is given in great detail in the API Technical Data Book—Petroleum Refining (op. cit.) and in the ASTM D 2892 test method. It includes a correction for the nature of the sample (paraffin, olefin, naphthene, and aromatic content) in terms of the UOP characterization factor, UOP-K, as given by

$$\frac{\text{IJOP-K} = (T_0)^{1/3}}{\text{SG}}$$

Where TB is the mean average boiling point in degrees Rankine, which is the arithmetic average of the molal average boiling point and the cubic volumetric average boiling point. Values of UOP-K for n-hexane, 1-hexene, cyclohexene, and benzene are 12.82, 12.49, 10.99, and 9.73, respectively. Thus, paraffins with their lower values of specific gravity tend to have high values, and aromatics tend to have low values of UOP-K.

The assay contains all the petroleum laboratory data, boiling point curves, light ends, property curves and bulk properties. Aspen HYSYS uses the supplied assay data to generate internal TBP, molecular weight, density and viscosity curves.

The cut/blend characterization in Aspen HYSYS splits the internal working curves for one or more assay into hypo-components. The blend tab of the oil characterization view provides two functions, cutting oil into hypo-components and blending two or more assays into one set of hypo-components.



Table 5. 1: Crude assay

	· · · · · · · · · · · · · · · · · · ·		Case N	lame: C1P	ram Files V typrolech V HYS	'S 3.1.3\Samplestd	vncnuto3 bec
2		TEAM LND				· · · · · · · · · · · · · · · · · · ·	7/10/00/01/30
4	HYPROTECH	Calgary, Alberta CANADA	Unit Se	1,000	SGPM		<u> </u>
5			Date/Ti	me: Sun Ma	y 03 02:57:56 2009		
7		Assay: As	eav_1				
8		Assay. As					
9			INPU	DATA			
10 11			Data Type:	ТВР			
12							
13 14	Assay Basis:		Liquid Volume	TBP Distillation			Atmospheric
15			tht Ends Status				
16 17	Percentage of Light Ends in A	ssay:	1.				Liquid Volume %
18			Light Ends Co				
19 20	nd Landbert Magnes	Light Ends	Methane	The Control of the	Compos	ifion	ত প্রক্রি ন ্ত্
21			Ethene			*****	6.500e-003 ° 2.250e-002 °
22			Propane				0.3200 •
23 24	grand to all sections — contradictions all sections and contradictions. The sections are contradictions and contradictions are contradictions.		I-Butane n-Butane			Contract to the section of the secti	0.2400 *
25			H2O				0.8200 ° 0.0000 °
26	•	E	uik Properties:	Used			
27 28	Molecular Weight			Viscosity1 Temp	perature	(F)	100.0 *
29	Mass Density	(Ib/ft3)	54.82	Viscosity1		(cP)	
30	Watson Uopk			Viscosity2 Temp	perature	(P)	210.0 *
31 32			Dietillat	ion Table		<u> </u>	
33	· · · · · · · · · · · · · · · · · · ·	Assay Percent		ion rable	Temperati	-	
34 35	North House Co. (Robert Month)	Assay Percent	0.		· · · · · · · · · · · · · · · · · · ·		15.00 *
38	The second secon		5				90.00
37	The second secon		9151				165.0
38		· · · · · · · · · · · · · · · · · · ·	20				240.0 °
39 40	and the second s	The second secon	30 '	•			435.0 *
41			40 °			Paradole or rate discours as a sign	524.0 °
42		THE RESERVE OF THE PERSON OF T	60 '		•		620.0 ° 740.0 °
씜	an i salamana kuningga panggangan na nagaru nahi saningga na di maya		70				885.0
4			76 '		***************************************	-	969.0 *
48	The state of the s		80 9	ļ	and the same of th		1015 *
47			Input Da	ata Status			1,555
48 49			Density:			/iscosity: Not Us	ad
50	Molecular Weight: Not	Used		G CURVES			
51 52		V-101		zaczowanie:	Mary Charles (1977)	Viscosity 1)	
53	Poliții # Moles	Curry, Molesc	NO	MANT	(f6/83)	(eP)	Viscosity 2 (eP)
54 i	0 0.000	0.0000	31.10 F	56.87	36.50	0.1425	8.290e-002
58 58	1 1.000e-00	2 1.0006-002	48.77 F 64.88 F	59.69 62.72	43.10 43.56	0.2614	0.1361
57	2 1.000e-00		81.66 F	65.90	44,02	0.2788	0.1469 0.1586
58	3 1,000e-00 4 1,000e-00	4 000- 000	98.48 F	69.53	44,47	0.3081	0.1670
59		F 000- 000	115.2 F	73.42	44.91	0.3135	0.1735
	5 1.000e-00		10405	99.50			
60	5 1,000e-00 6 1,000e-00	g 6,000e-002	131.8 F	77.36 81.30	45.34 45.75	0.3023	0.1732
60 61 62		8,000-002	148.1 F	77.36 81.30 2 (Build 4627)	45.75	0.3023 0.3077	0.1732 0.1868 Page 1.012



Table 5. 2: working curves

HYPROTECH

TEAM LND Calgary, Alberta Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc

Unit Set: Fleld-USGPM

Date/Time: Sun May 03 02:57:56 2009

Assay: Assay-1 (continued)

10				WORK	ING CURVES			
11 12	Point #	Moles	Cum, Moles	NBP	MVT	Mass Density (Ib/It3)	Viscosity 1 (cP)	Viscosity 2 (cP)
13	8	1.000e-002	8.0006-002	164.0 F	85.11	46.14	0.3268	0.1971
14	9	1.000e-002	9.000e-002	177.5 F	87.93	46.48	0.3456	0.2070
15	10	1.000e-002	1,000e-001	187.4 F	90.30	46.72	0.3605	0.2148
16	11	2.500e-002	0.1250	214.9 F	98.72	47.37	0.4081	0.2390
17	12	2.500e-002	0.1500	247.6 F	108.1	48.12	0.4769	0.2727
18	13	2.500e-002	0.1750	278.9 F	117.3	48.82	0.5562	0.3103
19	14	2.500e-002	0.2000	310.2 F	127.1	49.50	0.6495	0.3534
20	15	2.500e-002	0.2250	341.9 F	137.7	50.17	0.7601	0.4031
21	16	2.500e-002	0.2500	373.5 F	148.8	50,82	0.9258	0.4680
22	17	2.5006-002	0.2750	404.3 F	160.3	51,44	1.135	0.5392
23	18	2.500e-002	0.3000	432.9 F	171.4	52.00	1.381	0.6172
24	19	2.500e-002	0.3250	457.3 F	181.4	52.47	1.848	0.6953
25	20	2.500e-002	0.3500	479.2 F	190.9	52.89	1.936	0.7753
26	21	2,500e-002	0.3750	500.4 F	200.5	53.28	2.277	0.8635
27	22	2.500e-002	0.4000	521.4 F	210.4	53.67	2.892	0.9634
28	23	2,5006-002	0.4250	543,4 F	221.2	54.08	3.228	1.083
29	24	2.500e-002	0.4500	566.3 F	232.9	54.47	3.934	1.228
30	25	2,5000-002	0.4750	590.1 F	245.3	54.89	4.873	1.404
31	28	2.500e-002	0.5000	615.1 F	258.7	55,32	6.175	1.624
32	27	2,500e-002	0.5250	641.9 F	273.4	55.78	8.072	1.908
33	28	2.500e-002	0.5500	670.4 F	288.9	58.26	10.93	2.280
34	29	2.500e-002	0.5750	700.3 F	304.6	56.75	15.37	2.773
35	30	2.500e-002	0.6000	731.8 F	321.7	57.26	22.63	3.444
36	31	2,500e-002	0.6250	765.7 F	343.5	57.80	35.57	4.409
37	32	2.500e-002	0.6500	801.3 F	368.6	58,35	59.92	5.815
-	33	2.500e-002	0.6750	837.4 F	393.1	58.90	107.2	7.851
38	34	2.500e-002	0.7000	873.6 F	417.2	59.45	204.4	10.85
39		2.500e-002	0.7250	909.6 F	440.9	59,98	416.2	15.36
40	35	2.500e-002	0.7500	944.4 F	463.7	60.48·	891.2	22.09
41	• 38	2,500e-002	0,7750	976.7 F	485.1	60.94	1945	31.76
42	37	2.500o-002	0.8000	1005 F	504.4	61,34	4145	44.83
43	38	2.500e-002	0.8250	1027 F	519.3	61,65	7739	59.29
4	39		0,8500	1044 F	531.5	61.88	1.283e+004	74.13
45	40	2.500e-002	0.8650	1059 F	542.1	62.08	2.034e+004	90.68
46	41	1,500e-002	0.8800	1072 F	552.2	62.26	3.027e+004	110.0
47	42	1.500e-002	0.8960	1086 F	562.5	62.44	3.876e+004	134.1
48	43	1,500e-002	0.9100	1099 F	573.0	62.62	5.004e+004	164.5
49	44	1.500e-002	0.9250	1113F	583.8	62,80	6.515e+004	203.1
50	45	1.500e-002	0.9400	1126 F	594.9	62,98	8.555e+004	252.4
51	46	1,500e-002	0.9550	1140 F	608.5	63.16	1.138e+005	316.3
52	47	1.500e-002	0.9700	1154 F	618.6	63,34	1.536e+005	401.9
53	48	1.500e-002	0.9850	1161 F	624.2	63,44	1.798a+005	455,4
54	49	1,500e-002	1,000	1174 F	634.7	63,61	2.443e+005	580.5
55	50	1,500e-002	1,000					300,5

HP/SYSV8-18 (BOLD 4827)

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Page 2 of 2 Specified by user.



Table 5. 3: component properties

_									
1	_			Case	Name: C:\Prog	ram FilesV-ty	protechilitys	YS 3,1.3\Sem	ples/dyncrude3.hsc
3	HYPROT	ECH Cal	AM LND gary. Alberta	Unit :	Set: Fleid-U	SGRM			
3 4 5		CA!	NADA	Date	Time: Sun Ma	y 03 04:11:12	2009,		
6									
7 8			Blend: Bl	end-1					
9					DATA		<u> </u>	· · · · · · · · · · · · · · · · · · ·	
11				Oil Flow	Information				
12 13		O2	1.14.254 1.1	គ	ow Units		erija in jira	Oil Flow	Rate
14		Assay-1		Li	quid Vol				
15	7			Cut Ranges:	User Points				
16 17				Number of Cuts:	30				
18	j			Ru	ik Data				
19									
20 21	Molecular Weight: Viscosity 1 Temp:	100,00	- Mass Density: F * Viscosity 1:		Watson Uopk: Viscosity 2 Tem				
22	Viscosity (Temp.	100.00	VISCUSHY 1.	2000		y2	10.00 F	Viscosity 2:	_
23				CORR	ELATIONS				
24 25				Def	ault Set				
26	Low End T	High End T.	To		Pe		W		Cp
27	(F)	<u> </u>		**************************************		Brien, F.		प्रदेशका होता.	
28	-459.67	2192.00	l Le	e_Kesler	Lee_Kesier		L00_I	Kesler	Leo_Kesier
29				TA	ABLES				
29 30 31 32			Compon	ent Properties:	Blend-1				
				Mol Wt.			15000		
33	Comp Name	•	NBP (F)	NOI VVI.	Density (Ib/ft3)		Visco:	-	Viscosity 2 (cP)
3 3	1	NBP_52	51.88	58.		41,08		0.25831	0.12462
36		NBP_79	79.08	63. 70.		43.72		0.28582	0.15104
37		8P_111	111.32 143.48	78.		44.59 45.43		0.31007	0.16897
38 39		BP_143 BP_178	177.64	88.		46.33		0.33670	0.17622 0.20234
3 €		BP_208	208.15	92.		48.93		0.37490	0.22223
41	N	BP 240	239.84	101. 110.		47.56		0.42365	0.24673
42		BP_272	271.61 303.55	120.		48.32		0.49873	0.28226
43		BP 304 BP 336	335,68	130.		49.71		0.68004	0.32219 0.36749
44 45		BP_368	367.82	141.		50.39		0.80707	0.42237
46	N	BP 400	400,08	153. 166.		51.06 51.73	·	0.99671	0.49271
47		BP 433	432.70 464.48	179.	·····	52.40		1.2527	0.57726 0.68092
48		3P 464 3P 496	496.09	193.	97	53,01		2.0310	0.80043
49 50					- I	62 E0		2.5091	0.94159
		BP 528	527.77	208,		53.59			
51	N N	BP 528 BP 560	559.95	224.	28	54.16		3.3704	1,1135
52	N N	SP 592	559.95 591.91		98 22			3.3704 4.4384 5.9288	1.3249
52 53	N N N	SP 560 SP 592 SP 624	559.95	224. 240.	98 22 79	54.16 54.72		4.4384	
51 52 53 54	M M M M	SP 560 SP 592 SP 624 SP 656	559.95 591.91 623.92 656.08 686.17	224) 240: 256: 273. 290/	06	54.16 54.72 55.26 55.79 56.31		4.4384 5.9288 8.0822 11.294	1,3249 1,5840 1,9090 2,3236
52 53 54 55	N N N N N	SP 560 SP 592 SP 624	559.95 591.91 623.92 656.08 686.17 720.16	224. 240. 256. 273. 290. 307.	222 2799 278 284 211 211 211 211 211 211 211 211 211 21	54.16 54.72 55.26 55.79 56.31 56.82		4.4384 5.9288 8.0822 11.294 15,161	1,3249 1,5840 1,9090 2,3236 2,8532
52 53 54 55 56 57	NS N	SP 560 SP 592 SP 624 SP 656 SP 688 SP 720 SP 752	559.95 591.91 623.92 656.08 686.17 720.16 752.29	224) 240: 256: 273. 290/	79 78 94 94 94 94 94 94 94 94 94 94 94 94 94	54.16 54.72 55.26 55.79 56.31		4.4384 5.9288 8.0822 11.294 15.161 23.725	1.3249 1.5840 1.9090 2.3236 2.8532 3.5358
52 53 54 56 56 57	NS N	SP 560 SP 592 SP 624 SP 656 SP 688 SP 720 SP 752 SP 784	559.95 591.91 623.92 656.08 686.17 720.16	224) 240, 256, 273, 290, 307, 324,	79	54.18 54.72 55.26 55.79 56.31 58.82 57.32 57.82 58.55		4.4384 5.9288 8.0822 11.294 15,161	1,3249 1,5840 1,9090 2,3236 2,8532
52 53 54 68 56 57 58	NS N	SP 560 SP 592 SP 624 SP 656 SP 688 SP 720 SP 752 SP 784 SP 832	\$59.95 591.91 623.92 656.08 685.17 720.16 752.29 764.46 832.24 895.94	224) 240, 256, 273, 290, 307, 324, 344, 377, 420,	79	54.16 54.72 55.26 55.79 56.31 56.82 67.32 57.82 58.55 59.51		4.4384 5.9288 8.0822 11.294 16.161 23.725 36.110 73.226 221.62	1,3249 1,5840 1,9090 2,3236 2,8532 3,5356 4,4445 6,4545 11,297
52 53 54 56 57 59 60	NS N	SP 560 SP 592 SP 624 SP 656 SP 688 SP 720 SP 752 SP 784	559.95 591.91 623.92 656.08 686.17 720.16 752.29 784.46 832.24 895.94	224) 240, 256, 273, 290, 307, 324, 344, 377, 420, 463,	22 2 79 78 8 4 4 4 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1	54.16 54.72 55.26 55.79 56.31 56.82 57.32 57.32 57.82 58.56 59.51 60.48		4.4384 5.9288 8.0822 11.294 16.161 23.725 36.110 73.226 221.62 877.81	1.3249 1.5840 1.9090 2.3236 2.853 3.5356 4.4445 6.4545 11.297 21.929
52 53 54 55 56 57 59 60 61	100 100 100 100 100 100 100 100 100 100	SP 560 SP 592 SP 624 SP 656 SP 638 SP 720 SP 752 SP 784 SP 635 SP 960 SP 1027	559.95 591.91 623.92 656.08 685.17 720.16 752.29 784.46 832.24 885.94 960.31	224) 240, 256, 273, 290, 307, 324, 344, 377, 420, 463,	22 2 79 78 84 21 1 34 35 8 8 8 8 8 8 8 8 8 8 8 8 8 8 8 8 8 8	54.16 54.72 55.26 55.79 56.31 56.82 67.32 57.82 58.55 59.51		4.4384 5.9288 8.0822 11.294 16.161 23.725 36.110 73.226 221.62 877.81 5790.1	1.3249 1.5840 1.9090 2.2236 2.8532 3.5358 4.4445 6.4545 11.297 21.929 52.097
52 53 54 55 56 57 59 60 61 62	10 10 10 10 10 10 10 10 10 10 10 10 10 1	SP 560 SP 592 SP 624 SP 656 SP 688 SP 720 SP 752 SP 752 SP 784 SP 832 SP 836 SP 960 > 1027	559.95 591.91 623.92 656.08 686.17 720.16 752.29 784.46 832.24 895.94	224) 240, 256, 273, 290, 307, 324, 344, 377, 420, 463,	22 2 79 78 84 21 1 34 35 8 8 8 8 8 8 8 8 8 8 8 8 8 8 8 8 8 8	54.16 54.72 55.26 55.79 56.31 56.82 57.32 57.32 57.82 58.56 59.51 60.48	NOW	4.4384 5.9288 8.0822 11.294 16.161 23.725 36.110 73.226 221.62 877.81	1,3249 1,5840 1,9090 2,2236 2,8532 3,5558 4,4445 6,4545 11,297 21,929 52,097



Table 5. 4: component properties

1 2		TEAM LND	Caso Name	: C:\Program Files\Hyp	notech\HYSYS 3,1.3\Sam	ples/dyncrude3.hsc
3	HYPROTECH	Calgary, Alberta	Unit Set:	Field-USGPM		
5		CANADA	Date/Time:	Sun May 03 04:11:12	2009	
6						
7 8		Blend: Ble	end-1 (contin	ued)		
9			TABLE			
10 11						
12		Compon		end-1		
13 14	Comp Name	NBP (F)	Mol Wt.	Density (Ib/ft3)	Viscosity 1 (cP)	Viscosity 2 (cP)
15	NBP_1085	1085.44	557.90	62.38	34418	120.98
16 17	NBP_1153	1152.70 Critical Temp	612.85 Critical Press	63.25	1.3030e+005	352.69
18	Comp Name	(F)	(psia)	Accentric Fac.	Watson K	
19	NBP_52	357.75	566.97	0.13703	12.130	
20	NBP_79	391.15 426.48	554.07 534.80	0.16622 0.21071	11.595	
21 22	NBP_111 NBP_143	483.23	523.29	0.24964	11.591	
23	NBP_178	503.45	507.26	0.28193	11,573	*********************
24	NBP_208	537.38	480.05	0.30882	11.603	
25	NBP_240	571.40	450,45	0.33824	11.629	-
26	NBP_272	605.78	426.25	0.36763	11.818	
27	NBP_304	639.44	402.95	0.39869	11.614	
28	NBP_338	672.61	381.00 360.68	0.43111 0.46445	11.612	·
29	NBP_388	705.25 737.58	341.73	0.49874	11.509	
30	NBP_400	769,92	324.14	0.53402	11.595	
31	NBP_433 NBP_464	801.19	308,38	0.56881	11.582	
32 33	NBP_496	831.60	293.11	0.60470	11.577	
34	NBP_528	861.54	278.42	0.64174	11.576	
35	NBP_560	891.49	264.22	0.68028	11.578	
36	NBP_592	920.98	250.98 238.39	0.71921	11.579	
37	NBP_624	950.10	228.43	0.75899 0.79964	11.581	-
38	NBP_656	979.07 1007.8	215.21	0.84070	11.584	\$*************************************
39	NBP 688	1038,1	204.62	0.58225	11.588	
40	NBP_720 NBP_752	1084.3	194.49	0.92469	11.588	
41	NBP_784	1092.4	184.92	0.96356	11,589	*****
42	NBP_832	1133.9	171.78	1.0218	11.588	
43 44	NBP 896	1188.6	155.80	1,0973	11.586	
45	NBP_960	1243.6	141.36	1.1708	11.579	
46	NBP_1027	1301.1	128.24 117.64	1.2429 1.3042	11,581	
47	NEP_1085	1349.9	105.98	1,3748	11.561 11.550	
48	NBP_1153	1405.4		end-1		<u> </u>
49 53					Mass Flow	
51	Comp Name	Liquid Vol %	LiqVol % Cum	Vol Flow (USGPM)	(ID/hr)	Moter Flow (ibmote/hr)
52	***	0.0065	0.0068	-		
53	Methane	0.0225	0.0290			
54 57	Ethane Propane	0.3200	0.3490		***	
58 56	i-Butane	0.2400	0.5890			
57	n-Butane	0.8200	1.4090			
58	H2O	0.0000	1.4090			
59	NBP_52	1,3669	2.7759 4,6685			······································
60	NBP_79	1.8927	6.5608			A TO SEE THE SECOND SEC
_	NBP_111	1.8923	The second secon			······································
61		1,9426	8.5034	Dilla 4827)		

Licensed to: TEAM LND

Specified by user



Table 5. 5: component break down

[]	<u> </u>					
2		TEAM LND	Case Name	C:\Program Files\Hy	protech\HYSYS 3.1.3\Semp	les/dyncrude3.hsc
4		Calgary, Alberta CANADA	Unit Set:	Field-USGPM	<u> </u>	
5			Date/Time:	Sun May 03 04:11:12	2009	
6 7		Blend: Ble	end-1 (contin	uod)		
8		Dictio. Dic	sna-i (contin	ueu)		
9			TABLE	S		
11		Compone	nt Breakdown: Ble	end-1		
13		Liquid Vol %	LiqVel % Cum	Vol Flow		
14		2 5254		(USGPM)	Mass Flow (Bu/hr)	Motar Flow (lbmote/hr)
16		2.5971 2.0168	11.1006 13.1173			
17	A THE RESIDENCE OF THE PARTY OF	2.4670	15.5844			
18		2.5318	18.1162			
19	NBP_304	2.5191	20.6353			~ · · · · · · · · · · · · · · · · · · ·
20	N8P_336	2.4866	23.1219			
21	NBP_368	2.5083	25.6302	_		
22	NBP_400	2.6168	28.2470			···
23	NBP_433	2.9853	31,2324			-
24	NBP_464	3.5364	34,7687		-	····
25	NBP_496	3.7526	38.5213			
26	NBP_528	3.6375	42.1588			_
27	NBP_560	3.4150	45.5738 48.8084			
28	NBP_592	3.2346 2.9852	51.7938			
29	NBP_624	2.7726	54.5662	· · · · · · · · · · · · · · · · · · ·		
30 31	NBP_656 NBP_688	2.6323	57.1985			
32	NBP_720	2.4833	59.6819			
33	NBP_752	2.3150	61.9969			
34	NBP_784	2.2138	64.2107			
35	NBP_832	4.3232	68.5339	_		
36	NBP_896	4,3608	72.8947			
37	NSP_960	4.8190	77.7138			
38	NBP_1027	7.6512	85.3649		_	
39	NBP_1085	7.0358	92,4007			
40	NBP 1153	7.5993	100.0000			
41			Total:			
42 43			Molar Compo	sitions		
44		Blend-1				
45	Methano	0.00	03			
46	Ethane	0.00	08	***************************************		***************************************
47	Propone	0.00	88			
48	i-Butane	0.00				-
49	n-Butano	0.01				
50	H2O	0.00		*		
51	NBP_52	0.03	****			···· · · · · · · · · · · · · · · · · ·
52	NBP_79	0.04				
33	NBP_111	0.04				
4	NBP_143	0.05				
즤.	NBP 178	0.03				
	NBP_208	0.04	******			
_	NBP 240	0.04		L		
	NBP 272	0.03				
-	NBP_304 NBP_338	0.03		and the same of th		and the second
-	NBP_368	0.03	******		******************************	***************************************
-	NBP 400	0.03	OR .			
z Z	hyprotech (3th	NAS TURBLE SERVICE	HYSYSVAF3 (B	in 4820 to		Page 3 Gt 7/
	icensed to: TEAM LND					* Specified by user.



Table 5. 6: molar composition

1	_		Case Name:	C:\Program Files\Hyprotech\HYSYS 3	3.1.3\Samples\dyncrude3.hsc
2	HYPROTECH	TEAM LND Colgory, Alberta	Unit Set:	Field-USGPM	
4		CANADA		Sun May 03 04:11:12 2009	
5 6				The second secon	
7		Blend: Ble	nd-1 (continued)	
8 9					
10			TABLES		
11			Molar Composition	ns	
12 13		Blend-1			1
14	N8P_433	0.03			
15	NBP_464	0.03			
16 17	NBP_496 NBP_528	0.03			
18	NBP_560	0.03			
19	NBP_592	0.02			
20 31	NBP 624	0.02			
21 22	NBP_656 NBP_688	0.02			
23	NBP_720	0.01	172		
24	NBP_752	0.01			
25	NBP_784	0.01			
26 27	NBP_832 NBP_896	0.02			
	1101 000				
28	NBP_960	0.02	235		
	NBP_960 NBP_1027	0.03	343		
28 29 30	NBP_1027 NBP_1085	0.03	243 294		
28 29 30 31	NBP_1027	0.03	343 194 193		Paddaday
28 29 30 31 32 33	NBP_1027 NBP_1085	0.03 0.02 0.02	M3 194 1983 Oil Properties: Blend	······································	Bastisquid Volume
28 29 30 31 32 33	NBP_1027 NBP_1085	0.03 0.02 0.02	343 194 193	I-1 Accentric Factor	Basisquid Votume Molocutar Weight
28 29 30 31 32 33 34 35	NBP 1027 NBP 1085 NBP 1153	0.03 0.02 0.02 Critical Temp (F) -166.91	Oil Properties: Blend Critical Pressure (psia) 665.65	Accentric Factor -9.7924e-003	Molecular Weight
28 29 30 31 32 33 34 35	NBP 1027 NBP 1085 NBP 1153 Percent 0.00 1.00	0.03 0.02 0.02 Critical Temp (F) -166.91 305.74	Oil Properties: Blend Critical Pressure (psia) 665.65 550.67	Accentric Factor -9,7924e-003 0.20094	Molecular Weight 12.624 58.125
28 29 30 31 32 33 34 35 36 37	NBP 1027 NBP 1085 NBP 1153 Percent 0.00 1.00 2.00	0.03 0.02 0.02 Critical Temp (F) -166.91 305.74 354.10	Oil Properties: Blend Critical Pressure (psia) 665.65	Accentric Factor -9.7924e-003	Molecular Weight 12.624 58.125 58.737
28 29 30 31 32 33 34 35 36 37 38	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50	0.03 0.02 0.02 Critical Temp (F) -166.91 305.74	143 194 193 Oil Properties: Blend Critical Pressure (psla) 665.65 550.67 565.56	Accentric Factor -9.7924e-003 0.20094 0.14244	Molecular Weight 12.624 58.125
28 29 30 31 32 33 34 35 36 37 38 39 40	NBP 1027 NBP 1085 NBP 1153 Percent 0.00 1.00 2.00	0.03 0.02 0.02 Critical Temp (F) -166.91 305.74 354.10 388.11 415.14 462.63	143 1294 1293 Oil Properties: Blend Critical Pressure (psia) 665,65 590,67 565,55 556,05 540,77 523,47	-9,7924e-003 0,20094 0,14244 0,16175 0,19603 0,24907	Molecular Weight 12.624 58.125 58.737 63.046
28 29 30 31 32 33 34 35 36 37 38	NBP 1027 NBP 1085 NBP 1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00	Critical Temp (F) -166.91 305.74 354.10 388.11 415.14 462.63 506.52	143 1294 1293 Oil Properties: Blend Critical Pressure (psia) 665,65 550,67 595,55 556,05 540,77 523,47 505,23	-9,7924e-003 0,20094 0,14244 0,16175 0,19603 0,24907	Molecular Weight 12.624 58.125 58.737 33.045 68.344 78.167
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43	NBP 1027 NBP 1085 NBP 1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50	0.03 0.02 0.02 Critical Temp (F) -166.91 305.74 354.10 388.11 415.14 462.63 506.52 543.35	143 1294 1293 Oil Properties: Blend Critical Pressure (psia) 665,65 590,67 565,55 556,05 540,77 523,47	-9,7924e-003 0,20094 0,14244 0,16175 0,19603 0,24907	Molecular Weight 12.624 58.125 58.737 83.045 68.344 78.167 87.324
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00	Critical Temp (F) -166.91 305.74 354.10 388.11 415.14 462.63 506.52	Oil Properties: Blend Crâtical Pressure (psla) 665.65 550.07 565.55 556.05 540.77 523.47 505.23 474.75 443.64	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548	Molecular Weight 12.624 58.125 58.737 33.045 68.344 78.167
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45	NBP 1027 NBP 1085 NBP 1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50	Critical Temp (F) -166.91 305.74 354.10 338.11 415.14 462.63 506.52 543.35 580.59 614.51 647.74	294 294 293 Oil Properties: Blend Critical Pressure (psia) 665.65 550.67 565.55 556.05 540.77 505.23 474.75 443.84 420.14 397.36	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40666	Molecular Weight 12.624 58.125 58.737 63.046 68.344 78.167 87.324 94.311 103.59 113.08
28 29 30 31 32 33 34 35 38 37 38 39 40 41	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00 17.50 20.00 25.00	Critical Temp (F) -166.91 305.74 354.10 338.11 415.14 462.63 506.52 543.35 560.59 614.51 647.74 713.28	43 1934 1934 1934 1934 1935 1945 1955	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548	Molecular Weight 12.624 58.125 58.737 63.045 68.344 78.167 87.324 94.311 103.69 113.08 122.71
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00 17.50 20.00 25.00 30.00	0.03 0.02 0.02 0.02 0.02 0.02 0.02 0.02	294 294 293 Oil Properties: Blend Critical Pressure (psia) 665.65 550.67 565.55 556.05 540.77 505.23 474.75 443.84 420.14 397.36	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287	Molecular Weight 12.624 58.125 58.737 63.045 68.344 78.167 87.324 94.311 103.59 113.08 122.71 144.18 167.28
28 29 30 31 32 33 34 35 36 37 38 40 41 42 43 44 45 46 47 48	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00 17.50 20.00 25.00 30.00 35.00	Critical Temp (F) -166.91 305.74 354.10 338.11 415.14 462.63 506.52 543.35 560.59 614.51 647.74 713.28	43 1934 1934 1934 1934 1934 1935	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40666 0.47287 0.53669 0.58866 0.63823	Molecular Weight 12.624 58.125 58.737 63.046 68.344 78.167 87.324 94.311 103.69 113.08 122.71
28 29 30 31 32 33 34 35 36 37 38 40 41 42 43 44 45 46 47	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00 17.50 20.00 25.00 30.00	Critical Temp (F) -166.91 305.74 354.10 388.11 415.14 462.63 506.52 543.35 580.59 614.51 647.74 713.28 772.63 818.22 658.76 901.38	43 294 293 294 293 294 293 294 295 2	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287 0.53689 0.58866 0.63823 0.99321	Molecular Weight 12.624 58.125 58.737 83.045 68.344 78.167 87.324 94.311 103.59 113.08 122.71 144.18 167.26 187.56 207.21 229.41
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00 17.50 20.00 25.00 35.00 40.00 46.00	Critical Temp (F) -166.91 305.74 354.10 388.11 415.14 462.63 506.52 543.35 580.59 614.51 647.74 713.28 772.63 818.22 658.76 901.38 947.19	43 1934 1934 1934 1934 1934 1935	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40666 0.47287 0.53669 0.58866 0.63823	Molecular Weight 12.624 58.125 58.737 83.045 68.344 78.167 87.324 94.311 103.59 113.08 122.71 144.18 167.56 207.21 229.41
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53	NBP 1027 NBP 1085 NBP 1153 Percent 0.00 1.00 2.00 3.50 5.00 12.50 15.00 17.50 20.00 35.00 35.00 40.00 45.00 56.00	Critical Temp (F) -166.91 305.74 354.10 338.11 415.14 462.63 506.52 543.35 580.59 614.51 647.74 713.28 772.63 818.22 858.76 901.38 947.19	243 294 293 294 293 294 293 294 293 294 295	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287 0.53689 0.58866 0.63923 0.69321 0.75497 0.82698	Molecular Weight 12.624 58.125 58.737 83.045 68.344 78.167 87.324 94.311 103.59 113.08 122.71 144.18 167.26 187.56 207.21 229.41
28 29 30 31 32 33 34 35 38 39 40 41 42 43 44 45 46 47 48 49 50 51 51 52 53	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 12.50 15.00 17.50 20.00 35.00 35.00 40.00 45.00 55.00 60.00	Critical Temp (F) -166.91 305.74 354.10 388.11 415.14 462.63 506.52 543.35 580.59 614.51 647.74 713.28 772.63 818.22 658.76 901.38 947.19	43 1934 1934 1934 1934 1934 1934 1935 1934 1935	Accentric Factor -9.7924e-003 0.20094 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287 0.53669 0.58866 0.63823 0.69321 0.75497 0.82698 0.90978	Molecular Weight 12.624 58.125 58.737 63.046 68.344 78.167 87.324 94.311 103.59 113.08 122.71 144.18 167.26 207.21 229.41 2255.11 2265.21 317.96 363.67
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 55 53 54	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00 17.50 20.00 25.00 30.00 35.00 40.00 45.00 50.00 50.00 60.00	Critical Temp (F) -166.91 305.74 354.10 338.11 415.14 462.63 506.52 543.35 550.59 614.51 647.74 713.28 772.63 818.22 858.76 901.33 947.19 998.26 1054.3 1116.4	43 1934 1934 1934 1934 1934 1934 1935	Accentric Factor -9.7924e-003 0.20994 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287 0.53669 0.58666 0.63823 0.99321 0.75497 0.82698 0.90978	Molecular Weight 12.624 58.125 58.737 63.046 68.344 78.167 67.324 94.311 103.59 113.08 122.71 144.18 167.26 207.21 229.41 255.11 285.21 317.96 363.67
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 12.50 15.00 17.50 20.00 35.00 35.00 40.00 45.00 55.00 60.00	Critical Temp (F) -166.91 305.74 354.10 388.11 415.14 462.63 506.52 543.55 540.59 614.51 647.74 713.28 772.63 818.22 658.76 901.38 947.19 988.26 1054.3 1116.4 1179.8	43 1934 1934 1934 1934 1934 1934 1935 1934 1935	Accentric Factor -9.7924e-003 0 20994 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287 0.53669 0.58866 0.63923 0.99321 0.75497 0.82698 0.90978 0.90978	Molecular Weight 12.624 58.125 58.737 63.046 68.344 78.167 67.324 94.311 103.59 113.08 122.71 144.18 167.26 207.21 229.41 255.11 285.21 317.96 363.67 413.67
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 55 55 55 56 55 56 55 58	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00 17.50 20.00 25.00 30.00 35.00 40.00 45.00 56.00 66.00 65.00 70.00 75.00 80.00	Critical Temp (F) -166.91 305.74 354.10 338.11 415.14 442.63 506.52 543.33 580.59 614.51 647.74 713.28 772.63 818.22 858.76 901.38 947.19 999.26 11064.3 1119.4 1179.8 1240.2	43 1934 1934 1934 1934 1934 1934 1935	Accentric Factor -9.7924e-003 0.20994 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287 0.53669 0.58666 0.63823 0.99321 0.75497 0.82698 0.90978	Molecular Weight 12.624 58.125 59.737 63.045 68.344 78.167 67.324 94.311 103.59 113.08 122.71 144.18 167.26 207.21 229.41 2255.11 285.21 317.96 363.67 413.67
28 29 30 31 32 33 34 35 38 39 40 41 42 43 44 45 46 47 48 49 50 51 55 55 56 57 58 59	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 17.50 20.00 25.00 30.00 25.00 30.00 40.00 45.00 50.00 65.00 60.00 65.00 70.00 75.00 80.00	0.03 0.02 0.02 0.02 0.02 0.02 0.02 0.02	243 294 293 294 293 294 293 294 295	Accentric Factor -9.7924e-003 0 20994 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287 0.53669 0.58866 0.63923 0.99321 0.75497 0.82698 0.90976 0.99723 1.0851 1.1662	Molecular Weight 12.624 58.125 58.737 63.045 65.344 78.167 87.324 94.311 103.59 113.08 122.71 144.18 167.28 187.66 207.21 229.41 255.11 285.21 317.96 363.67 413.67
28 29 30 31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 53 54 55 56 57 58	NBP_1027 NBP_1085 NBP_1153 Percent 0.00 1.00 2.00 3.50 5.00 7.50 10.00 12.50 15.00 17.50 20.00 25.00 30.00 35.00 40.00 45.00 56.00 66.00 65.00 70.00 75.00 80.00	Critical Temp (F) -166.91 305.74 354.10 338.11 415.14 442.63 506.52 543.33 580.59 614.51 647.74 713.28 772.63 818.22 858.76 901.38 947.19 999.26 11064.3 1119.4 1179.8 1240.2	143 1934 1934 1934 1934 1934 1935 193	Accentric Factor -9.7924e-003 0 20994 0.14244 0.16175 0.19603 0.24907 0.28434 0.31388 0.34603 0.37548 0.40665 0.47287 0.53699 0.58866 0.63823 0.89321 0.75497 0.82598 0.90978 0.90978	Molecular Weight 12.624 58.125 59.737 63.045 68.344 78.167 67.324 94.311 103.59 113.08 122.71 144.18 167.28 187.66 207.21 229.41 225.11 285.21 317.96 363.67 413.67 461.12 501.95

Licensed to: TEAM LNO

* Specified by user.



Table 5. 7: oil properties

1 2				Case Name: 0	:\Program Files\Hyprote	ch\HYSYS 3.1.3\Sample	s\dyncrude3.hsc		
3	HYPROTE	TEAM LND Calgary, Alberta		Unit Set; F	Field-USGPM				
4	Contest ness	CANADA		Date/Time: Sun May 03 04:11:12 2009					
5			MARKALLAN PLAN	The state of the s					
		Blend	: Blend-1	(continued))				
9		TALL DO ALSO ALAIM SEE		TABLES	CONTRACTOR AND				
11			Oil	Properties: Blend-	1		Bastiquid Volume		
13	Percent	Critical Temp (F)	Cr	itical Pressure (psia)	Accentric Facto	r Mol	ecular Weight		
15	96.50		07.7	105.55		1.3777	615.10		
16	98.00	The second secon	19.0	103.46		1.3922	626.37		
17	99.00		26.6	102.11		1.4018	633.87		
18	100.00		34.2	100.79		1.4115	641.38		
19 20	Percent	Density (Ib/ft3)		Viscosity 1 (cP)	Viscosity 2 (cP)	25 17 18 18 10 18 17 28 18			
21	0.00		.809	6.1706e-003		37e-002			
22	1.00		.340	0.13571	7.59	067e-002			
23	2.00		.589	0.23781 0.28201		0.11965 0.15017			
24	3.50 5.00		.343	0.30346		0.16459			
6	7.50		.325	0.30355		0.17608			
7	10.00		.295	0.33971		0.20418			
8	12.50	Control of the second s	.944	0.38248		0.22609			
9	15.00	47	.656	0.44096	***************************************	0.25526			
0	17.50	48	.404	0.51739		0.29209	***		
1	20.00		.098	0.60437 0.84778		0.33282			
2	25.00		.450	1.2784		0.43838 0.58535	· · · · · · · · · · · · · · · · · · ·		
3	30,00		.686	1.8249		0.74528			
4	35.00	CONTRACTOR OF STREET CONTRACTOR OF STREET	432	2.5389		0.92730			
5	40.00	THE RESERVE AND ADDRESS OF THE PARTY OF THE	240	3.6887		1.1794	N. S.		
7	45.00 50.00	55	.093	5.7545		1.5555			
8	55.00	56	.025	10.076		2.1742			
9	60.00	57	.030	20.612		3.2705			
ō	65.00		.127	53.665		5.4873			
1	70.00		.240	182.31		10.254 20.973			
2	75.00	AND THE RESIDENCE OF THE PARTY	294	798.87 3690.7		42.431			
3	80.00		.163	14207		77.441			
4	85.00		.350	43372		138.81	- 1 - 1 - 1 - 1 - 1 - 1 - 1 - 1 - 1 - 1		
5	90.00	THE RESERVED THE PROPERTY OF THE PARTY OF TH	.634	70128		194.08			
6	92.50	The second secon	.953	1.0785e+005		285.93			
7	95.00	THE RESERVE OF RESERVE AND ADDRESS OF THE PERSON NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT OF THE PERSON NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT NAMED IN COLUMN TWO IS NOT THE PERSON NAMED IN COLUMN TRANSPORT NAMED IN COLUMN	.160	1.3760e+005		368.49			
8	96.50 98.00	63	344	1.8078e+005		458.85			
0	99.00	The state of the s	.466	2.1685e+005		531.10			
9 1	100.00	63	.588	2.6012e+005	/-	614.72			
2			Oil BP Tem	peratures: Blend-			Bastiquid Volume		
4	Percent	ТВР	ASTM D86 (F)	D86 Crack Reduced (F)	D1160 Vac. (F)	D1160 ATM (F)	D2887 (F)		
5		-3.1676	86.282	86.282	-125.20	8.9307	-23.883		
3	0.00	31.122	108.74	108.74	-101.73	40.314	-10.048		
7	1.00	50.475	121.95	121.95	-88.247	58.266	3,4240		
4	2.00 3.50	75.461	139.58	139.58	-70.571	81.703	23.09		
9	5.00	100.84	158.09	158.09	-52.335	105.77	42.32		
1	7.50	142.94	190.00	190.00	-21.501	146.22	74.07		
2	10.00	180.34	219.52	219.52 YSYS v3.1.3 (Build 48	6.4667	182.64	107.26		
-	Hyprotech Ltd.		WHEN THE RESIDENCE OF THE PARTY	1010 VO. 1.0 (Dulld 40		the state of the s	Page 5 of		

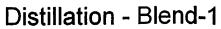


Table 5. 8: oil BP temperature

2		754444		Case Name: C	:\Program Files\Hyprotec	hiHYSYS 3,1.3\Semples\c	lyncrude3.hsc
3 p	HYPROTE	TEAM LND Calgary, Albe	erta	Unit Set:	ield-USGPM		
<u>4</u>		CANADA		Date/Time: S	un May 03 04:11:12 2009		
+					<u> </u>		
7		Blei	nd: Blend-1	(continued))		
<u> </u>							
9				TABLES			
11			Oil BP Tem	peratures: Blend-	1		Basisquid Volum
13	Percent	TBP (F)	ASTM D86 (F)	D86 Crack Reduced (F)	D1160 Vac. (F)	D1160 ATM (F)	D2887 (F)
5	12.50	213.68	246.67	246.67	31.822	215.43	140.4
6	15.00	248.32	275.60	275.60	58.541	249.77	172.0
긱	17.50	279.81	302.48	302.48	83,151	281.20	202.5
8	20.00	311.54	330.03	330.03	108.20	312.99	232.6
9	25.00	375.79 435.44	387.05 440.72	387.05 440.72	159.61	377.62	293.6
21	30.00 35.00	482.10	482.52	482,52	207.93 245.99	437.63	358.7
2	40.00	524.81	520.93	513.84	280.87	484.39	422.90
23	45.00	570.63	582.45	551.72	318.03	526.88 571.75	477,45 630.41
4	50.00	620.71	608.16	591.58	358,98	620.73	530.4° 587.9°
5	55.00	677.49	660,47	634.09	406.77	677.29	648.1
6	60.00	740.81	719.14	676.98	461.08	740,79	701.8
7	65.00	812.06	785.21	718,70	522.88	812.08	759.9
8	70.00	585.56	853.19	754.68	587.57	885.58	824.5
9	75.00	958.30	918.19 969.52	783.39 802.83	650.77	956.31	889.4
<u> </u>	80.00	1012.8 1055.8	1008.2	815.91	701.86 741.19	1012.7 1065.7	951.8
1 2	85.00	1095.4	1043.4	826.80	777.76	1095,4	1008.8
3	90.00 92.50	1118.2	1063.5	832.65	798.97	1118.2	1066.6 1102.4
1	95.00	1141.5	1083.8	838,32	820.70	1141.5	1143.8
5	96.50	1155.5	1095.9	841,60	833.82	1155.5	1170.9
3	98.00	1169.4	1107.9	844.77	846.88	1169,4	1199.5
7	99.00	1178.6	1115.8 1123.5	846.81 848.79	855.52 864.08	1178,6	1219.3
-	100.00	1187.6		Properties: Blend-		7107.5	1239.
<u> </u>			Oil User P	Toperties. Bielig-			Bastsiquid Volume
4-	Percent						· · · · · · · · · · · · · · · · · · ·
4	1.00						
}	2.00						
-	3.50						
	5.00						
	7.50						
	10.00						
1	12.50	*****					******
.	15.00				and the second s		•
	17.50						
ł	20.00						***************************************
	25.00 30.00						
ł	35.00						
†	40.00						
1	45.00						
1	50.00						••••••••••••••••••••••••••••••••••••••
	55.00						
1	60.00	ar may a a san					
					with the opening the state of t	L	
]	65.00	and the games and a games of the same of t					
	65.00 70.00 yyydfach 11d		The state of the s	SYSTATIA (Bidli 482			Page & of



Figure 5. 1: TBP curve



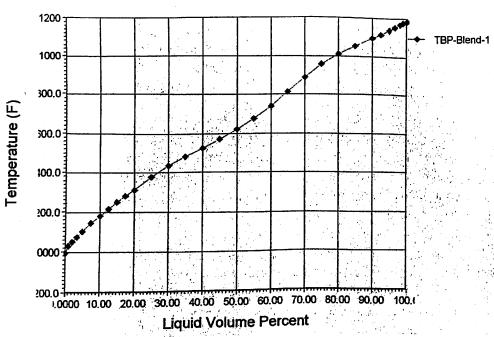


Table 5. 9: Oil distribustion table

	Oil distribustion table-	HYSYS	
Name	Begin T [F]	End T [F]	Fraction
Off Gas	42.30155119	49.9998901	1.15E-02
Lt St Run	49.99998901	157.999989	7.32E-02
Naphtha	157.999989	355.999989	0.149917137
Kerosene	355.999989	464.0000439	9.49E-02
Light Diesel	464.0000439	554.0000439	0.102819001
Heavy Diesel	554.0000439	644.0000439	8.89E-02
Atm Gas Oil	644.0000439	698.0000439	4.56E-02
Residue	698.0000439	1187.622578	0.433166452
_			L



Atmospheric Crude Colum (ADU)

Atmospheric Crude Columns are one of the most important pieces of equipment in the petroleum refining industry. Typically located after the Desalter and the Crude Furnace, the Atmospheric Tower serves to distil the crude oil into several different cuts. These include naphtha, kerosene, light diesel, heavy diesel, and AGO.

The column of atmospheric distillation is the unit where the most part of the crude fractioning is produced. The model is composed of a column with 29 theoretical trays, with a partial condenser and without a reboiler. It has 3 side columns of 3 theoretical trays each, which are of kerosene with a reboiler, Diesel and AGO without a condenser or reboiler. The column includes 3 pumparounds in charge of recirculating the liquid between trays 2 and 1, 17 and 16, 22 and 21. At tray 28, there is a trim heater that allows a temperature variation at the flash zone. The crude is fed to the main column at tray 28, while at tray 29 steams is fed. The products of the main column are: the condenser off-gas, naptha, and the bottoms (TOPPED); from trays 9, 17 and 22 come out three streams that feed tray 1 of the side columns: they have as bottom products the streams of Kerosene, Diesel and Atmospheric Gas Oil (AGO). These side columns give in return to the main column, the most volatile components of the separation to trays 8, 16 and 21 respectively. For the separation in the side columns, these are fed with steam at tray number 3.

Operation variables: flow, temperature, and feeding pressure of steam to the main column (STEAM); flow, temperature and feeding pressure of steam to side column 1 (STEAM-1); flow, temperature and feeding pressure of steam to side column 2 (STEAM-2); flow, temperature and feeding pressure of steam to side column 3 (STEAM-3); rate of reflux.

Model variables: flow of vapour from the condenser.

All the process flow diagram and simulated results are shown below



Figure 5. 2: PFD of ADU

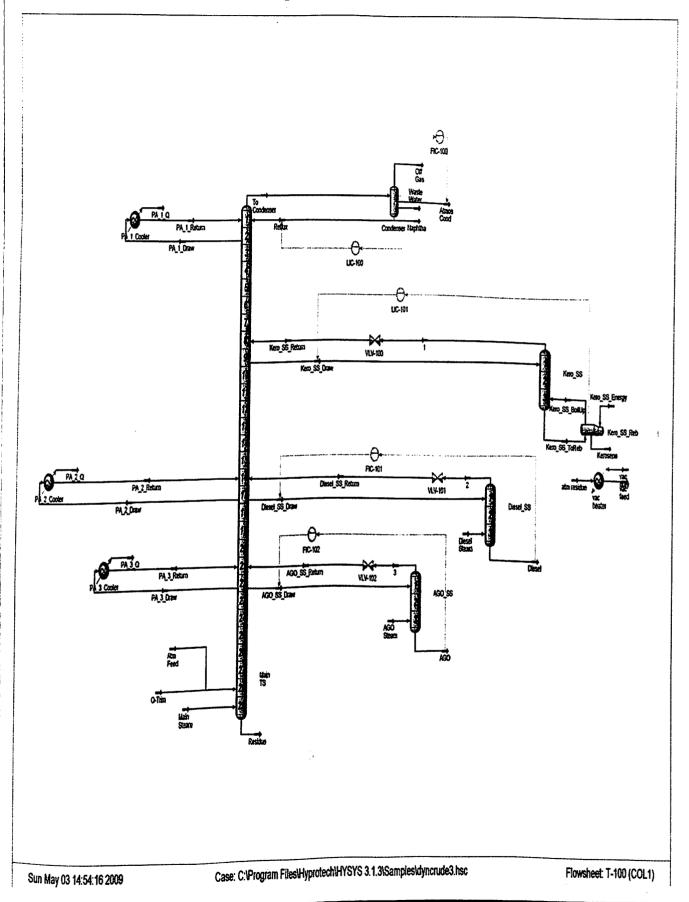




Table 5. 10: connection of ADU

П				To	se Name; C			· · · · · · · · · · · · · · · · · · ·		
\exists		TEAML	ND		an Marine: C	::Program Files	typrotech\HYSYS 3.1.3\	Samplesv	упстисез.п	sc
3	HYPROTEC	Calgary.	Alberta	Un	il Set: F	Told-USGPM			6. <u>1</u>	
4	Circavett innevat	CANAD	A	Da	te/Time:	un May 03 14:58:	57 2009			
5 6								*******		
\Box		Colu	mn S	ub-Flowshee	t: T-100	@Main				
8						17				
9 10				CO	NNECTIONS					
11	The second second				niet Stream					
12	STREAM	NAME	_ _	Stage			FROM UNIT OPE	RATION		
13	Main Steam		29	Main TS Main TS						
14 15	Q-Trim Atm Feed			Main TS		Valve				/LV-100
16	Kero_SS_Energy			ro_SS_Reb			* ***			
17	Diesel Steam			_Diesel_SS	*****					
18	AGO Steam		· ³ <u>-</u>	_AGO_SS			To the district of the second			
19 20	atm residue vac duty	***************************************								
21	Vac Coxy	and the second second			utlet Stream		2. 1. 1. 1. 1. 1. 1. 1. 1. 1. 1. 1. 1. 1.	a ingg	1017	
22	STREAM	NAME		Stage			TO UNIT OPER			
23	Residue		29							
24	Atmos Cond			ondenser ondenser						
25 26	Off Gas Waste Water			ondenser						
27	Naphtha		Co	ondenser						
28	Kerosene			ro_SS_Reb						
29	Diesel		Account to the second second	_Diesel_SS _AGO_SS						
30	AGO									
31 32	PA_1_Q PA_2_Q			12		****				
33	PA 3 Q		P/	_3						
33 34 35	vac feed									
35				!	MONITOR					Ì
38 37		学生的人		Specif	cetions Summary	udali (may hai)	:制品色类10.2000	2	建建	
37 38	Mark Control of the C	Specified	Value	Current Value	Wt. Error	W1, Tol.	Abs. Tol.	Active	Estimate	Used
39	Kero SS Prod Flow	271.2	USGPM .	271.2 USGPM		1.000e-002 *	4.403 USGPM *	On	On	On
40	Diesel SS Prod Flow		USGPM *	562.2 USGPM 131.3 USGPM		1,000e-002 *	4.403 USGPM * 4.403 USGPM *	On On	On On	On On
41	AGO SS Prod Flow		USGPM *	1458 USGPM		1,000e-002 *	4.403 USGPM *	On	On	On
42	PA 1 Rate(Pa)	-5,500e+0				1,000e-002 °	0.9478 Bluffer *	On	On	On
43 44	PA 1 Duty(Pa) PA 2 Rate(Pa)		USGPM *	874,8 USGPM		1.0006-002 *	4,403 USGPM *	On	On	On
45	PA 2 Duty(Pa)	-3.500e+0				1.000e-002 *	0.9478 Btu/hs * 4.403 USGPM *	On	On	On On
46	PA 3 Rate(Pa)		USGPM *	874.8 USGPM		1,000e-002 °	0.9478 Btu/hr *	On On	On On	On On
47	PA 3 Duly(Pa)	-3,500e+0	USGPM *	671,0 USGPM		1.000e-002 *	4,403 USGPM *	On	On	On On
48	Naphtha Prod Rate		USGPM *	92.19 USGPM		1.000e-002 °	4.403 USGPM *	On	On	On
49 50	Liquid Flow Kero Reb Duty	7.500e+0	06 Btu/hr *	7.500e+008 Btu/hr		1.0000-002	9.478 Btu/hr *	On	On	On
51	Vap Prod Flow	0.0000	puole/hr *	0.0000 ibmole/hr 76.37		1,000e-002 * 1,000e-002 *	2.205 (bmole/hr * 1.000e-002 *	On Off	On On	On Off
52	Reflux Ratio		1.000 *	70.01	epecc	1 110000 000			7 (3)	- VII
53					SPECS					
54			V. T.	Column Sp	octioation Param	COOK .			- 11 · 12 · 12	19. J.A. 1
55 56	Transfer Time				SS Prod Flow	,				
8 57				A Ottomote: Print	ary Lower Bo	ind	- Upper E	2.33		9.1
58	Fixed / Ranged:	Fixed		/ Andipoto.		with the same of t	Oppar E	NUINT.		
59	Stream:	Kerosene	Flow Ba	010.	_SS Prod Flox		1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1			
60								Ar grant or		
61 62	Fixed / Ranged:	Fixed	Primary	/ Alternate: Print			Upper B	cently.	o loset	
63	Hyprotech Lol	9		HVAX6	(3)1.3 (Build 46			ep = 100	Specified b	110 29
~1	Licensed to: TEAM LND								-hornen D	y user.



Table 5. 11: Specs of ADU

1				Case N	ame: C:\Program Files\Hyprotech\H	YSYS 3.1.3\Samples\dyncrude3.hsc
3	1	EAM LI algary.	ND Albeita	Unit Sel	: Field-USGPM	
5]	ANAD		Date/Ti	me: Sun May 03 14:58:57 2009	
3						
,]	olu	mn Sub-Flows	sheet:	T-100 @Main (contin	ued)
3	and the second of the second o				ation Parameters	
0					Prod Flow	<u>a (Magilia) and an Albandia and an Albandia and Albandia</u>
1					FIOU FIOW	
3		esel	Flow Basis:	Volume		1
4				AGU_SS	Prod Flow	
5		ixed	Primary / Alternate: Flow Basis:	Primary Volume	Lower Bound:	Upper Bound:
7	·	AGO	FROW DESIG.		304-(0-)	
8	=9				Rate(Pa)	
9		best	Primary / Alternate: Pumperound:	Primary PA 1	Lower Bound: Flow Besis: Votence	Upper Bound:
2	Spec Type: Flow!	·	- umpercuite.		704018	<u>L</u>
2					Outy(Pa)	
3		bed	Primary / Alternate:	Primary PA_1	Lower Bound:	Upper Bound:
5	Spec Type:	Duty	Pumparound:		Rate(Pa)	
3	<u> </u>					T
7		ixed	Primary / Alternate:	Primory PA_2	Lower Bound: — Flow Basis: Volume	Upper Bound:
	Spec Types Flow I	Rate	Pumparound:			
,					outy(Pa)	
L	1 mod 1 total and	bed	Primary / Alternate:	Primary PA 2	Lower Bound:	Upper Bound:
	Spec Type:	Duty_	Pumparound:)-4-(D-)	
1					Rate(Pa)	
3	FIXOU / PANIGOU.	tood	Primary / Alternate:	Primary PA 3	Lower Bound: — Flow Basis: Volume	Upper Bound:
	Spec Type: Flow F	Cates	Pumparound:		outy(Pa)	<u> </u>
1						Tarana da la companya
	FD000 / PC000f000	bex	Primary / Alternate:	Primery PA 3	Lower Bound:	Upper Bound:
	Spec Type:	Duty	Pumparound:		Proof Prop	
ł					Prod Rate	
Ì	Fixed / Planiged: Fi		Primary / Alternate:	Primary Volume	Lower Bound:	Upper Bound
Į	Stream: Napl	ntha	Flow Basis:			
I				Liquid		T
	Fixed / Ranged: F	tod	Primary / Alternate:	Primary	Lower Bound:	Upper Bound:
İ	Stage: 27 Main	15	Flow Busia:	Volume	Liquid Specification: Light	
Ì				Kero Re	PD Duty	I
ł	Fixed / Ranged: Fi		Primary / Alternate:	Primary	Lower Bound:	Upper Bound:
İ	Energy Stream: Kero SS En	187	*18			
ĺ				Vap Pro	od Flow	
ŀ	Fixed / Ranged: Fi	xed_	Primary / Alternate:	Primary	Lower Bound:	Upper Bound:
ŀ	Stage: Conder		Flow Basis:	Molar		
İ				Reflux	Ratio	
ļ	Sivial (Discourt: R	xod.	Primary / Alternate:	Primary	Lower Bound:	Upper Bound:
ŀ	Fixed / Ranged: FI Stage: Conden		Flow Basis:	Moter	Liquid Specification: Light	
t					_	
		mar attorio	and dentile Sandard H	YSYS VAIC	B(BH6 2827)	Page 201
	TRACES WHITE A VALLEY OF THE PARTY OF THE PA					* Specified by user.



Table 5. 12: profile of ADU

		Case Name;	C:\Program Files\	typrotech\HYSYS 3.1	.3\Samples\dyncrude3.hsc
HYPROTECH	TEAM LND Calgary, Alberta	Unit Set:	Field-USGPM		
	CANADA	Date/Time:	Sun May 03 14:58:	57 2009	
	Column Sub	-Flowsheet: T-10	0 @Main (d	continued)	
		SUBCOOLING	;		Application of the state of the
		Condenser			
Degrees of Subcooling			0.0000 *		
Subcool to		User Variable			
		OSCI VARIABLE	•		
		PROFILES			
	W. The British of the State of	General Paramete T-100 (COL1) Numb	rs er of Stages:		Constitution of the second
Sub-Flow Sheet:	49. 149 . 12	Profile Estimat			2 ATREM GROSSING NEW
		Temperature		Liquid	Net Vapour
		(F)		GPM)	(USGPM)
	Condense 1_Main T	* 1		486.7 2698	2.138e-0
	1_Main 13 2_Main T3		*******************	1475	118
	3_Main T	·· I	3	1527	21
	4_ Main T	354	*****	1538	22
	5_Main T		•• · · • • • • · · · · · · · · · · · ·	1529	
	6_Main T			1510	22
	7_Main TS	The second secon		1478	
	8_Main TS 9_Main TS	404		1032	21
and the second s	10 Main T			1002	18
	11_Main T	434		984.3	190
	12_Main T	444		967.9	19:
	13_Main TS	450		945.9	193
The same of the same and the sa	14 Main TS		***	910.0 839.9	19
	15_Main TS 16_Main TS	40.4		2080	18
water to the control of the control	17_Main T		9	477,6	20
	18_Main TS		8 .	394.4	20
	19_Main TS	200	3	347.9	19
	20_Main TS			309.4	18
	21_ Main TS			1486 451.9	183
	22_ Main TS			395.0	20
	23_Main TS 24_Main TS			343.8	
	25_Main T	0.40		297.0	19
	26 Main T	655		234.2	19
	27 Main TS	661		102.1	18
	28_Matn TS			1360 1283	17
	29 Main TS	40.7		379.2	92.
	1_ Kero_SS 2_ Kero_SS			395.3	108
	2_Kero_S	448		406.3	124
	Kero_SS_Rel	455		271.3	134
to a large desired and the second desired and	1_Diesel_SS	501		620.5	94.
	2 Diesel_SS			600.0	66,
	3 Diesel_SS	483		581.5	44.
The second section of the second section of the second section	1_AGO_SE			154.1 144.7	41.5 27.1
	2_AGO_SS	587	4	131,3	18,
	3 AGO SS	1745787313 (BUE	(Decision)		Page 3 on



Table 5. 13: operation summary of ADU

1 2				Case I	lame: C:\Progra	m Files/Hyprotech/HYSYS 3,1.3	Samplesidyncrude3.hsc		
3	HYPROTECH	TEAM LND Calgary, Albert	a	Unit Sc	ot Field-USC	3PM.			
4 5	CITTETOLE PORTATION	CANADA		Date/T	ime: Sun May	03 14:58:57 2009			
5							The second second		
1		Column	Sub-Flow	sheet:	T-100 @M:	ain (continued)			
4					والإرام والمراجع والمراجع والمراجع والمراجع والمراجع والمراجع والمراجع والمراجع والمراجع والمراجع والمراجع				
<u>}</u>			NBP[0]1124*	Stage E	fficiencies		<u>。我就是我的</u> 我说:"他是 我 做。		
i	Stages C 23 Main TS	Overall Efficiency 1,000	1.00	<u> </u>					
12	24_Main TS	1.000	1.00						
13	25_Main TS	1.000	1.00	0					
14	26_Main TS	1.000	1.00						
15	27_Main TS	1.000	1.00						
16	28 Main TS	1,000	1.000						
<u>'</u>	29_Main TS	1,000	1,000 1,000			Paramanan (1, 1, 1, 1, 1, 1, 1, 1, 1, 1, 1, 1, 1, 1			
18	1_Kero_SS	1,000	1.00						
19 20	2_Kero_SS 3_Kero_SS	1.000	1,00						
<u>.</u>	Kero_SS_Reb	1.000	1.00						
2	1_Diesel_SS	1,000	1.00						
23	2 Diesel SS	1.000	1.00	2					
24	3_Diesel_SS	1.000	1.00						
25	1_AGO_SS	1.000	1.00						
6	2_AGO_SS	1,000	1,000						
4	3 AGO SS	1,000	1.00						
8				SO	LVER				
9		Negraphs at Negra	Cohirma Solvin	g Algorithms.	HYSIM inside Out	Grant to the Market	. Bedeline weren e		
약		Solving Optio				Acceleration Parame			
1 2	Maximum Iterations:	Costing		10000	Accelerate K Value	& H Model Parameters:	Of		
#	Equilibrium Error Toleranc	e:		1.000e-05	4				
#	Heat/Spec Error Tolerance			5.000a-004	4				
5	Save Solutions as Initial Es			On On	-				
6	Super Critical Handling Mo	del:		Simple K Low	1				
7	Trace Level:			Off	 	Damping Paramete	rs		
빏	Init from Ideal Ks:	timate Generate	r Parameters		Azeotrope Check:				
익				Off	Fixed Damping Fac	ctor:	Of		
4	Iterative IEG (Good for Che	STRCAIS).							
1					<u> </u>				
4				SIDE ST	RIPPERS				
4						artificación a la company			
,			All Carlo	Siderantible	TE THIND CO.	Product Flow	Salar Maria Cara Cara Cara Cara Cara Cara Cara		
3		# Stages	Liquid Draw Stag	1	spour Return Stage	(ibmole/hr)	Reboiler Duty (Btu/ter)		
1			9 Ma			718.1	7.500e+006		
1	Kero_SS	3:	17 Ma	n TS		1135	7.0000*000		
1	Diesel SS	- 3.	22 M2			202.4			
4	AGO_SS	J			CTIFIERS				
4									
+				PUMP A	ROUNDS				
			en and a second of the				1.2. Car Courses was weeken		
:	tagen and the second	The Area	THE STATE OF THE S			Product Flow	Condenser Duty		
۲			v Stage	Retur	n Stage	(lbmole/hr)	Congenser Duty (Btu/hr)		
1			2 Main TS		1_Main TS		-6.600e+007		
_	PA_1		2 Main TS 17 Main TS		16 Main TS		-3.500e+007		
L	PA_2		22 Main TS		21_Main TS		-3.500a+00°		
L	PA_3		1_Kero_SS		8 Main TS		0.0000		
	PA_4				21 Main TS		0.000		
_	PA 5		1 AGO SS						



Table 5. 14: rating of ADU

1									
2	TEAM	LND		Case N	Name: C:\Prógr	am Files\Hyprotech\H\	/SYS 3.1.3\S	emples/dyncrude3.hsc	
3	HYPROTECH Calgar	y, Alberta		Unit Se	el: Field-US	GPM	1.		
4 5	CANAL	DA		Date/T	ime: Sun May	03 14:58:57 2009			
6				wsheet: T-100 @Main (continued)					
7 8	Colu	ımn s	SWOIT-QUE	ineet:	1-100 @M	ain (contin	ued)	• • • • • • • • • • • • • • • • • • • •	
9				Pump Around Summary					
10 11		Draw :	Stage	Return Stage Pro				Condensor Duty	
12	PA_6		Diesel_SS		16_Main TS	(inholomdi)		(Btu/hr) 0.0000	
13 14				VAP BY	PASSES				
15				- DA	TING				
16									
17 18				Tray S	Sections				
19	Tray Section		Main TS		Kero SS	Diesel	SS	AGO_SS	
20	Tray Diameter	(ft)	44.95		3.937	9.84		4.921	
21	Weir Height	(ft)	0.1640		0.1640	0.16	60 ·	0.1640	
22	Weir Length	(ft)	32.81	 -	5.577	8.85	8 ·	3.937	
23	Tray Space	(ft)	1,969_		1.640	1.64	• •	1.640	
24	Tray Volume	(83)	3123		19.97	124.	8	31.20	
25	Disable Heat Loss Calculations		No		No	No		No	
26	Heat Model		None		None	Non	0	None	
27	Rating Calculations		No		No	No		No	
28	Tray Hold Up	(fl3)	260.3		1.997	12.4	8	3.120	
29				Ves	ssels				
30			3		e consideration of	to the state of the state of	2000 F 12		
31	Vossel		Condense		Kero SS Reb	and a substitution			
32	Diameter		13.12		5.235				
33	Length	(11)	13.12		3.281				
34	Volume	(ft3)	1759		70.63				
35	Orientation		Horizonta	<u>'</u>	Vertical				
36	Vessel has a Boot		Yes		No No				
37	Boot Diameter		2.187	- i					
38	Boot Length	(ft)	4,374	$\overline{}$	35.35				
9	Hold Up	(11 3)	528.5						
삗			Other Eq	juipment ir	n Column Flowsh	eet		•	
17									
201	PA 1 Cooler		PA 2 Cooler			Cooler		VLV-100	
12	PA 1 Cooler VLV-101		PA 2 Cooler VLV-102			Cooler heater		VLV-100	
13 14				Pressur	vac re Profile	heater			
13 14 15	VLV-101				vac re Profile		Prosture		
3 4 5	V.V-101		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.33 0 pila	heater	Pressure		
13 14 15 7	VLV-101		VLV-102	Pressu 28.7	vac re Profile re (psia)	heater	Prosture		
3 5 6 7	VLV-101 Condenser 1 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.33 0 pila	heater	Prostour		
3 4 5 7 8	VLV-101 Condenser 1 Main TS 2 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Pressure		
3 5 7 8	VLV-101 Condenser 1 Main TS 2 Main TS 3 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Pressure		
3 4 5 7 8 9	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS	XX X X X	VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Pressure		
3 4 5 6 8 9	VLV-101 Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Prossure		
3 4 5 6 8 9 0	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Prosture		
3 4 5 6 8 9 0 1	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS 7 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Pressure		
3 4 5 7 8 9 0 1 1 2 3	Condenser 1		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Pressure		
3 4 5 6 7 8 9 0 1 1 2 3 4 8	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS 7 Main TS 8 Main TS 9 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Pressure		
3 4 4 5 6 7 8 8 9 9 0 1 1 2 3 4 4 7	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS 7 Main TS 8 Main TS 9 Main TS 10 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Proseur		
3 4 5 6 7 7 8 9 9 1 1 1 2 3 3 8 8 8	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS 7 Main TS 8 Main TS 9 Main TS 10 Main TS 11 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Prospur		
3 4 5 6 7 7 8 9 9 1 1 1 2 3 3 8 8 8	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS 7 Main TS 8 Main TS 9 Main TS 10 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.75 0 pila	heater	Prosput		
13 14 15 16 17 18 19 10 11 12 2 3 3 4 4 8 9	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS 7 Main TS 8 Main TS 9 Main TS 10 Main TS 11 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.33 0 pila	heater	Prossur		
13 14 15	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS 7 Main TS 8 Main TS 9 Main TS 10 Main TS 11 Main TS 12 Main TS 13 Main TS 14 Main TS		VLV-102	Pressu 28.7	vac re Profile re (pile) (1915) 2.33 0 pila	heater	Pressure		
3 4 5 5 6 6 7 7 8 8 9 9 9 9 9 9 9 9 9 9 9 9 9 9 9 9	Condenser 1 Main TS 2 Main TS 3 Main TS 4 Main TS 5 Main TS 6 Main TS 7 Main TS 8 Main TS 9 Main TS 10 Main TS 11 Main TS 12 Main TS 13 Main TS		VLV-102	28.77 37.70	vac re Profile re (pile) (1915) 2.33 0 pila	No affect	Pressure		



Table 5. 15: properties atm feed of ADU

			Case Name: C:	Program Files\Hyprotec	hiHYSYS 3,1,3\Samples\	dyncrude3.hsc
HYPROTECH	TEAM LND Calgary, Alberta	1	Unit Set: Fi	eld-USGPM		
	CANADA		Date/Time: St	m May 03 14:58:57 2009		
					:	
	Column	Sub-Flowsi	neet: T-100 (@Main (cont	inueaj	
16_Moin						
17_Main						
18_Main 19_Main						
20_Main						
21_Main				·		
22_Main						
23 <u>Main</u> 24 <u>Main</u>						
25_Main	TS .					
26_Main						
27_Main 28_Main			-			
28_Main 29_Main			30.97 psla	•		
1_Kero_					-	
2_Kero_	SS		29.58 psia			
3_Kero_			29.58 psia			
Kero_SS_					****	
1_ Diesel 2_ Diesel						
3_Diesel		and the second s	30.15 psla			
1_AGO						
2_AGO_			30.56 psia	•	***	
3_AGO	<u>ss</u>	Dr	essure Solving Optic	ons		
					1.000 * Max Press item	ations 10
Pressure Tolerance 1.00	00e-004 Press	sure Drop Tolerance	PROPERTIES			
			•		and the second second second	
			• Atm Food			
			Perties : Atm Feed	Liquid Phase	Aqueous Phase	
The state of the s		Overall	Vapour Phase 0.6972		Aqueous Phase 0.0000	
Vapour/Phase Fraction			Vapour Phase 0.6972 612.6	Liquid Phase 0.3028 612.6	Aqueous Phase 0.0000 612.8	
Vapour/Phase Fraction Temperature:	(F) (psia)	Overall 0.6972 612.6 30.92	Vapotir Phase 0.6972 612.6 30.92	Uquid Phase 0.3028 612.6 30.92	Aqueous Phase 0.0000 612.6 30.92	
Vapour/Phase Fraction Temperature: Pressure:	(f)	Overall 0.6972 612.6 30.92 6231	Vapour Phase 0.6972 612.6 30.92 4344	Liquid Phase 0.3028 612.6	Aqueous Phase 0.0000 612.8	
Vapour/Phase Fraction Temperature: Pressure: Motor Flow Mass Flow	(F) (psia) (bmolefit) (bhu)	Overall 0.6972 612.6 30.92 6231 1.285e+008	Vapotir Phase 0.6972 612.6 30.92	tiquid Phase 0,3028 812.6 30.92 1887	Aqueous Phase 0.0000 612.6 30.92 0.0000	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Std Ideal Lig Vol Flow	(F) (psia) (bmole/hr) (bhrir) (USGPM)	Overall 0.6972 612.6 30.92 6231 1.2850+008 2916	Vapour Phase 0.6972 612.6 30.92 4344 5.6430+005	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 0.0000 -1.278e+005	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Std Ideal Liq Vol Flow Molar Enthalpy	(F) (psia) (bmole/hr) (bhr) (USGPM) (Btu/(bmole)	Overall 0.6972 612.6 30.92 6231 1.285e+008	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005	Aqueous Phase 0.0000 612.8 30.92 0.0000 0.0000 0.0000 -1.278e+005 -819.6	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Std Ideal Liq Vol Flow Molar Esthalpy Mass Enthalpy	(F) (psia) (bmole/hi) (bhri) (USGPM) (Btu/bmole) (Btu/fb)	Overall 0.6972 612.6 30.92 6231 1.285e+006 2916 -1.198e+005 -581.0 162.9	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 0.0000 -1.278e+006 -619.6	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Std Ideal Liq Vol Flow Molar Enthalpy Mass Enthalpy Molar Entropy (E	(psia) (psia) (bmolefiri) (bfir) (bfir) (USGPM) (Btu/bmole) (Btu/bhole-F) (Btu/bF)	Overall 0.5972 612.6 30.92 6231 1.2850-008 2916 -1.1930+005 -581.0 162.9 0.7899	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005	Aqueous Phase 0.0000 612.8 30.92 0.0000 0.0000 0.0000 -1.278e+005 -819.6	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Std Ideal Liq Vol Flow Molar Entrapy Molar Entropy Mass Entrapy Heat Flow Heat Flow	(F) (psia) (bmolefir) (bhr) (bhr) (USGPM) (Bturbmole) (BturbholeF) (Bturb+F) (Bturb+F)	Overall 0.6972 612.6 30.92 6231 1.2856+008 2916 -1.1986+005 -581.0 162.9 0.7899 -7.4656+008	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.979e+008 2.785e-003	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9 0.7817 -4.487e+008 0.1174	Aqueous Phase 0.0000 612.8 30.92 0.0000 0.0000 -1.278+005 -619.8 162.9 0.7699 0.0000 0.1849	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Mase Flow Std Ideal Lig Vol Flow Molar Enthalpy Mose Enthalpy Mass Enthalpy Mass Enthalpy Heat Flow Molar Entropy Heat Flow Molar Density	(F) (psia) (bbmolefir) (bbfr) (bbfr) (USGPM) (Btu/bmole) (Btu/bb) (btu/bmole/F) (Btu/fb-F) (Btu/fb-F) (btu/fb-F) (btu/fb-F) (btu/fb-F)	Overall 0.6972 612.6 30.92 6231 1.2856+008 2916 -1.1986+005 -581.0 162.9 0.7899 -7.4656+008 3.966e+003 0.8157	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.978e+008 2.785e-003 0.3820	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9 0.7617 -4.487e+003 0.1174 44.83	Aqueous Phase 0.0000 612.8 30.92 0.0000 0.0000 -1.278e+005 -410.9 0.7899 0.0000 0.1849 34.01	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Stil Ideal Lig Vol Flow Motor Enthalpy Moss Enthalpy Molar Entropy Mass Entropy Mest Flow Molar Entropy Mass Entropy Molar Entropy Mass Entropy Molar Entropy Molar Entropy Molar Entropy Molar Entropy Molar Entropy Molar Entropy Molar Density Molar Density	(F) (psia) (bmolefir) (bhr) (bhr) (USGPM) (Bturbmole) (BturbholeF) (Bturb+F) (Bturb+F)	Overall 0.6972 612.6 30.92 6231 1.2856+008 2916 -1.1986+005 -581.0 162.9 0.7899 -7.4856+008 3.9560+003 0.8157 54.93	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8236 -2.979e+003 2.785e-003 0.3620 50.15	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9 0.7817 -4.487e+008 0.1174	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 -1.278e+005 -419.6 162.9 0.7899 0.0000 0.1849 34.01	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mass Flow Mass Flow Molar Entropy Mass Entrapy Molar Entropy Mass Entrapy Molar Entropy Heat Flow Molar Density Molar Density Mass Density Stil Ideal Lig Mass Density	(F) (psia) (bmole/hr) (USGPM) (Btu/bmole) (Btu/bmole) (Btu/bh) (Btu/bh) (Btu/hr) (Btu/hr) (bmole/f3) (bmole/f3) (bhf3)	Overall 0.6972 612.6 30.92 6231 1.2856+008 2916 -1.1936+005 -581.0 162.9 0.7899 -7.4656+008 3.9566-003 0.8157 54.93	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.978e+008 2.785e-003 0.3820	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9 0.7617 -4.487e+008 0.1174 44.83	Aqueous Phase 0.0000 612.8 30.92 0.0000 0.0000 -1.278e+005 -410.9 0.7899 0.0000 0.1849 34.01	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Std Ideal Liq Vol Flow Motor Enthalpy Mass Enthalpy Mose Entropy Entropy Heat Flow Motor Density Mass Density Std Ideal Liq Mass Density Liq Mass Density	(F) (psia) (bmole/hr) (b/hr) (USGPM) (Btu/bmole) (Btu/bmole-F) (Btu/hr) (Btu/hr) (bmole/f3) (bmole/f3) (b/f3) d (b/f3) Stu/hmole-F)	Overall 0.6972 612.6 30.92 6231 1.2856+008 2918 -1.1986+005 -581.0 162.9 0.7899 -7.4856+008 3.9666-003 0.8157 54.93 54.70	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.979e+003 2.786e-003 0.3620 60.15	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9 0.7617 -4.487e+008 0.1174 44.83 59.35 58.98 264.3 0.6921	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 -1.278e+005 -619.8 162.9 0.7899 0.0000 0.1649 34.01 54.93 54.70 151.7	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Mase Flow Molar Enthalpy Molar Enthalpy Molar Enthalpy Molar Enthalpy Molar Enthalpy Molar Enthalpy Molar Enthalpy Molar Entropy Heat Flow Molar Density Mass Density Liq Mass Density Liq Mass Density Liq Mass Density Molar Enthalpy Mass Density	(F) (psia) (bmole/hr) (USGPM) (Btu/bmole) (Btu/bmole) (Btu/bh) (Btu/bh) (Btu/hr) (Btu/hr) (bmole/f3) (bhf3) (bhf3) d (bhf3) stu/bmole-F) (Btu/br)	Overall 0.6972 612.6 30.92 6231 1.2856+008 2916 -1.1936+005 -581.0 162.9 0.7899 -7.4656+008 3.9566-003 0.8157 54.93 54.70 136.4 0.6618	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.978e+003 2.785e-003 0.3620 50.15 50.71 80.91 0.6228 2.042e-002	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9 0.7617 -4.487e+003 0.1174 44.83 59.35 55.98 264.3 0.6921 6.110e-002	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 0.0000 -1.278e+005 -619.6 162.9 0.7899 0.0000 0.1849 34.01 54.93 54.70 151.7 0.7356	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mase Flow Std Ideal Liq Vol Flow Molar Enthalpy Molar Enthalpy Molar Enthalpy Molar Enthalpy Molar Entropy Melet Flow Molar Density Mass Density Molar Density Molar Density Molar Density Molar Density Molar Density Molar Density Molar Density Molar Density Molar Density Molar Density Molar Density Molar Density Molar Molar Density Molar Molar Density Molar Molar Density Mass Density	(F) (psia) (bmole/hr) (USGPM) (Btu/bmole) (Btu/bmole) (Btu/bh) (Btu/bh) (Btu/bh) (Btu/hr) (bmole/f3) (bhft3) (bhft3) d (bhft3) stu/bmole-F) (Btu/br-F) (Btu/br-F) (Btu/br-F) (Btu/br-F)	Overall 0.6972 612.6 30.92 6231 1.2856+008 2918 -1.1986+005 -581.0 162.9 0.7899 -7.4856+008 3.9666-003 0.8157 54.93 54.70	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.978e+003 2.785e-003 0.3620 50.15 50.71 80.91	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -822.7 290.9 0.7617 -4.487e+008 0.1174 44.83 59.35 55.98 264.3 0.6921 6.110e-002	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 0.0000 -1.278+005 -619.6 162.9 0.7899 0.0000 0.1649 34.01 54.93 54.70 151.7 0.7356 2.070e-002	
Vapour/Phase Fraction Temperature: Pressure: Notar Flow Mase Flow Mase Flow Std Ideal Lig Vol Flow Motor Enthalpy Moss Enthalpy Motor Enthalpy Motor Enthalpy Motor Enthalpy Motor Enthalpy Lig Mass Density Mass Density Mass Density Mass Density Mass Density Mass Density Mass Density Mass Density Mass Density Mass Density Mass Density Mass Density Motor Heat Capacity Motor Heat Capacity Thermal Conductivity Viscosity	(F) (psia) (bmole/hr) (USGPM) (Btu/bmole) (Btu/b) (Btu/bh) (Btu/bh) (Btu/bh) (Btu/bh) (Btu/hr) (Btu/hr) (bmole/f3) (bhf3) (bhf3) d (bhf3) Stu/bmole-F) (Btu/hr-F) (Btu/hr-F) (Btu/hr-F) (Btu/hr-F)	Overall 0.6972 612.6 30.92 6231 1.2856-008 2916 -1.1936+005 -5.81.0 162.9 0.7899 -7.4656+008 3.9666-003 0.8157 54.93 54.70 136.4 0.6616	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.979e+003 2.785e-003 0.3620 50.15 50.71 80.91 0.6223 2.042e-002 1.155e-002	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9 0.7617 -4.487e+008 0.1174 44.83 59.35 58.98 264.3 0.6921 6.110e-002 0.3041	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 0.0000 -1.278e+006 -162.9 0.7899 0.0000 0.1649 34.01 54.93 54.70 151.7 0.7356 2.070e-002 0.1323	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mass Flow Std Ideal Liq Vol Flow Molar Enthalpy Moss Enthalpy Moss Enthalpy Moss Enthalpy Heat Flow Molar Entropy Liq Mass Density Std Ideal Liq Mass Density Liq Mass Density @Std Con Molar Heat Capacity Thermal Conductivity Viscosity Surface Tension	(F) (psia) (bmole/hr) (USGPM) (Btu/bmole) (Btu/bmole) (Btu/bh) (Btu/bh) (Btu/bh) (Btu/hr) (bmole/f3) (bhft3) (bhft3) d (bhft3) stu/bmole-F) (Btu/br-F) (Btu/br-F) (Btu/br-F) (Btu/br-F)	Overall 0.6972 612.6 30.92 6231 1.2856+008 2918 -1.1986+005 -581.0 162.9 0.7899 -7.4856+008 3.9666-003 0.8157 54.93 54.70 136.4 0.6616	Vapour Phase 0.6972 612.6 30.92 4344 5.643e+005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.978e+003 2.785e-003 0.3620 50.15 50.71 80.91 0.6228 2.042e-002	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -822.7 290.9 0.7617 -4.487e+008 0.1174 44.83 59.35 55.98 264.3 0.6921 6.110e-002	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 0.0000 -1.278+005 -619.6 162.9 0.7899 0.0000 0.1649 34.01 54.93 54.70 151.7 0.7356 2.070e-002	
Vapour/Phase Fraction Temperature: Pressure: Molar Flow Mass Flow Std Ideal Liq Vol Flow Motor Enthalpy Moss Enthalpy Molar Entropy (E) Mass Entropy Heat Flow Molar Density Mass Density Std Ideal Liq Mass Density Liq Mass Density GStd Con Molar Heat Capacity Thermal Conductivity Viscosity	(F) (psia) (bmole/hr) (USGPM) (Btu/bmole) (Btu/b) (Btu/bh) (Btu/bh) (Btu/bh) (Btu/bh) (Btu/hr) (Btu/hr) (bmole/f3) (bhf3) (bhf3) d (bhf3) Stu/bmole-F) (Btu/hr-F) (Btu/hr-F) (Btu/hr-F) (Btu/hr-F)	Overall 0.6972 612.6 30.92 6231 1.2856-008 2916 -1.1986+005 -5.81.0 162.9 0.7899 -7.4656+008 3.9666-003 0.8157 54.93 54.70 136.4 0.6618	Vapour Phase 0.6972 612.6 30.92 4344 5.843e-005 1403 -6.857e+004 -527.8 107.0 0.8238 -2.979e+008 2.796e-003 0.3620 50.15 50.71 80.91 0.6223 2.042e-002 1.155e-002 129.9	Uquid Phase 0.3028 612.6 30.92 1887 7.205e+005 1513 -2.378e+005 -622.7 290.9 0.7617 -4.487e+003 9.1174 44.83 59.35 58.98 264.3 0.6921 6.110e-002 0.3041 12.81 381.9	Aqueous Phase 0.0000 612.6 30.92 0.0000 0.0000 0.0000 -1.278e+005 -619.6 162.9 0.7899 0.0000 0.1649 34.01 54.93 54.70 151.7 0.7356 2.070s-002 0.1323 5.741	



Table 5. 16: properties of stream of ADU

_								
爿			Case Name:	C:\Program Files\Hyprot	echiHYSYS 3.1.3\Semple	s\dyncrudo3.hsc		
3	HYPROTECH Catgary, A		Unit Set:	Field-USGPM				
5	CANADA CANADA		Date/Time:	Sun May 03 14:58:57 20	Ģ9			
6								
7	Colum	in Sub-Flows	heet: T-100 @Main (continued)					
8 9		Pro	perties : Main Ste	àm .				
10		Overall	Vapour Phase	Liquid Phase	Aqueous Phase			
11	Vapour/Phase Fraction	1.0000	1.0000	0.0000	0.0000			
12	Temperature: (F)	375.0	375.0	375.0	375.0			
13	Pressure: (psia)	31.01	31.01	31.01	31.01			
14	Moter Flow (ibmole/hr)	416.3 7500	416.3 7500	0,0000	0,0000			
15 16	Mass Flow (fb/hr) Std Ideal Liq Vol Flow (USGPM)	15,01	15.01	0.0000	0.0000			
17	Molar Enthalpy (Btu/fbmole)	-1.012e+005	-1.012++005	-1.170e+006	0,0000 -1,170e+005			
18	Mass Enthalpy (Btu/fb)	-5619	-5619	-6493	-1.17067403			
19	Molar Entropy (Btu/bmole-F)	43.57	43.57	43.57	43,57			
20	Mass Entropy (Btu/lb-F)	2.418	2.418	2.418	2.418			
21	Heat Flow (Btu/hr)	-4.214e+007	-4.214e+007	0000.0	0.0000			
22	Molar Density (ibmole/ft3)	3.497e-003	3.4976-003	2.993	2.993			
23	Mass Density (lb/fi3)	6.300e-002	6.300e-002	53.93	53.93			
24	Std ideal Liq Mass Density (lb/fi3)	62.30	62.30	62.30	62.30			
25	Liq Mass Density @Std Cond (tb/fl3)	63.33	63.33	63.33	63.33			
26	Moter Heat Capacity (Btu/lbmole-F)	8.428	8.428 0.4679	20.75 1,152	20.75			
27	Mass Heat Capacity (Btu/lb-F)	0.4879 1,877e-002	1.877e-002	0.3878	1.152 0.3876			
28	Thermal Conductivity (Btu/tr-ft-F)	1,5720-002	1,5726-002	0.1407	0.1407			
쁴	Viscosity (cP)	1,5720-02	-	39.94	39.94			
<u>30</u>	Surface Tension (dyne/cm)	18.02	18,02	18.02	18.02			
31 32	Molecular Weight Z Factor	0.9899	0.9899	1.156e-003	1.1566-003			
32		Pro	perties : Diesel St	eam				
34		Overali	Vapour Phase	Liquid Phase	Aqueous Phase			
35	Vapour/Phase Fraction	1,0000	1.0000	0,000,0	0.0000			
36	Temperature: (F)	300.0	300.0	300.0	300.0			
37	Pressure: (psia)	30.62	30.62 166.5	30.62 0.0000	30,62 0,0000			
38	Motar Flow (ibmole/hr)	166.5 3000	3000	0.0000	0.0000			
39	Mass Flow (lb/hr)	6,003	8.003	0.0000	0.0000			
40	Std Ideal Liq Vol Flow (USGPM)	-1,019e+005	-1.019e+005	-1,185e+005	-1.185e+005			
41	Molar Enthalpy (Btu/Brnole)	-5854	-5654	-6577	-6577			
12	Mass Enthalpy (Sturib)	42.80	42.80	42.80	42.80			
ß	Molar Entropy (Btu/fbmole-F)	2,378	2.376	2.378	2.376			
44	Marsa Ermopy	-1,696e+007	-1.696e+007	0.000.0	0.0000			
45	Plost Flori	3.805003	3.805e-003	3.135	3.135			
46	MOUNT DOCKS	6,855e-002	6.855e-002	56.48	56.48			
17	Mass Density (units) Std Ideal Liq Mass Density (Ib/R3)	62.30	62.30	62.30	62.30			
	Liq Mess Density @Std Cond (ib/ft3)	63.33	63.33	63.33	63.33			
8	Motar Heat Capacity (Stuffbmole-F)	8.348	8.348	19.65	19.65			
<u>;</u>	Mass Heat Capacity (Btu/ib-F)	0,4634	0.4634 1.668e-002	0.3968	0.3968	······································		
52	Thermal Conductivity (Blufts-R-F)	1.668e-002	1.3996-002	0.1825	0.1825			
3	Viscosity (cP)	1,3996-002		48.91	48.91			
и	Surface Tension (dyne/cm)	18.02	18.02	18.02	18.02			
	Molecular Weight		0.9870	1,195e-003	1 1980 002			
Ø		Pro	perties AGO Ste	in. 2010 (8 1596)		mineral leaves of the leaves		
_	Z Factor		Manager Dhoma	Liquid Phase	Aqueous Phase			
8	Z Factor		Vapour Phase					
98 97	Contract Con	Overali	1,0000	0.0000	0.0000			
58 57 59	Vapour/Phase Fraction			0.0000 300.0	0.0000 300.0			
58 57 59 50	Vapour/Phase Fraction Tomperature: (F)	Overall 1,0000	1,0000 300.0 30,84	300.0 30.84	300.0 30.84			
58 57 59 50	Vapour/Phase Fraction	Overall 1,0000 300.0 30.84	1.0000 300.0 30.84	300.0 30.84	300.0 30.84			

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Table 5. 17: properties of stream of ADU

1			Case Name:	C:\Program Files\Hyprote	ch\HYSYS 3,1.3\Sample	s\dyncrude3.hsc	
3	HYPROTECH TEAM LND Calgary, Albe	rta	Unit Set:	Field-USGPM			
4	CANADA		Date/Time:	Sun May 03 14:58:57 200	9		
5							
7	Column	Sub-Flows	heet: T-100				
9		Pro	perties : AGO Stea	im .			
10		Overall	Vapour Phase	Liquid Phase	Aqueous Phase		
11	Mass Flow (lb/hr)	2500	2500	0.0000	0.0000		
2	Std Ideal Liq Vol Flow (USGPM)	5.003	5.003	0.0000	0.0000		
3	Molar Enthalpy (Btu/lbmole)	-1.019e+005	-1.019e+005	-1.185e+005	-1.185e+005		
4	Mass Enthalpy (Btu/lb)	-5654	-5654	-6577	-6577		
5	Molar Entropy (Btu/lbmole-F)	42.79	42.79	42.79	42.79		
6	Mass Entropy (Btu/lb-F)	2.375	2.375	2.375	2.375		
7	Heat Flow (Btu/hr)	-1.413e+007	-1.413e+007	0.0000	0.0000		
8	Molar Density (Ibmole/ft3)	3.833e-003	3.833e-003	3.135	3,135		
9	Mass Density (lb/ft3)	6,905e-002	6.905e-002	56.48	56.48		
0	Std Ideal Liq Mass Density (lb/ft3)	62.30	62.30	62.30	62.30		
1	Liq Mass Density @Std Cond (lb/ft3)	63.33	63.33	63.33	63.33		
2	Molar Heat Capacity (Btu/lbmole-F)	8.349	8,349 0,4635	19.65	19.65		
3	Mass Heat Capacity (Btu/lb-F)	0.4635	1.669e-002	0.3968	1.091		
4	Thermal Conductivity (Btu/hr-ft-F)	1.669e-002	1.399e-002	0.1825	0.3968 0.1825		
25	Viscosity (cP)	1.399e-002	1.3336-002	48.91	48.91		
6	Surface Tension (dyne/cm)	18.02	18.02	18.02	18.02		
7	Molecular Weight	0.9869	0.9869	1,207e-003	1.207e-003		
8	Z Factor		perties : Naphtha	es Alexandra	B. J. Prancisky was	NEW COLUMN SERVICE	
9		COLUMN TO THE OWNER OF THE OWNER OWN	Vapour Phase	Liquid Phase	Aqueous Phase	THE PERSON NAMED IN COLUMN	
0		Overall 0.0000	0,0000	1.0000	0.0000		
1	Vapour/Phase Fraction	114.0	114.0	114.0	114.0		
2	Temperature: (F)	19.70	19.70	19.70	19.70		
3	Pressure: (psia)	2759	0.0000	2759	0.0000		
4	Molar Flow (fbmole/ht)	2,464e+005	0.0000	2.464e+005	0.0000		
5	Mass Flow (lb/hr) Mass Flow (USGPM)	671.0	0.0000	670.9	0.0000		
6	Std ideal Liq Vol 1 iov	-8,366e+004	-5.354e+004	-8.365e+004	-1.220e+005		
7	Molar Enthalpy (Btu/lbh)	-936.9	-980.4	-936.7	-6773		
8	Mass Enthalpy (Btu/lbmole-F)	31.32	47.40	31.32	31.32		
9	Molar Entropy (Btu/lb/Hole-17)	0.3507	0.8679	0.3507	1.738		
0	Mass Entropy (Sturb-1) (Btu/hr)	-2.308e+008	0.0000	-2.308e+008	0.0000		
1	Heat Flow	0.4993	3.296e-003	0.4993	3.437		
2	Molar Density (Ib/ft3)	44,59	0.1800	44.59	61.91		
3	Mass Density	45.78	37.64	45.78	62.30		
4	Std Ideal Liq Mass Density	46.35	38.55	46.35	63.33	Maria La Companya da Companya da Companya da Companya da Companya da Companya da Companya da Companya da Compa	
5	Liq Mass Density @ord Cont	44.27	23.21	44.27	18.58	The contract of	
6	Motal Fleat Cuptory	0.4958	0.4250	0.4958	1.031	<u>U </u>	
7	Mass rieal Capacity	6.587e-002	1.028e-002	6.587e-002	0.3688		
8	Thermal Collddowny (cP)	0.3528	8.008e-003	0.3528	0.5880		
9	Viscosity	18.00	-	18.00	68.52		
0	Surface rension	89.30	54.61	89.30	18.02		
1	Molecular Weight	6.4076-003	0.9708	6.407e-003	9.309e-004	The state of the s	
2	Z Factor	Pro	perties : Residue	NEW YORK OF THE PARTY OF THE PA	Aguerus Dt	201	
3	white the state of	Overall	Vapour Phase	Liquid Phase	Aqueous Phase		
1		0.0000	0.0000	1.0000	0.0000		
3	Vapour/Phase Fraction (F)	667.1	667.1	667.1	667.1		
3	Temperature: (peia)	30.97	30.97	30.97	30.97		
4	Pressure:	1594	0.0000	1594	0.0000		
8	Molar Flow (ibmole/hr)	6.981e+005	0.0000	6.981e+005	0.0000		
9	Mass Flow (ib/hr)	1445	0.0000	1445	0.0000		
0	Std Ideal Liq Vol Flow (USGPM)	-2.564e+005	-9.948e+004	-2.564e+005	-9.948e+004		
-	Molar Enthalmy (Btu/lbmole)		-1385	-585.5	-1385		
	Molar Enthalpy (Btu/lbmole) Mass Enthalpy (Btu/lb)	-585.5	YSYS v3.1.3 (Build 4			Page 14 of 2	

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Table 5. 18: properties of stream of ADU

_			Case Name:	C:\Program Files\Hyprote	ANHVEVE 2 1 7 Comple	nidmanuda? haa
-	TEAM LND				CINTIO 10 3.1.30ampio	SWYINI (GGOOLISC
3	HYPROTECH Calgary, Alb	erta		Field-USGPM		
5	GALLON.		Date/Time:	Sun May 03 14:58:57 200	19	
6 7	Colum	n Sub-Flows	heet: T-100	@Main (con	tinued)	
8		Pro	perties : Residue	in the same of the same		
9 10		Overall	Vapour Phase	Liquid Phase	Aqueous Phase	
ii	Molar Entropy (Btu/ibmole-F)	352.2	352.2	352.2	352.2	
	much Ex	0.8043	4,905	0.8043	4,905	
12 13	mass Emilopy	-4.088e+008	0.0000	-4.088e+008	0.0000	
_	Heat Flow (Brum) Molar Density (ibmole/fi3)	0.1023	2.590e-003	0.1023	2.590e-003	
14	William Delivery	44.79	0.1860	44.79	0.1860	
15	Wazas Deliany	60.24	55.71	60.24	55.71	
16	SIG ROOM CIG IMMSS DOLLOW	60.17	58.35	60,17	58.35	
17	DQ MESS Desistry Quito Commission (5)	312.0	43.90	312.0	43.90	
18	MONEY FROM CONTRACTOR EN	0.7125	0.6115	0.7125	0.6115	
19	mass real Capacity	6.6236-002	2.3676-002	6.6230-002	0.1439	
20	Thermas Condocurry (cP)	0.2821	1.789e-002	0.2821	2.621e-003	
21	Viscosay	12.92		12.92	3.724	
22	Surface Tension (dyne/cm)	437.9	71.80	437.9	71.80	
23	Molecular Weight	2.504e-002	0.9889	2.504e-002	0.9889	
24	Z Factor	2,5048-002 1	perties : Kerosen	e	5 200 00 00 00 2000	对对外的执行 的。
25		Overall	Vapour Phase	Liquid Phase	Aqueous Phase	
26		0.0000	0.0000	1.0000	0.0000	
27	Vapour/Phase Fraction	459.6	459.6	459.8	459.6	
28	Temperature: (F)	30.00	30.00	30.00	30.00	
29	Pressure: (psla)	718.1	0.0000	718.1	0.0000	
30	Molar Flow ((bmole/fur)	1,140++005	0.0000	1.140e+005	0.0000	
31	Mass Flow (lb/hr)	271.2	0.0000	271.2	0.0000	
32	Std Ideal Liq Vol Flow (USGPM)	-1.163e+005	-9.277e+004	-1.163e+005	-1.163e+005	
33	Moler Enthalter (Btu/fbmole)	-732.9	-619.1	-732.9	-732.9	
34	Mars Fotholow (Btufb)	82,97	82.97	82.97	82.97	
35	Moter Entropy (Btu/ibmole-F)	0.5228	0.5537	0.5228	0.5228	
36	Mass Entropy (Btuffb-F)	-8.352e+007	0.0000	-8,352++007	0.0000	
37	Heat Flow (Blu/hr)		3.2960-003	0.2566	0.2573	
38	Molar Density (ibmole/ft3)	0.2568	0.4939	40.72	40.83	
39 39	((6/13)	40.72	51.93	52,39	52.39	
_	Mass Density Std Ideal Liq Mass Density (Ib/R3)	52.39	. 52.18	52,62	• 52.62	
49	Sto local Lip Mass South Cond (ID/R3)	52.62	81.66	103.2	103.2	
41	Liq Mass Density @Std Cond (Ib/ICS) Makes Meet Canacity (Sturibmole-F)	103.2	0.5450	0.6505	0.6505	
42	Rtuffs-F)	0.6505	1,4186-002	5,6826-002	5.682e-002	
43	Mass Hear Cupacity	5.6826-002	8,496e-003	0.1797	0.1800	
44	(cP)	0.1797		10.47	10.47	
45	Viscosity (Ame/cm)	10.47	149.8	158,7	158.7	
46	Suriace Terratori	158.7	0.9227	1,185e-002	1,1820-002	
47	Molecular Weight	1,185e-002	Diesel		STATES OF THE PARTY OF THE PART	A THE WHENVERS
49	Z Factor	S. C. J. L. L. L. L. L. L. L. L. L. L. L. L. L.	Vapour Phase	Liquid Phase	Aqueous Phase	
49	Z Factor	Overail	0.0000	1.0000	0.0000	
50		0.0000	485.9	485.9	485.9	
51	Vapous/Phase Fraction (F)	485.9	30.15	30.15	30.15	
52	Temperature: (osia)	30:15	0.0000	1135	0.0000	
53	Pressure:	1135	0.0000	2,4660+005	0.0000	·
54	Motas Pions (0-fbs)	2.468e+005	0.0000	562.2	0.0000	
55	Mass Flow (USGPM)	562.2	-1,040e+005	-1,553e+005		
58		-1.553e+005		-715.1	-1.145e+005	
57	Molar Enthalpy (Bru/Ibmole)	-715.1	-1369	130.4	-6357	
58	Mass Enthatpy (Btu/lb)	130.4	130.4		130.4	
59	Motar Entropy (Btu/fbmole-F)	0.6003	1.716	0.6003	7.237	
60	Mass Entropy (Btu/ib-F)	-1.764e+008	0.0000	-1.764e+008	0.0000	
-	Heat Flow (Btu/hr)	0.1987	3.078e-003	0.1987	2.221	1

Moler Density

Etyprotech Ltd.

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Table 5. 19: properties of stream of ADU

1 2			Case Name:	C:\Program Files\Hyprote	ech\HYSYS 3.1.3\Sampl	es/dyncrude3.hsc		
3	HYPROTECH Calgary, Aib	erta	Unit Set:	Field-USGPM				
	CANADA CANADA		Date/Time:	Sun May 03 14:58:57 200	19			
		The Arthur						
	Colum	n Sub-Flows	sheet: T-100 @Main (continued)					
		Pr	operties : Diesei					
2		Overall	Vapour Phase	Liquid Phase	Aqueous Phase			
1	Mass Density (fb/ft3)	43.15	0.2339	43.15	40.01			
	Std Ideal Liq Mass Density (ib/ft3)	54.70	54.68	54.70	62.30			
	Liq Mass Density @Std Cond (fb/ft3)	54.78	57.34	54.78	63.33			
	Molar Heat Capacity (Btu/ibmole-F)	140.7 0.6476	41.35 0.5442	140.7	24.20			
5	Mass Heat Capacity (Btu/lb-F) Thermal Conductivity (Btu/hr-B-F)	6.3450-002	1.777e-002	0.6476	1.343			
7	Viscosity (cP)	0.2180	1.3536-002	6.345e-002	0.3545			
,	Surface Tension (dyno/cm)	12.99	1,2000 000	0.2180	0,1061			
5	Molecular Weight	217.2	75.99	12.99 217,2	25.61	<u> </u>		
,	Z Factor	1,496e-002	0.9654	1.496e-002	18.02			
ı	网络美国新疆军员 化多克克特 经销售库	Pn	operties : AGO		1.338e-003	in manager to the		
2		Overall	Vapour Phase	Liquid Phase	Aqueous Phase	orange in great the first tiping field.		
	Vapour/Phase Fraction	0.0000	0.0000	1.0000	0.0000	 		
	Temperature: (F)	568.8	588.8	568.8	568,8			
4	Pressure: (paia)	30.56	30.56	30.56	30,58			
4	Molar Flow (fbmole/fir)	202.4	0.0000	202.4	0.0000			
4	Mass Flow (Ib/hr)	6.002e+004	0.0000	6.002e+004	0.0000			
4	Std Ideal Liq Vol Flow (USGPM)	131.3	0.0000	131.3	0.0000			
4	Molar Enthalpy (Btu/Bmole)	-1.951e+005	-1.056e+005	-1.951e+005	-1.123e+005			
1	Mass Enthatpy (Btu/lb) Motor Entropy (Btu/lbmole-F)	-657.9	-2017	-657.9	-6238			
	Mass Entropy (Btu/lb-F)	304.5 1,027	304.5 5,814	304,5	304.5			
ì	Heat Flow (Bts/hr)	-3.948e+007	0.0000	1.027	16.90			
	Molar Density (lbmole/fi3)	0,1471	2.8206-003	-3.948e+007	0,0000			
ì	Mass Density (Ib/R3)	43,61	0.1477	0.1471 43.61	1.873			
1	Std Ideal Liq Mass Donsity (Ib/R3)	56.97	57.62	56.97	33.75 62.30			
1	Lig Mass Density @Std Cond (tb/ft3)	56.88	60.95	56.88	63.33			
ı	Motar Heat Capacity (Btu/ibmole-F)	. 201.3	29,49	201.3	35.69			
ı	Mass Heat Capacity (Btu/lb-F)	0,6788	0.5830	0.6788	1.981			
1	Thermal Conductivity (Btu/hr-ft-F)	6,532e-002	2.191e-002	6,532e-002	0.3144			
l	Viscosity (cP)	0.1292	1.7766-002	0.1292	9.0610-002			
l	Surface Tension (dyne/cm)	13.15		13.15	14,68			
I	Molecular Weight	296.5	52.38	298.5	18,02			
Į	Z Factor	1,882e-002	0.9817	1,882e-002	1.478e-003			
l		LA LA PARTIE DE LA	perties : Off Gas					
ŀ		Overag	Anhort Linize	Liquid Phase	Aqueous Phase			
ŀ	Vapour/Phase Fraction	1,0000	1.0000	0,0000	0.0000			
ł	Temperature: (F)	114.0	114.0	114.0	114.0			
ł	Pressure: (psss) Melor Flow ((brook/hr)	7.308	19.70 7.308	0.0000	19.70			
ł	Molar Flow ((bmolefur) Mass Flow ((b/tur)	399.4	399,1	0.0000	0.0000			
t	Std Ideal Liq Vol Flow (USGPM)	1.322	1.322	0.0000	0.0000			
t	Motar Enthalpy (Btu/lbmole)	-5.358e+004	-5.354e+004	-8.365e+004	-1.220e+005			
ľ	Mass Enthalpy (Stuffb)	-981.2	-980.4	-938.7				
ľ	Motar Entropy (Stuffemole-F)	48.94	48.94	28.92	-6773			
Ī	Mass Entropy (Stuftb-F)	0.8596	0.8596	0.3239	48.94 2.606			
ſ	Heat Flow (Btu/hr)	-3.916e+005	-3.9136+005	0.0000	0.0000			
l	Molar Density (tbmole/ft3)	3.2966-003	3.296e-003	0.4993	3.437			
I	Mass Density (b/ft3)	0.1800	0.1800	44.59	51.91			
L	Std Ideal Liq Mass Density (Ib/R3)	37.64	37.64	46.78	62.30			
Ļ	Liq Mass Density @Std Cond (Ib/R3)	38.55	38.55	48.35	63.33			
Ļ	Molar Heat Capacity (Btu/ibmole-F)	23.21	23.21	44.27	18.58			
ı	Hyperdech (bit)	H	YSYS v3/1.9 (Birld 2	62 <i>(</i>)		Page 18 of 2		



Table 5. 20: properties of stream of ADU

2				Case Name:	C:\Program Files\Hypro	olech\HYSYS 3,1,3\Sam	ples\dyncrude3.hsc	
3	HYPROTECI	TEAM LND Calgary, Alberta		Unit Set:	Field-USGPM			
;	Circevette incovere	CANADA		Date/Time:	Sun May 03 14:58:57 2	009		
ŝ	ay i Sufferient, and		0.00 April 10 (10 April 10 Apr					
		Column	Sub-Flows	heet: T-100 @Main (continued)				
9	Section Action		DATE OF THE PARTY	perties : Off Gas	ANTO THE STATE OF			
0	SECURITY OF THE SECURITY OF TH	CAURIO DE VIDE SE SE MOS	Overall	Vapour Phase	Liquid Phase	Amuseus Phase		
11	Mass Heat Capacity	(Btu/lb-F)	0.4250	0.4250	0.4958	Aqueous Phase		
2	Thermal Conductivity	(Btu/hr-ft-F)	1.028e-002	1.028e-002	6.587e-002	0.3688		
13	Viscosity	(cP)	8.007e-003	8.007e-003	0.3528	0.5880		
14	Surface Tension	(dyne/cm)			18.00	68.52	?	
5	Molecular Weight	15.0	54.61	54.61	89.30	18.02	2	
7	Z Factor	Water Company	0.9708 Pro	0.9708 perties : Waste W	6.407e-003	9.309e-004		
8		THE PARTY OF THE P	Overall	Vapour Phase	Liquid Phase	E. Manager 1, 1, 1, 1, 1, 1, 1, 1, 1, 1, 1, 1, 1,		
19	Vapour/Phase Fraction		0.0000	0.0000	0.0000	Aqueous Phase		
20	Temperature:	(F)	114.0	114.0	114.0	1.0000		
21	Pressure:	(psia)	19.70	19.70	19.70	114.0		
22	Molar Flow	(lbmole/hr)	698.9	0.0000	0.0000	698.5		
23	Mass Flow	(lb/hr)	1.259e+004	0.0000	0.0000	1.259e+004		
24	Std Ideal Liq Vol Flow	(USGPM)	25.19	0.0000	0.0000	25.19	9	
25	Molar Enthalpy	(Btu/Ibmole)	-1.220e+005	-5.354e+004	-8.365e+004	-1.220e+005		
26	Mass Enthalpy	(Btu/lb)	-6773 14.69	-980.4 14.69	-936.7	-6773		
27	Molar Entropy Mass Entropy	(Btu/lbmole-F) (Btu/lb-F)	0.8154	0.2690	14.69 0.1645	14.69		
9	Heat Flow	(Btu/hr)	-8.527e+007	0.0000	0.0000	-8.527e+007		
10	Molar Density	(lbmole/ft3)	3.437	3.296e-003	0.4993	3.43		
31	Mass Density	(lb/ft3)	61.91	0.1800	44.59	61.91		
32	Std Ideal Liq Mass Densit	y (lb/ft3)	62.30	37.64	45.78	62.30		
33	Liq Mass Density @Std C	ond (lb/ft3)	63.33	38.55	46.35	63.33	3	
4	Molar Heat Capacity	(Btu/lbmole-F)	18.58	23.21	44.27	18.58		
5	Mass Heat Capacity	(Btu/lb-F)	1.031	0.4250 1.028e-002	0,4958	1.03		
6	Thermal Conductivity	(Btu/hr-ft-F)	0,3688	8.008e-003	6.587e-002 0.3528	0.3688		
37 18	Viscosity Surface Tension	(dyne/cm)	68,52		18.00	0.5880		
9	Molecular Weight	1	18.02	54.61	89.30	18.00		
0	Z Factor	Land	9,309e-004	0.9708	6,407e-003	9.309e-004		
1				SUMMARY	7/21/20/20/20/20/20/20/20/20/20/20/20/20/20/			
2	A STATE OF THE STA	42.0446	ATT OF THE PERSON			and the same of		
3	Flow Basis:			Molar Control III	The corr	position option is select	led	
4			Main S	Feed Compositio	The state of the state of the state of		THE REPORT OF THE PARTY OF THE	
5	15 35 7 10 10 10 10 10 10	Atm Feed 6.230685e+03	416.3		esel Steam 166.5270	AGO Steam	(2) TARGET OF 1 (2)	
6	Flow Rate (lbmole/hr)	6.2300036103	410.5		100.5270	138.7725		
7	Methane	0.0003	0.00	00	0.0000	0.0000		
9	Ethane	0.0006	0.00	Printed Street, Street	0.0000	0.0000		
ō	Propane	0.0086	0.00	00	0.0000	0.0000	******************************	
ĭ	j-Butane	0.0054	0,00	*****************	0.0000	0.0000		
2	n-Butane	0.0193	0.00		0.0000	0.0000		
3	H2O	0.0000	1,00	The second secon	1.0000	1.0000		
4	NBP[0]49*	0.0364	0.00		0.0000	0.0000		
3	NBP[0]79*	0.0436	0.00	Secretary and the second secon	0.0000	0.0000		
3	NBP[0]111*	0.0427 0.0416	0.00	and the second s	0.0000	0.0000		
	NBP[0]144*	0.0418	0.00		0.0000	0.0000		
-	NBP[0]176*	0.0451	0.00	(alternative control of the control	0.0000	0.0000	-	
		0.0437	0.00		0.0000	0.0000		
8	NBP[0]208*				- and the last of	0.0000		
7 8 9 0	NBP[0]240*		0.00	00	0.0000	0.0000		
8 9	Control of the Contro	0.0418	0.00	the state of the s	0.0000	0.0000		



Table 5. 21: feed composition of ADU

1			Case Na	me: C:\Program Files	Hyprotech\HYSYS 3.1.3\S	amples\dyncrude3.hsc
3	HYPROTECH	TEAM LND Calgary, Alberta	Unit Set:	Field-USGPM		
-	CIPECTE INNOVATION	CANADA	Date/Tin	ne: Sun May 03 14:58	3:57 2009	
1	CT SUPURSION SEVER	10.10.000.000.000.000	A DOMESTIC STORY OF THE STORY O			The second second second
		Column Sub	-Flowsheet:	T-100 @Main (continued)	
3						
0			SUMM	MARY		
1	TO SASTA MERCENTAGE T	Atm Feed	Main Steam	Diesel Steam	AGO Steam	Carplety, Paris
2	NBP[0]336*	0.0362	0.0000	0.0000	0.0000	
3	NBP[0]368*	0.0342	0.0000	0.0000	0.0000	
4	NBP[0]400*	0.0334	0.0000	0.0000	0.0000	
5	NBP[0]433*	0.0356	0.0000	0.0000	0.0000	
6	NBP[0]464*	0.0394	0.0000	0.0000	0.0000	
7	NBP[0]496*	0.0393	0.0000	0.0000	0.0000	-
9	NBP[0]528*	0.0356 0.0314	0.0000	0.0000	0.0000	
9	NBP[0]560* NBP[0]592*	0.0314	0.0000	0.0000	0.0000	
1	NBP[0]624*	0.0279	0.0000	0.0000	0.0000	-4
2	NBP[0]656*	0.0213	0.0000	0.0000	0.0000	
3	NBP[0]688*	0.0192	0.0000	0.0000	0.0000	
4	NBP[0]720*	0.0172	0.0000	0.0000	0.0000	
5	NBP[0]752*	0.0153	0.0000	0.0000	0.0000	
6	NBP[0]784*	0.0138	0.0000	0.0000	0.0000	
7	NBP[0]830*	0.0229	0.0000	0.0000	0.0000	
8	NBP[0]888*	0.0212	0.0000	0.0000	0.0000	
9	NBP[0]947*	0.0209	0.0000	0.0000	0.0000	* 10 mm days are a second and a
0	NBP[0]1009*	0.0267	0.0000	0.0000	0.0000	*****
1	COLUMN TO SELECT THE PROPERTY OF THE PROPERTY	0.0364	0.0000			
	NBP[0]1062°	0,0304	0.0000	0.0000	0.0000	
2	NBP[0]1124*	0.0358	0.0000	0.0000	0.0000	
2	And who has been been a selected and company on the select transfer and the		0.0000 Molar	0.0000 Th		lected
2 3 4	NBP[0]1124*	0.0358	0.0000 Molar Feed	0.0000 Th	0.0000 e composition option is se	lected
2 3 4 5	NBP[0]1124* Flow Basis:	0.0358 Atm Feed	0,0000 Molar Feed Main Steam	0.0000 Th Flows Diesel Steam	0.0000 e composition option is se	lected
2 3 4 5	NBP[0]1124*	0.0358	0,0000 Molar Feed Main Steam 416.3174	0.0000 Th	0.0000 e composition option is se	lected
2 3 4 5	NBP[0]1124* Flow Basis: Flow Rate (lbmole/hr)	0.0358 Atm Feed 6.230685e+03	0.0000 Molar Feed Main Steam 416.3174	0.0000 Th Flows Diesel Steam 166.5270	0.0000 e composition option is se AGO Steam 138.7725 —	lected
2 3 4 5 6 7	NBP[0]1124* Flov/ Basis: Flow Rate (lbmole/hr) Methane (lbmole/hr)	0.0358 Atm Feed 6.230685e+03 1.7716	0.0000 Molar Feed Main Steam 416.3174 0.0000	0.0000 Th Flows Diesel Steam 166.5270 — 0.0000	0.0000 e composition option is se AGO Steam 138.7725 0.0000	lected
2 3 4 5 6 7 8	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 1.7716 3.8868	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 — 0.0000 0.0000	0.0000 e composition option is se AGO Steam 138.7725 0.0000 0.0000	lected
2 3 4 5 7 8 9	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000	0.0000 e composition option is se AGO Steam 138.7725 0.0000 0.0000 0.0000	lected
2 3 4 5 6 7 8	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Eihane (ibmole/hr) Propane (ibmole/hr) i-Butane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is se AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000	lected
2 3 4 5 6 7 8 9	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Eihane (ibmole/hr) Propane (ibmole/hr) i-Butane (ibmole/hr) n-Butane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 1.7716 3.8868 53.6976 33.8880 120.1634	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is se AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000	lected
2 3 4 5 6 7 8 9	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr) i-Butane (ibmole/hr) n-Butane (ibmole/hr) H20 (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is se AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725	lected
2 3 4 5 6 7 8 9 0 1	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr) i-Butane (ibmole/hr) n-Butane (ibmole/hr) H20 (ibmole/hr) NBP[0]49* (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 416.3174	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 166.5270	0.0000 e composition option is se AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000	lected
2 3 4 5 6 7 8 9 0 1	NBP[0]1124* Flovr Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000	0.0000 e composition option is se AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725	lected
2 3 4 5 6 7 8 9 9 9 9 9 9 9 9 9 9 9 9 9 9 9 9 9 9	NBP[0]1124* Flovr Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000 0.0000	0.0000 e composition option is see AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000	lected
2 3 4 5 6 7 8 9 0 1 2 3 4 7	NBP[0]1124* Flovr Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is see AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000	lected
2 3 4 5 6 7 8 9 0 1 1 5 6 7 8	NBP[0]1124* Flovr Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr) n-Butane (ibmole/hr) h2C (ibmole/hr) NBP[0]49* (ibmole/hr) NBP[0]79* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]115* (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is set AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000	lected
22 33 34 44 55 56 67 77 77 77	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr) i-Butane (ibmole/hr) n-Butane (ibmole/hr) H2O (ibmole/hr) NBP[0]19* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]114* (ibmole/hr) NBP[0]176* (ibmole/hr) NBP[0]176* (ibmole/hr) NBP[0]176* (ibmole/hr)	0.0358 Aim Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 The Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is se AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected
2 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Eihane (ibmole/hr) Propane (ibmole/hr) I-Butane 0.0358 Atm Feed 6.230685e+03 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 Th Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is set AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected	
22 33 34 44 55 55 63 63 63 63 63 63 63 63 63 63 63 63 63	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 The Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is sei AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected
2 3 3 3 4 4 4 4 4 4 4 7 7 7 7 7 7 7 7 7 7	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr) i-Butane (ibmole/hr) i-Butane (ibmole/hr) n-Butane (ibmole/hr) NBP[0]149* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]114* (ibmole/hr) NBP[0]144* (ibmole/hr) NBP[0]145* (ibmole/hr) NBP[0]146* (ibmole/hr) NBP[0]240* (ibmole/hr) NBP[0]272* (ibmole/hr) NBP[0]272* (ibmole/hr) NBP[0]272* (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778 260.4584	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 The Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is sei AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected
22 33 34 44 44 55 55 55 66 67 77 77 77 77 77 77 77 77 77 77 77	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Elhane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 The Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 0.0000 166.5270 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0.0000 e composition option is see AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected
2 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr) i-Butane (ibmole/hr) i-Butane (ibmole/hr) n-Butane (ibmole/hr) NBP[0]149* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]114* (ibmole/hr) NBP[0]144* (ibmole/hr) NBP[0]145* (ibmole/hr) NBP[0]146* (ibmole/hr) NBP[0]240* (ibmole/hr) NBP[0]272* (ibmole/hr) NBP[0]272* (ibmole/hr) NBP[0]272* (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778 260.4584 243.5817 225.5531	0.0000 Molar Feed	0.0000 The Flows Diesel Steam 166.5270 0.0000	0.0000 e composition option is see AGO Steam 138.7725 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected
2 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778 260.4584 243.5817 225.5531 213.0128	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000	0.0000 The Flows Diesel Steam 166.5270 0.0000	0.0000 e composition option is see AGO Steam 138.7725 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected
2 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778 260.4584 243.5817 225.5531 213.0128 208.0350	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000	0.0000 The Flows Diesel Steam 166.5270	0.0000 e composition option is set AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected
2 3 3 4 4 4 4 7 7 7 7 7 7 7 7 7 7 7 7 7 7	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr) Propane (ibmole/hr) i-Butane (ibmole/hr) n-Butane (ibmole/hr) n-Butane (ibmole/hr) NBP[0]49* (ibmole/hr) NBP[0]49* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]114* (ibmole/hr) NBP[0]115* (ibmole/hr) NBP[0]203* (ibmole/hr) NBP[0]203* (ibmole/hr) NBP[0]304* (ibmole/hr) NBP[0]305* (ibmole/hr) NBP[0]368* (ibmole/hr) NBP[0]368* (ibmole/hr) NBP[0]368* (ibmole/hr) NBP[0]368* (ibmole/hr) NBP[0]368* (ibmole/hr) NBP[0]368* (ibmole/hr) NBP[0]368* (ibmole/hr) NBP[0]368* (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778 260.4584 243.5817 225.5531 213.0128 208.0350 222.1032	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000	0.0000 The Flows Diesel Steam 166.5270 0.0000 0.0000 0.0000 166.5270 0.0000	0.0000 e composition option is set AGO Steam 138.7725 0.0000	lected
2 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Ethane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778 260.4584 243.5817 225.5531 213.0128 208.0350 222.1032 245.5658	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000	0.0000 The Flows Diesel Steam 166.5270 0.0000	0.0000 e composition option is set AGO Steam 138.7725 0.0000	lected
2 3 4 5 6 7 8 9 0 1 2 3 4 5 6 7 8	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Eihane (ibmole/hr) Propane (ibmole/hr) Propane (ibmole/hr) I-Butane (ibmole/hr) n-Butane (ibmole/hr) NBP[0]49* (ibmole/hr) NBP[0]79* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]111* (ibmole/hr) NBP[0]114* (ibmole/hr) NBP[0]208* (ibmole/hr) NBP[0]208* (ibmole/hr) NBP[0]304* (ibmole/hr) NBP[0]306* (ibmole/hr) NBP[0]306* (ibmole/hr) NBP[0]306* (ibmole/hr) NBP[0]368* (ibmole/hr) NBP[0]468* (ibmole/hr) NBP[0]433* (ibmole/hr) NBP[0]436* (ibmole/hr) NBP[0]436* (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778 260.4584 243.5817 225.5531 213.0128 208.0350 222.1032 245.5658 244.7402	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000	0.0000 The Flows Diesel Steam 166.5270	0.0000 e composition option is set AGO Steam 138.7725 0.0000 0.0000 0.0000 0.0000 138.7725 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	lected
2 3 3 3 4 4 4 5 5 6 8 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	NBP[0]1124* Flow Basis: Flow Rate (ibmole/hr) Methane (ibmole/hr) Eihane (ibmole/hr) Propane (ibmole/hr) Propane (ibmole/hr)	0.0358 Atm Feed 6.230685e+03 — 1.7716 3.8868 53.6976 33.8880 120.1634 0.0000 226.6922 271.5848 266.1628 259.2997 273.0171 280.7032 272.5778 260.4584 243.5817 225.5531 213.0128 208.0350 222.1032 245.5658 244.7402 221.9375	0.0000 Molar Feed Main Steam 416.3174 0.0000 0.0000 0.0000 0.0000 416.3174 0.0000	0.0000 The Flows Diesel Steam 166.5270 0.0000	0.0000 e composition option is set AGO Steam 138.7725 0.0000	lected



Table 5. 22: product composition of ADU

+			Case	Name: C:\Program Files\	Hyprotech\HYSYS 3.1.3\Samp	ples/dyncrude3.hsc
	HYPROTECH	TEAM LND Calgary, Alberta	Unit S	et: Field-USGPM		
Ц	CIPICYCLE INNOVATION	CANADA	Date/T	ime: Sun May 03 14:58	:57 2009	
5	7 C. 10 C. 1	VEHAL TO SEE THE SECOND	The second second second			
		Column Sub	-Flowsheet:	T-100 @Main (continued)	
9			SUM	MARY		
10	the transfer of the first			- Parallella de la company		
11	100000000000000000000000000000000000000	Atm Feed	Main Steam 0.0000	Diesel Steam 0.0000	AGO Steam	国际。但首组的大场社会
12	NBP[0]656* (lbmole/hr)	132.5358 119.3544	0.0000	0.0000	0.0000	The first of the same and the same of the
14	NBP[0]688* (Ibmole/hr)	107.1919	0.0000	0.0000	0.0000	
15	NBP[0]720* (lbmole/hr) NBP[0]752* (lbmole/hr)	95.1681	0.0000	0.0000	0000.0	
16	NBP[0]784* (lbmole/hr)	86.1128	0.0000	0.0000	0.0000	
17	NBP[0]830* (lbmole/hr)	142.9682	0.0000	0.0000	0.0000	
18	NBP[0]888* (lbmole/hr)	131.8298	0.0000	0.0000	0.0000	
19	NBP[0]947* (lbmole/hr)	130,3381	0.0000	0.0000	0.0000	
20	NBP[0]1009* (lbmole/hr)	166.3298	0.0000	0.0000	0.0000	
21	NBP[0]1062* (lbmole/hr)	227.1031	0.0000	0.0000	0.0000	*************
22	NBP[0]1124* (lbmole/hr)	223.1021	0.0000	0.0000	0.0000	
23				oducts		
24	Flow Basis:		Molar Product (Compositions	e composition option is select	ed
25		0#0	Naphtha	Waste Water	Doub	Market British
26		Off Gas	2,758898e+03	698.8582	Residue 1.594396e+03	Kerosene
27	Flow Rate (lbmole/hr)	7.3082	2,7503900103	698.8582	1.594396e+03	718.0960
28		0.0512	0.0003	0.0000	0.0000	
29	Methane	0.0512 0.0347	0.0010	0.0000	0.0000	0.0000
30	Ethane	0.0347	0.0173	0.0000	0.0000	0.0000
31	Propane	0.0517	0.0117	0.0000	0.0000	0.0000
_	i-Butane n-Butane	0.1332	0.0419	0.0000	0.0000	0.0000
33 34	n-Butane H2O	0.0713	0.0012	1.0000	0.0067	0.0000
35	NBP[0]49*	0.1775	0.080.0	0.0000	0.0001	0.0000
35 36	NBP[0]79*	0.1350	0.0968	0.0000	0.0001	0.0000
37	NBP[0]111*	0.0758	0.0956	0.0000	0.0001	0.0000
38	NBP[0]144*	0.0414	0.0935	0.0000	0.0002	0.0000
39	NBP[0]176*	0.0240	0.0987	0.0000	0.0002	0.0000
40	NBP[0]208*	0,0133	0.1016	0.0000	0,0003	0.0001
41	NBP[0]240*	0.0067	0.0985	0.0000	0,0004	0.0008
42	NBP[0]272*	0.0032	0.0929	0.0000	0.0006	0.0055
43	NBP[0]304*	0.0014	0.0829	0.0000	0.0008	0.0149
44	NBP[0]336*	0.0005	0.0655	0.0000	0.0011	0.0475
45	NBP[0]368*	0.0001	0.0197	0.0000	0.0016	0.1970
46	NBP[0]400*	0.0000	0.0008	0.0000	0.0022	0.2533
47	NBP[0]433*	0.0000	0.0000	0.0000	0.0036	0,2452
48	NBP[0]464*	0.0000	0.0000	0.0000	0.0058	0.1893
49	NBP[0]496*	0.0000	0.0000	0,0000	0.0085	0.0424
50	NBP[0]528*	0.0000	0.0000	0.0000	0.0113	0.0023
51	NBP[0]560*	0.0000	0.0000	0.0000	0.0144	0.0001
52	NBP[0]592*	0.0000	0.0000	0.0000	0.0182	0.0000
53	NBP[0]624*	0.0000	0.0000	0.0000	0.0221	0.0000
54	NBP[0]656*	0.0000	0.0000	0.0000	0.0265	0.0000
55	NBP[0]688*	0.0000	0.0000	0.0000	0.0316	0.0000
56	NBP[0]720*	0,0000	0.0000	0.0000	0.0366	0.0000
57	NBP[0]752*	0.0000	0.0000	0.0000	0.0410	0.0000
58	NBP[0]784*	0.0000	0,0000	0,0000	0.0465	0.0000
59	NBP[0]830*	0.0000	0.0000	0.0000	0.0959	0.0000
60	NBP[0]888*	0,0000	0.0000	0.0000	0.0929	0.0000
61	NBP[0]947*	0.0000	0.0000	0.0000	0.0924	0.0000
62	NBP[0]1009*	0.0000	0.0000	0.0000 .1.3 (Build 4827)	0.1181	0.0000
63	Hyprotech Ltd.	A STATE OF THE PARTY OF THE PAR	HI313 V3	(Dullu 402/)		Page 19 o



Table 5. 23: product composition of ADU

1		TEAMIND	Case Na	me: C:\Program Files	\Hyprotech\HYSYS 3.1.3\Sam	ples\dyncrude3.hsc		
1	HYPROTECH	TEAM LND Calgary, Alberta	Unit Set:	Field-USGPM	BURNES BURNES			
7	LIFECYCLE INNOVATION	CANADA	Date/Tin	ne: Sun May 03 14:5	8:57 2009	Marie Co.		
+	, head with the land and the		Charles the part of the sec	5111 (May 65 (4.5)		515 (44.5) 5 (- 1979) 56 (- 1979)		
1		Column Sub	-Flowsheet:	Γ-100 @Main ((continued)			
5		19	SUMN	SUMMARY				
1		Off Gas	Naphtha	Waste Water	Residue	Kerosene		
3	NBP[0]1062*	0.0000	0.0000	0.0000	0.1614	0.0000		
4	NBP[0]1124*	0.0000 Diesel	0.0000 AGO	0.0000	0.1586	0.0000		
5	Flow Rate (lbmole/hr)	1.135494e+03	202.4290		A TRACK BUT SOME OF BUTCH	Carlo And A		
6	r low reals (ibitioleyit)	- 1,1004940.00	_	************************				
7	Methane	0.0000	0.0000					
8	Ethane	0.0000	0.0000	And the second s	************************************			
9	Propane	0.0000	0.0000					
0	i-Butane	0.0000	0.0000					
1	n-Butane	0.0000	0.0000					
2	H2O	0.0062	0.0077					
3	NBP[0]49*	0.0000	0.0000					
4	NBP[0]79*	0.0000	0.0000					
5	NBP[0]111*	0.0000	0.0000					
6	NBP[0]144*	0.0001	0.0000					
7	NBP[0]176*	0.0001	0.0000			***************************************		
8	NBP[0]208*	0.0003	0.0000					
9	NBP[0]240*	0.0006	0.0000	**************************************				
0	NBP[0]272*	0.0013 0.0027	0.0000	The second secon				
12	NBP[0]304*	0.0027	0,0001					
3	NBP[0]368*	0.0093	0.0001					
4	NBP[0]400*	0.0170	0.0005					
5	NBP[0]433*	0.0354	0.0013	Al recording to the highest community to any other to the property of the second secon				
6	NBP[0]464*	0.0887	0.0035	3.0				
7	NBP[0]496*	0.1748	0.0080					
8	NBP[0]528*	0,1769	0.0155					
9	NBP[0]560*	0.1504	0.0272			*****************************		
0	NBP[0]592*	0.1239	0.0453			*****		
1	NBP[0]624*	0.0943	0.0716	•	A STATE OF THE STA	and an entering to your day have been properly and the		
2	NBP[0]656*	0.0648	0.1142			V		
3	NBP[0]688*	0.0353	0.1803	***************************************				
4	NBP[0]720*	0.0109	0.2212					
5	NBP[0]752*	0.0017	0.1818					
6	NBP[0]784*	0.0002	0.1003	78-1-1-1-1-1-1-1-1-1-1-1-1-1-1-1-1-1-1-1		***************************************		
7	NBP[0]830*	0.0000	0.0211					
8	NBP[0]888*	0.0000	0.0003					
9	NBP[0]947*	0.0000	0.0000	***************************************				
0	NBP[0]1009*	0.0000	0.0000					
1	NBP[0]1062*	0,0000	0.0000	***************************************	***************************************			
2	NBP[0]1124*	0.0000	0.0000 Molar					
3	Flow Basis:		Produc	Flows	e composition option is select	ed		
4		Off Gas	Naphtha	Waste Water	Book T			
5	Eleva Parte (Ib. 1 h h h	7.3082	2.758898e+03 *	698.8582 *	Residue	Kerosene		
3	Flow Rate (lbmole/hr)			090,0302	1.594396e+03 ·	718,0960		
7	Mathema (harris da)	0.3744	0.8192 *	0,0000	0.0004			
9	Methane (lbmole/hr)	0.2537	2.8162	0.0000	0.0001	0.0000		
0	Ethane (lbmole/hr) Propane (lbmole/hr)	1.3070	47.7343 -	0.0000	0.0003	0.0000		
1	i-Butane (lbmole/hr)	0.3778	32.1524	0.0000	8800.0	0.0000		
2	n-Butane (ibmole/hr)	0.9737	115.7150	0.0000	0.0093	0.0000		
- 60		The second second second second		3 (Build 4827)	0.0381	0.0001		



Table 5. 24: product flow of ADU

	TEAM LND	Case Name	C:\Program Files\	Hyprotech\HYSYS 3,1.3\Sampl	es/dyncrude3.hsc			
HYPROTECH	Calgary, Alberta	Unit Set:	Field-USGPM					
CITICIDE INNOVATION	CANADA	Date/Time:	Sun May 03 14:58	:57 2009				
	O-luwer Cub	Elowebest: T	100 @##=:- /					
	Column Sub-	Flowsheet: T-	· ioo @iviain (continuea)				
		SUMMA	SUMMARY					
	Off Gas	Naphtha	Waste Water	Residue	Kerosene			
H2O (lbmole/hr)	0.5211	3.1823	698.8582 +	10.6450	0.0000			
NBP[0]49* (lbmole/hr)	1.2971	220.8109 •	0.0000 •	0.1117 •	0.0010			
4 NBP[0]79* (lbmole/hr)	0.9867	267.1707	0.0000 -	0.1657	0.0030			
NBP[0]111* (lbmole/hr)	0.5539 *	263.7163	0.0000 +	0.2086	0.0092			
NBP[0]144* (lbmole/hr)	0.3024	257.9673	0.0000 -	0.2687	0.0298			
NBP[0]176* (lbmole/hr)	0.1757	272.2966	0.0000 •	0.3615	0.0972			
NBP[0]208* (lbmole/hr)	0.0972	280.3578	0.0000 •	0.5205	0.3685			
NBP[0]240* (lbmole/hr)	0.0491	271.7889	0.0000	0.7168	1.2797			
NBP[0]272* (lbmole/hr)	0.0235	256,2762	0.0000	0.9808	3.9615			
NBP[0]304" (lbmole/hr)	0.0104	228.7481	0,0000	1.3246 •	10.7080			
2 NBP[0]336* (lbmole/hr)	0.0039	180.7228	0.0000	1.7891 •	34.0955			
3 NBP[0]368* (ibmole/hr)	0.0006	54.4279	0.0000	2.4806	141.4605			
NBP[0]400* (lbmole/hr)	0.0000	2.1494	0.0000	3.5796 -	181.8632			
NBP[0]433* (lbmole/hr)	0.0000	0.0448	0.0000	5.6996	176.1014			
6 NBP[0]464* (lbmole/hr)	0.0000	0.0008	0.0000	9.2951	135.9441			
NBP[0]496* (lbmole/hr)	0.0000	0.0000	0.0000	13.5893	30.4621			
NBP[0]528" (lbmole/hr)	0.0000	0.0000	0.0000	17.9759	1.6550			
NBP[0]560* (lbmole/hr)	0.0000	0.0000	0.0000	22.9605	0.0546			
NBP[0]592* (lbmole/hr)	0.0000	0,0000	0.0000	29.0421	0.0015			
1 NBP[0]624* (lbmole/hr)	0,0000	0.0000	0.0000 •	35.2450	0.0000			
2 NBP[0]656* (lbmole/hr)	0,0000	0.0000	0.0000	42.2105	0.0000			
3 NBP[0]688* (lbmole/hr)	0.0000	0.0000	0.0000 -	50.3987	0.0000			
4 NBP[0]720* (lbmole/hr)	0.0000	0.0000	0.0000	58,2931	0.0000			
NBP[0]752* (ibmole/hr)	0.0000	0.0000	0.0000	65.3378 * 74.2189 *	0.0000			
6 NBP[0]784* (ibmole/hr)	0.0000	0.0000	0.0000	152.8483	0.0000			
NBP[0]830* (lbmole/hr)	0.0000	0.0000	0.0000 -	148.1189	0.0000			
8 NBP[0]888* (lbmole/hr) 9 NBP[0]947* (lbmole/hr)	0.0000	0.0000	0.0000	147.3449	0.0000			
	0.0000	0.0000	0.0000	188.3514	0.0000			
NBP[0]1009* (lbmole/hr)	0.0000	0.0000	0.0000	257.3466	0.0000			
1 NBP[0]1062* (lbmole/hr)	0.0000	0.0000	0.0000	252.9099 -	0.0000 •			
2 NBP[0]1124* (ibmole/hr)	Diesel	AGO	CANONINA SUCCES	A STATE OF THE PARTY OF THE PAR	0.0000			
3 D. Marsin Dri	1.135494e+03 *	202,4290 *			A CHESTODANS CACH.			
Flow Rate (lbmole/hr)			***************************************	***************************************				
5 Mathewa (hmola/hr)	0.0000 -	0.0000	***************************************					
Methane (lbmole/hr)	0.0000	0.0000						
Ethane (lbmole/hr) Propane (lbmole/hr)	0.0001	0.0000	***************************************					
Propane (Ibmole/hr) i-Butane (Ibmole/hr)	0.0002	0.0000 •	***************************************					
n-Butane (lbmole/hr)	0.0011	0.0000						
1 H2O (lbmole/hr)	7,0192	1.5635 •						
NBP[0]49* (lbmole/hr)	0.0058	0.0000 *						
NBP[0]79* (lbmole/hr)	0.0129	0,0000 +			***************************************			
NBP[0]111* (lbmole/hr)	0.0267	0.0001	***************************************		**************			
NBP[0]144* (lbmole/hr)	0.0575	0.0001		- CARREST STORMS AND A STORM AND A STORMS AND A STORMS AND A STORMS AND A STORMS AND A STORM AND A STORMS AND A STORMS AND A STORMS AND A STORMS AND A STORM AND A STORMS AND A STORMS AND A STORMS AND A STORMS AND A STORM AND A STORMS AND A STORMS AND A STORMS AND A STORMS AND A STORM AND A S				
NBP[0]176" (lbmole/hr)	0.1253	0.0002			The same of the sa			
NBP[0]208* (ibmole/hr)	0.3071	0.0005			***************************************			
NBP[0]240* (ibmole/hr)	0.7026	0.0012 *						
9 NBP[0]272* (lbmole/hr)	1.5272	0,0028						
0 NBP[0]304* (lbmole/hr)	3.0827	0.0065						
1 NBP[0]336* (lbmole/hr)	5,8238	0.0152 *						
	With the Administration of the Conference of the	0.0370			***************************************			
2 NBP[0]368* (lbmole/hr)	10.6005	HYSYS v3.1.3						



Table 5. 25: product recoveries of ADU

2		TEAM LND	Case N	ame: C:\Program File	s\Hyprotech\HYSYS 3,1.3\Sa	mples\dyncrude3.hsc
1	HYPROTECH	Calgary, Alberta	Unit Se	t: Field-USGPM		
1	CITICOLE INTOVATION	CANADA	Date/Ti	me: Sun May 03 14:	58:57 2009	
j			March 10 Miles	Co. Co. Co. Co. Co. Co. Co. Co. Co. Co.		(1) (1) (1) (1) (1) (1) (1) (1) (1) (1)
4		Column Su	b-Flowsheet:	T-100 @Main	(continued)	
+			A STATE OF THE STA	25.5 604 (6.4)	A RECEIPTION OF THE	
0			SUM	MARY		
1	MARINES OF THE PARTY OF THE	Diesel	AGO		CANAGORAL CONTACT	
3	NBP[0]400* (lbmole/hr)	19.3232	0.0936			
4	NBP[0]433* (lbmole/hr) NBP[0]464* (lbmole/hr)	40.2346 100.6787	0.2606 •			
5	NBP[0]496* (ibmole/hr)	198.4789	1.6126			
6	NBP[0]528* (lbmole/hr)	200.9207	3.1284 -			
7	NBP[0]560* (lbmole/hr)	170.7404	5.5132 •			
8	NBP[0]592* (lbmole/hr)	140.6510	9.1767 •			
9	NBP[0]624* (lbmole/hr)	107.0700	14.4953			
1	NBP[0]656* (lbmole/hr) NBP[0]688* (lbmole/hr)	73.5300	23.1158			
2	NBP[0]720* (lbmole/hr)	40.0959 12.3762	36.4920 - 44.7791 -			
3	NBP[0]752* (lbmole/hr)	1.9266	36.7959			
4	NBP[0]784* (lbmole/hr)	0.1725	20.3028			
5	NBP[0]830* (lbmole/hr)	0.0023	4.2747 -			****
6	NBP[0]888* (lbmole/hr)	0.0000	0.0564			
7	NBP[0]947* (Ibmole/hr)	0.0000	0.0005			
В	NBP[0]1009* (lbmole/hr)	0.0000	0.0000			
9	NBP[0]1062* (lbmole/hr)	0.0000	0.0000	**************************************		
1	NBP[0]1124* (lbmole/hr)	0.0000	0.0000 • Molar		L .	
• 1	I ION DUSIS.					
2	美拉斯尼斯斯斯特美国斯 斯尼亚	A LONG THE REAL PROPERTY OF THE PERSON NAMED IN COLUMN TO PERSON NAMED		lecoveries	e composition option is selec	ted
-		Off Gas		ecoveries Waste Water	AND STORY OF STREET	ELECTIVE PROPERTY
3	Flow Rate (fbmole/hr)	Off Gas 7.3082	Product R	ecoveries	Residue 1.594396e+03	Kerosene
4		7.3082	Product R Naphtha 2.758898e+03	Waste Water 698.8582	Residue	ELECTIVE PROPERTY
3	Methane (%)	7.3082 21.1318	Product R Naphtha 2.758898e+03 46.2430	Waste Water 698.8582 — 0.0000	Residue 1.594396e+03 — 0.0036	Kerosene
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	Methane (%) Ethane (%)	7.3082 21.1318 6.5263	Product R Naphtha 2.758898+03 46.2430 72.4541	Waste Water 698.8582 0.0000 0.0000	Residue 1.594396e+03 — 0.0036 0.0085	Kerosene 718.0960
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	Methane (%) Ethane (%) Propane (%)	7.3082 21.1318 6.5263 2.4340	Product R Naphtha 2.758898+03 46.2430 72.4541 88.8946	Waste Water 698.8582 0.0000 0.0000	Residue 1.594396e+03 0.0036 0.0085 0.0164	Kerosene 718.0960 — 0.0000
3 3 5 5 6 6 6 6 6 6 6 6 6 6 6 6 6 6 6 6	Methane (%) Ethane (%) Propane (%) i-Butane (%)	7.3082 21.1318 6.5263	Product R Naphtha 2.758898+03 46.2430 72.4541	Waste Water 698.882	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273	Kerosens 718.0960 0.0000 0.0000 0.0000 0.0000
3 4 5 6 7 8 9	Methane (%) Ethane (%) Propane (%)	7.3082 	Product R Naphtha 2.758898+03 46.2430 72.4541 88.8946 94.8784	Waste Water 698.8582 0.0000 0.0000 0.0000 0.0000 0.0000	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317	Kerosens 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001
3 4 5 3 7 8 9	Methane (%) Ethane (%) Propane (%) i-Butane (%) n-Butane (%)	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103	Product R Naphtha 2.758898+03 46.2430 72.4541 88.8946 94.8784 96.2980	Waste Water 698.882	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273	Kerosens 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001
2 3 4 5 6 7 3 9 9	Methane (%) Ethane (%) Propane (%) i-Butane (%) n-Butane (%) H2O (%)	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722	Product R Naphtha 2.758898+03 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410	Waste Water 698.8582 0.0000 0.0000 0.0000 0.0000 0.0000 96.8462	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0000 0.0000
3 4 5 6 7 3 9	Methane (%) Ethane (%) Propane (%) i-Butane (%) n-Butane (%) H2O (%) NBP[0]49* (%)	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081	Product R Naphtha 2.758898e+03	Waste Water 698.8582 0.0000 0.0000 0.0000 0.0000 0.0000 96.8462 0.0000	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0000 0.0004 0.0011
3 4 4 5 6 6 6 6 6 6 6 6 6 6 6 6 6 6 6 6 6	Methane (%) Ethane (%) Propane (%) I-Butane (%) n-Butane (%) H2O (%) NBP[0]49* (%) NBP[0]111* (%) NBP[0]144* (%)	7.3082 	Product R Naphtha 2.758898+03	Waste Water 698.8582 0.0000 0.0000 0.0000 0.0000 0.0000 96.8462 0.0000 0.0000 0.0000	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0000 0.0004 0.0011 0.0035
3 4 5 5 3 3 5 5 5 5 6 5 6 6 6 6 6 6 6 6 6 6	Methane (%) Ethane (%) Propane (%) I-Butane (%) n-Butane (%) H2O (%) NBP[0]49" (%) NBP[0]111" (%) NBP[0]1144" (%) NBP[0]176" (%)	7.3082 —- 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644	Product R Naphtha 2.758898+03	Waste Water 698.8522	Residue 1.594396e+03	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0000 0.0004 0.0011
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	Methane (%) Ethane (%) Propane (%) i-Butane (%) n-Butane (%) h-Butane (%) h-Poly (%) NBP[0]49* (%) NBP[0]79* (%) NBP[0]111* (%) NBP[0]144* (%) NBP[0]176* (%) NBP[0]208* (%)	7.3082 	Product R Naphtha 2.758898+03	Waste Water 698.8582	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1036 0.1324 0.1854	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115
3 4 5 5 3 3 5 5 5 5 6 5 6 6 6 6 6 6 6 6 6 6	Methane (%) Ethane (%) Propane (%) I-Butane (%) n-Butane (%) H2O (%) NBP[0]49* (%) NBP[0]111* (%) NBP[0]114* (%) NBP[0]144* (%) NBP[0]176* (%) NBP[0]208* (%) NBP[0]240* (%)	7.3082 —- 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346	Product R Naphtha 2.758898+03 46.2430 72.4541 38.8946 94.8784 96.2980 0.4410 97.4056 98.3747 99.0808 99.4861 99.7361 99.8769	Waste Water 698.8582 0.0000 0.0000 0.0000 0.0000 99.8462 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0000 0.0004 0.0011 0.0035 0.0115 0.0356
3 3 5 5 5 6 6 6 6 6 6 6 6 6 6 6 6 6 6 6	Methane (%) Ethane (%) Propane (%) i-Butane (%) n-Butane (%) hButane (%) H2O (%) NBP[0]49* (%) NBP[0]111* (%) NBP[0]114* (%) NBP[0]144* (%) NBP[0]176* (%) NBP[0]208* (%) NBP[0]208* (%) NBP[0]272* (%)	7.3082 	Product R Naphtha 2.758898+03	Waste Water 698.8582	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0000 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	Methane (%) Ethane (%) Propane (%) i-Butane (%) n-Butane (%) H2O (%) NBP[0]49* (%) NBP[0]111* (%) NBP[0]1144* (%) NBP[0]144* (%) NBP[0]208* (%) NBP[0]208* (%) NBP[0]2072* (%) NBP[0]2172* (%) NBP[0]2172* (%) NBP[0]2174* (%)	7.3082 — 21.1318 6.5263 2.4340 1.1148 0.8103 0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090	Product R Naphtha 2.758898e+03 — 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410 97.4056 98.3747 99.0808 99.4861 99.7361 99.8769 99.7106 98.3943	Waste Water 698.8582 0.0000 0.0000 0.0000 0.0000 99.8462 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961
3 3 4 4 5 5 5 5 5 5 5 5 5 5 5 5 5 5 5 5	Methane (%) Ethane (%) Propane (%) i-Butane (%) n-Butane (%) hButane (%) H2O (%) NBP[0]49* (%) NBP[0]111* (%) NBP[0]114* (%) NBP[0]144* (%) NBP[0]176* (%) NBP[0]208* (%) NBP[0]208* (%) NBP[0]272* (%)	7.3082 	Product R Naphtha 2.758898e+03 — 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410 97.4058 98.3747 99.0808 99.4861 99.7361 99.7569 99.7106 98.3943 93.9102	COVER'S Waste Water 698.8582	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	Methane (%) Ethane (%) Propane (%) Propane (%) i-Butane (%) n-Butane (%) H2O (%) NBP[0]49* (%) NBP[0]111* (%) NBP[0]111* (%) NBP[0]114* (%) NBP[0]176* (%) NBP[0]203* (%) NBP[0]204* (%) NBP[0]2072* (%) NBP[0]203* (%) NBP[0]304* (%) NBP[0]304* (%) NBP[0]336* (%)	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090 0.0043 0.0017 0.0003	Product R Naphtha 2.758898+03	Waste Water 698.8582	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438 0.7932	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094
	Methane (%) Ethane (%) Propane (%) I-Butane	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090 0.0043 0.0017 0.0003 0.0000	Product R Naphtha 2.758898+03	Recoverles Waste Water 698.8582 0.0000	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438 0.7932 1.1645	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094 87.4195
	Methane (%) Ethane (%) Propane (%) I-Butane	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090 0.0043 0.0017 0.0003 0.0000 0.0000	Product R Naphtha 2.758898+03 — 46.2430 72.4541 38.8946 94.8784 96.2980 0.4410 97.4056 98.3747 99.0808 99.4861 99.7361 99.8769 99.7106 98.3943 93.9102 80.1243 25.5515 1.0332 0.0202 0.0003	Recoverles Waste Water 698.8582	Residue 1.594396e+03 0.0036 0.0035 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438 0.7932 1.1645 1.7207 2.5662 3.7852	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094 87.4195 79.2881
	Methane (%) Ethane (%) Propane (%) I-Butane	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090 0.0043 0.0017 0.0003 0.0000 0.0000	Product R Naphtha 2.758898+03 — 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410 97.4056 98.3747 99.0808 99.4861 99.7361 99.8769 99.7106 98.3943 93.9102 80.1243 25.5515 1.0332 0.0202 0.0003 0.0000	Recoverles Waste Water 698.8582	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438 0.7932 1.1645 1.7207 2.5662 3.7852 5.5526	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094 87.4195
	Methane (%) Ethane (%) Propane (%) I-Butane	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0348 0.0180 0.0090 0.0043 0.0017 0.0003 0.0000 0.0000 0.0000	Product R Naphtha 2.758898+03 — 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410 97.4056 98.3747 99.0808 99.4861 99.7361 99.8769 99.7106 98.3943 93.9102 80.1243 25.5515 1.0332 0.0202 0.0003 0.0000	Recoverles Waste Water 698.8582 0.0000	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438 0.7932 1.1645 1.7207 2.5662 3.7852 5.5526 8.0995	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094 87.4195 79.2881 55.3596
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	Methane (%) Ethane (%) Propane (%) i-Butane (%) n-Butane (%) h-Butane	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090 0.0003 0.0003 0.0000 0.0000 0.0000 0.0000	Product R Naphtha 2.758898+03 — 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410 97.4056 98.3747 99.0808 99.4861 99.7661 99.7706 98.3943 93.9102 80.1243 25.5515 1.0332 0.0202 0.0003 0.0000 0.0000	### Coverles Waste Water	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438 0.7932 1.1645 1.7207 2.5662 3.7852 5.5526 8.0995	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094 87.4195 79.2881 55.3596 12.4467
3 4 5 5 5 5 5 5 5 5 5 5 5 5 5 5 5 5 5 5	Methane (%) Ethane (%) Propane (%) I-Butane (%) n-Butane (%) NBP[0]49° (%) NBP[0]111° (%) NBP[0]111° (%) NBP[0]111° (%) NBP[0]111° (%) NBP[0]240° (%) NBP[0]240° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]520° (%) NBP[0]520° (%)	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090 0.0003 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	Product R Naphtha 2.758898e+03 — 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410 97.4058 98.3747 99.0808 99.4861 99.7361 99.8769 99.7106 98.3943 93.9102 80.1243 25.5515 1.0332 0.0202 0.0003 0.0000 0.0000 0.0000	### Coverles Waste Water	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438 0.7932 1.1645 1.7207 2.5662 3.7852 5.5526 8.0995 11.7420 16.7205	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0004 0.0011 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094 87.4195 79.2881 55.3596 12.4467 0.7457
3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3 3	Methane (%) Ethane (%) Propane (%) Propane (%) I-Butane (%) n-Butane (%) NBP[0]49° (%) NBP[0]111° (%) NBP[0]111° (%) NBP[0]111° (%) NBP[0]240° (%) NBP[0]364° (%) NBP[0]368° (%) NBP[0]368° (%) NBP[0]368° (%) NBP[0]368° (%) NBP[0]368° (%) NBP[0]466° (%) NBP[0]466° (%) NBP[0]466° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%) NBP[0]528° (%)	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090 0.0003 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	Product R Naphtha 2.758898+03 — 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410 97.4056 98.3747 99.0808 99.4861 99.7661 99.7706 98.3943 93.9102 80.1243 25.5515 1.0332 0.0202 0.0003 0.0000 0.0000	Recoverles Waste Water 698.8582 0.0000	Residue 1.594396e+03	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0001 0.0001 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094 87.4195 79.2881 55.3596 12.4467 0.7457 0.0279 0.0009 0.0000
	Methane (%) Ethane (%) Propane (%) I-Butane (%) n-Butane (%) NBP[0]49° (%) NBP[0]111° (%) NBP[0]111° (%) NBP[0]111° (%) NBP[0]111° (%) NBP[0]240° (%) NBP[0]240° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]304° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]400° (%) NBP[0]520° (%) NBP[0]520° (%)	7.3082 21.1318 6.5263 2.4340 1.1148 0.8103 -0.0722 0.5722 0.3633 0.2081 0.1166 0.0644 0.0346 0.0180 0.0090 0.0003 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	Product R Naphtha 2.758898e+03 — 46.2430 72.4541 88.8946 94.8784 96.2980 0.4410 97.4058 98.3747 99.0808 99.4861 99.7361 99.8769 99.7106 98.3943 93.9102 80.1243 25.5515 1.0332 0.0202 0.0003 0.0000 0.0000 0.0000 0.0000	Waste Water 698.8582	Residue 1.594396e+03 0.0036 0.0085 0.0164 0.0273 0.0317 1.4752 0.0493 0.0610 0.0784 0.1036 0.1324 0.1854 0.2630 0.3766 0.5438 0.7932 1.1645 1.7207 2.5662 3.7852 5.5526 8.0995 11.7420 16.7205	Kerosene 718.0960 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0001 0.0001 0.0035 0.0115 0.0356 0.1313 0.4695 1.5210 4.3961 15.1164 66.4094 87.4195 79.281 55.3596 12.4467 0.7457 0.0279 0.0009



Table 5. 26: product recoveries of ADU

(2) (1) (1) (1) (1) (1) (1) (1) (1) (1) (1	0.0000 Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0000 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458 2.0245 0.2003 0.0016 0.0000 0.0000 0.0000	0.0000 AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0001 0.0002 0.0004 0.0017 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.6529 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747 38.6641 23.5770 2.9900 0.0428 0.0003 0.0000 0.0000 0.00000	0.0000 0.0000 ROFILES The Flows Option is s	11:	3.3171	0.0000
[0]1124* (%) Late (lbmole/hr) Late (%) Late (lbmole/hr) Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458 2.0245 0.2003 0.0016 0.0000 0.0000 0.0000	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.05589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747 38.6641 23.5770 2.9900 0.0428 0.0003 0.0000 0.0000	0.0000	11	3.3171		
[0]1124* (%) Late (lbmole/hr) Late (%) Late (lbmole/hr) Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458 2.0245 0.2003 0.0016 0.0000 0.0000 0.0000	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0025 0.0067 0.0174 0.0450 0.1173 0.2868 0.6589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747 38.6841 23.5770 2.9900 0.0428 0.0003 0.0000	**************************************	11	3.3171		
[0]1124* (%) Itale (lbmole/hr) Itale (lbmole/hr) Itale (lbmole/hr) Itale (%) Itale	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458 2.0245 0.2003 0.0016 0.0000 0.0000	AGO 202.4290 — 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0025 0.0067 0.0174 0.0450 0.1173 0.2868 0.5589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747 38.6841 23.5770 2.9900 0.0428 0.0003 0.0000	**************************************	11	3.3171	
[0]1124* (%) Italiana (%) It	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458 2.0245 0.2003 0.0016 0.0000	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.6589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747 38.6641 23.5770 2.9900 0.0428 0.0003	**************************************	11	3.3171	
(ID) 1124* (%) Italia (Ibmole/hr) Italia (Ib	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458 2.0245 0.2003 0.0016 0.0000	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.05589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747 38.6641 23.5770 2.9900 0.0428	**************************************	11	3.3171	
[0]1124* (%) Interpolation (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458 2.0245 0.2003	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.5589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747 38.6841 23.5770	**************************************	11	3.3171	
[0]1124* (%) Late (lbmole/hr) Late (%) Late (Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458 2.0245	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.0174 0.0450 0.1173 0.2868 0.6589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747 38.6641	**************************************	11	3.3171	
[0]1124* (%) Late (lbmole/hr) Late (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940 11.5458	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0028 0.0067 0.0174 0.0450 0.1173 0.2868 0.6589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745 41.7747	**************************************	11	3.3171	
[0]1124* (%) Late (lbmole/hr) Late (%) Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141 55.4794 33.5940	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.6589 1.4096 2.8195 5.2833 9.6004 17.4412 30.5745	**************************************	11	3.3171		
[U]1124* (%) Intale (lbmole/hr) Intale (lbmole/hr) Intale (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0987 81.0987 80.5303 87.3168 80.9774 70.9141 55.4794	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.2167 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.6589 1.4096 2.8195 5.2833 9.6004 17.4412	**************************************	11	3.3171	
[0]1124* (%) Itale (lbmole/hr) Itale (lbmole/hr) Itale (lbmole/hr) Itale (%) Itale	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774 70.9141	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.2167 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.6589 1.4096 2.8195 5.2833 9.6004	**************************************	11	3.3171	
(ID) 1124* (%) Italiana (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303 87.3168 80.9774	AGO 202.4290 — 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174 0.0450 0.1173 0.2868 0.6589 1.4096 2.8195 5.2833	**************************************	11	3.3171	
(ID) 1124* (%) Italiana (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978 90.5303	AGO 202.4290 — 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0028 0.0067 0.0174 0.0450 0.0174 0.0450 0.0173 0.2868 0.6589 1.4096	**************************************	11	3.3171	
[0]1124* (%) Late (lbmole/hr) Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987 81.0978	AGO 202.4290 — 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0028 0.0067 0.0174 0.0450 0.1173 0.2868 0.6589	**************************************	11	3.3171		
[0]1124* (%) Late (lbmole/hr) Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153 40.9987	AGO 202.4290 — 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.00174 0.0450 0.1173 0.2868	**************************************	11	3.3171		
[0]1124* (%) Late (lbmole/hr) Late (%) Late (Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884 18.1153	AGO 202.4290	**************************************	11	3.3171	
(U)1124* (%) (atale (lbmole/hr) (bthane (%) Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764 9.2884	AGO 202.4290	**************************************	11	3.3171		
(U)1124* (%) (atale (lbmole/hr) (bthane (%) Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820 4.9764	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067 0.0174	**************************************	11	3.3171		
(U)1124* (%) Intaine (%) Inta	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656 2.5820	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0002 0.0004 0.0011 0.0026 0.0067	**************************************	11	3.3171	
(ID) 1124* (%) Intaine (%) In	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863 1.2656	AGO 202.4290 — 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0002 0.0004 0.0011 0.0026	**************************************	11	3.3171	
(ID) 1124* (%) Italiana (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094 0.2578 0.5863	AGO 202.4290	**************************************	11	3.3171	
[0]1124* (%) Late (lbmole/hr) Late (lbmole/hr) Late (lbmole/hr) Late (lbmole/hr) Late (lbmole/hr) Late (lbmole/hr) Late (%) Late (lbmole/hr) Late (lb	Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459 0.1094	AGO 202.4290 0.0000 0.0000 0.0000 0.2167 0.0000	**************************************	11	3.3171	
(U)1124* (%) (ate (lbmole/hr) (athane (%) (bthane (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222 0.0459	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.2167 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	**************************************	11	3.3171	
(U)1124* (%) sate (lbmole/hr) sthane (%) thane (%) opane (%) Sutane (%) Sutane (%) P(U)19* (%) P(U)11** (%) P(U)11** (%) P(U)14** (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100 0.0222	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.2167 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	**************************************	11	3.3171	
(i) (i) (ii) (iii)	Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047 0.0100	AGO 202.4290 0.0000 0.0000 0.0000 0.2167 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	**************************************	11	3.3171	
(%) (%) (%) (%) (%) (%) (%) (%) (%) (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025 0.0047	AGO 202.4290 0.0000 0.0000 0.0000 0.2167 0.0000 0.0000 0.0000	**************************************	11	3.3171	
(U)1124* (%) tate (lbmole/hr) whate (%) thane (%) page (%) tutane (%) sutane (%) 120 (%) P[0]49* (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727 0.0025	AGO 202.4290 0.0000	**************************************	11	3.3171	
(0)1124* (%) tate (lbmole/hr) othane (%) thane (%) opane (%) butane (%) outne (%) 120 (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009 0.9727	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.2167	**************************************	11	3.3171	
jUj1124* (%) sale (lbmole/hr) sthane (%) shane (%) opane (%) sutane (%) sutane (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000 0.0001 0.0006 0.0009	AGO 202.4290 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	**************************************	11	3.3171	
io[1124* (%) tate (lbmole/hr) othane (%) thane (%)	Diesel 1.135494e+03 0.0000 0.0000 0.0000	AGO 202.4290 0.0000 0.0000 0.0000 0.0000	**************************************	11	3.3171	
[0]1124* (%) tate (lbmole/hr) othane (%)	Diesel 1.135494e+03 0.0000 0.0000	AGO 202.4290 0.0000 0.0000	**************************************	11	3.3171	
IO[1124* (%) Late (lbmole/hr)	Diesel 1.135494e+03 0.0000	AGO 202.4290 0.0000	**************************************	11	3.3171	
[0]1124* (%) tate (lbmole/hr)	Diesel 1.135494e+03	AGO 202.4290 	**************************************	11	3.3171	
[0]1124° (%)	Diesel	AGO	**************************************	11	3.3171	
[0]1124° (%)	Diesel	AGO	**************************************	11	3.3171	
Settle County and County Street, and County Street, Co.	and the second of the second o	A CONTRACTOR OF THE PARTY OF TH	**************************************	11	3.3171	
Settle County and County Street, and County Street, Co.	0.0000		**************************************		*************************	0.0000
[0]1062* (%)	0.0000	0.0000		- 11.	******************************	
[0]1009* (%)	0,0000	0.0000	0.0000	44	3.2397	0.0000
P[0]947* (%)	0.0000	0.0000	. 0.0000		3.0482	0.0000
P[0]888* (%)	0.0000	0.0000	0.0000		2.3561	0.0000
P[0]830° (%)	0,0000	0.0000	0.0000	***************************************	6.9107	0.0000
(%)	0.0000	0.0000	0.0000			0.0000
TO ACCUSE ANY ADDRESS OF A PARTY OF THE PARTY.	Marie Control of the	0.0000	0.0000			0.0000
MO1720* (SC)						Kerosene
reason than a section						
og der sette Atte orde	AND PROPERTY OF THE PARTY OF	SUMM	ARY		TO STATE OF THE ST	
	Column Sub-Flo	wsheet: T	-100 @Main (contin	ued)	
111111111111111111111111111111111111111						
	CANADA	Date/Time	Sun May 03 14:58	3:57 2009		
ROTECH	Calgary, Alberta	Unit Set:	Field-USGPM			
•	TEAM LND	Case Hall	C. Program Files	rryprotection	11515 3.1.3\Samp	les/dyncrude3.hsc
	(0)720° (%) (0)752° (%) (0)752° (%)	CANADA Column Sub-Flo Off Gas (0)7720* (%) 0,0000 (0)752* (%) 0,0000 (0)784* (%) 0,0000	TEAM LND Calgary, Alberta CANADA Column Sub-Flowsheet: T SUMMA Off Gas Naphtha (0)7720* (%) 0.0000 0.0000 (0)752* (%) 0.0000 0.0000 (0)752* (%) 0.0000 0.0000 (0)754* (%) 0.0000 0.0000	TEAM LND Calgary, Alberta CANADA Column Sub-Flowsheet: T-100 @Main (SUMMARY Off Gas Naphtha Waste Water (0)7720* (%) 0.0000 0.0000 0.0000 (0)752* (%) 0.0000 0.0000 0.0000 (0)752* (%) 0.0000 0.0000 0.0000 (0)752* (%) 0.0000 0.0000 0.0000	TEAM LND Calgary. Alberta CANADA Unit Set: Field-USGPM Date/Time: Sun May 03 14:58:57 2009 Column Sub-Flowsheet: T-100 @Main (conting summary) SUMMARY Off Gas Naphtha Waste Water Re (0)720* (%) 0.0000 0.0000 0.0000 54 0)752* (%) 0.0000 0.0000 0.0000 68 (0)754* (%) 0.0000 0.0000 0.0000 86	TEAM LND Calgary, Alberta CANADA Date/Time: Sun May 03 14:58:57 2009 Column Sub-Flowsheet: T-100 @Main (continued) SUMMARY Off Gas Naphtha Waste Water Residue (0)720* (%) 0,0000 0,0000 0,0000 54,3820 (0)752* (%) 0,0000 0,0000 0,0000 68,6552 (0)784* (%) 0,0000 0,0000 0,0000 86,1879



Table 5. 27: column profiles of ADU

T				Case Name: C	:\Program Files\Hyprote	ch\HYSYS 3.1.3\Sample	es/dyncrude3.hsc
P	TYPROTEC	TEAM LND Calgary, Alber	la	Unit Set: F	ield-USGPM		
₹,	CIFECULE PRIOVA	CANADA		Date/Time: S	un May 03 14:58:57 200	9	
	SAFETY STAR	a wangalyaya ta	(6) (2) (8) (9)				
		Column	Sub-Flows	sheet: T-100	@Main (con	tinued)	
100	Menny to the Mary	o producence statistics		COLUMN PROFILES			
1	WORLD WITH THE TOTAL	The Manager of State Annual Con-	MARKETEN AND AND A CASE.	Column Profiles Flow		1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1	CONTRACTOR SECURIOR
船		Temperature	Pressure	Net Liq (lbmole/hr)	Net Vap (lbmole/hr)	Net Feed (lbmole/hr)	Net Draws (Ibmole/I
	*Condenser	remperature	1100000	1985	- (controlering)	1481 eed (ibindie1a)	3465
	1 Main TS	276.6	28.70	9273	5421		
1	2_Main TS	317.1	28.78	4919	7915		4794
	3_Main TS	343.5	28.86	4910	8355		
	4_Main TS	357.0	28.94	4829	8347		
1	5 Main TS	366.3	29.02	4717	8266		——————————————————————————————————————
	6 Main TS	374.5	29.11	4568	8155	_	
	7_Main TS	383,1	29.19	4363	8008	****	
Γ	8_Main TS	393.9	29.27	4096	7805	_	
	9_Main TS	407.0	29.35	2867	7332		928.0
+	10_Main TS	423.8	29.43	2668	7036		-
1	11 Main TS	437.3	29.51	2534	6839		
-	12 Main TS	447.2	29.59	2429	6712		
1	13 Main TS	455.0	29.67	2321	6612	_	
1	14 Main TS	462.1	29.75	2180	6509		
1	15 Main TS	470.5	29.84	1945	6373	***	****************
t	16 Main TS	483.7	29.92	4539	6145	***	
1	17 Main TS	511.5	30.00	971.2	6538		2470
╁	18 Main TS	552.6	30.08	710.1	6150		3172
t	19 Main TS	571.7	30.16	590.5	5894	***	
+	20 Main TS	580.6	30.24	508.2	5777	_	
1	21 Main TS	586.0	30.32	2467	5697		
+	22 Main TS	613.0	30.40	717.5	6017		1703
1	23 Main TS	634.3	30.48	601.9	5970		
+	24 Main TS	643.9	30.57	509.0	5853		
1	25 Main TS	650.2	30.65	426.4	5759		T
4	26 Main TS	655.5	30.73	321.0	5674		
-		662.4	30.81	123.7	5566	-	
4	27_Main TS	676.1	30.89	1556	5366	6231	
1	28 Main TS	667.1	30.97		566.7	416.3	
-	29 Main TS	430.2	29.35	1027			1594
1	1_Kero_SS		29.47	1067	309.5		210.1
_	2_Kero_SS	440.7 449.0	30.00	1089	349.5		<u> </u>
4	3_Kero_SS	449,0	30.00		371.3		
-	Kero SS Reb	E044	30,00	1274			718.1
	1_Diesel_SS	504.1	30.08	1223	304.8		383.1
L	2_Diesel_SS	498.1 485.9	30.15		254.6	166.5	
4	3 Diesel SS	water the state of	30.40	243.6			1135
	1_AGO_SS	600.3	30.48	225.8	180.0	-	211.0
1	2_AGO_SS	589.6 568.8	30.56		162.2	138.8	
100	3 AGO SS	SUB-U		Column Profiles Ene		MARKET CONTRACTOR	202.4
		Te	mperature	Liquid Enthalpy	Vapour E	nthalpy	Heat Loss (Btu/hr)
۲	*Condone		mar dather were	And the safety of the safety o			
-	*Condenser		276.6	-9.829e+004	-6.850	+004	
-	1_Main TS	and the second s	317.1	-1.004e+005		-7.331e+004	
-	2_Main TS		343.5	-1.032e+005	-7.705	With the property of the control of the second of the seco	
-	3_Main TS	contract the track of the point of the second of the secon	357.0	-1.051e+005	-7,866		The second secon
-	4_Main TS		366.3	-1.068e+005	-7.955	The box of the late of the lat	
	5_Main TS		374.5	-1.085e+005	-8.015	Contract of the Contract of th	
1	6_Main TS			-1.107e+005		of the first of the first of the second of t	
2	7 Main TS 383.1		I	-1.107e+005 -8.064 HYSYS v3.1.3 (Build 4827)			



Table 5. 28: column profiles of ADU

$\frac{1}{2}$				Case	e Name: C	:\Program Files\Hypr	rotech\HYSYS 3,1.3	Samples\dyncrude3.h:	5Ç
]	HYPROT	TEAM LN Calgary.		Unit	Set: F	ield-USGPM			
$\frac{1}{1}$	Circovett in	CANADA		Date	Time: S	un May 03 14:58:57	2009		
1		Colu	nn Sub	-Flowsheet	: T-100	@Main (cc	ontinued)		
1				COLUM	IN PROFILES				
1		1100000	Temperature	Liq	uid Enthalpy	Vapour	Enthalpy	Heat Loss (Btu/l	hr)
1	8_Main 1	rs	393.9	-1	.136e+005		2e+004		
	9_Main 7	rs	407.0		.174e+005	-8.15	6e+004		
١	10_Main	TS	423.8		.216e+005	-8.22	6e+004		********
ļ	11Main	TS	437.3		.247e+005	-8.28	8e+004		
1	12Main	TS	447.2		.271e+005	-8.33	1e+004		
ļ	13_Main	TS	455.0		.291e+005		3e+004		
ļ	14_Main	TS	462.1		.314e+005		7e+004		
1	15_Main		470.5		.348e+005		8e+004		
ł	16_Main		483.7	made any more property and the contract of the	.417e+005		38+004		
l	17_Main		511.5		.460e+005		3e+004		
ł	18_Main	A D. (200 A D. (552.6	*******	1.659e+005		32e+004 13e+004		,
ł	19_Main	****	571.7	A STATE OF THE PARTY OF THE PAR	1.698e+005		55e+004		
1	20_Main	The state of the s	580.6		1.733e+005		73e+004		
1	21_Main		586.0	The second secon	1.736e+005		17e+004		
1	22_Main		613.0		1.788e+005		35e+004		
4	23_Main		634.3 643.9		1.835e+005		58e+004		
1	24_Main		650.2		1.879e+005		39e+004		
-	25_Main 26 Main		655.5		1.931e+005		18e+004		
1	27 Main		662.4		2.041e+005	A STATE OF THE PERSON NAMED IN COLUMN 2 IN	36e+004		-
1	28 Main		676.1		2.403e+005		28e+004		
3	29 Main		667.1	-	2.564e+005	-9.94	48e+004		
1	1 Kero_		430.2	-	1.153e+005	-8.50	51e+004		
5	2 Kero_		440.7	-	1.150e+005	-8.8	79e+004	***	
3	3_Kero_		449.0		1.152e+005	-9.0	78e+004		
7	Kero_SS_							***	
3	1 Diesel		504.1	-	1.497e+005	-9.93	23e+004		
9	2 Diesel	SS	498.1		1.519e+005	-1.00	27e+005		
5	3_Diesel	ss	485.9		1.553e+005	-1.0-	40e+005		.,
1	1_AGO_	ss	600.3		1.831e+005	-1.0	57e+005		
2	2_AGO_	ss	589.6		1.881e+005		69e+005		
3	3_AGO	ss	568.8	distribution of the second	1.951e+005	-1.0	56e+005		
5			. ipati		S / PRODUCT	'S			
5 7	Flow Basis:	Stream	Туре	Duty (Blu/hr	State	Flows (lbn	nole/hr Enthalpy	(Blu/lbmoi Temp	(F)
3		Atmos Cond	Energy	1.084e+008	-				
,	· · · · · · · · · · · · · · · · · · ·	Off Gas	Draw		Vapour	7.308	3345	114.0	
	*Condenser	Naphtha	Draw		Liquid	2759	5223	114.0	
1		Waste Water	Draw	-	Water	698.9	* -7617	114.0	-
2	4 14.1 70	<pa_1></pa_1>	Energy	-5.501e+007 *					
3	1Main TS	PA_1_Return	Feed		Liquid		-4.809e+	004 • 139.8	***************************************
ı	2_Main TS	PA 1 Draw	Draw		Liquid	4794	-6267	- 317.1	
	3 Main TS	N. Base				-			
3	4_Main TS					_			
	5 Main TS								
;	6_Main TS								-
į	7_Main TS	TALK.							
)	8_Main TS	<pa_4></pa_4>	Energy		Vancus				
1		Kero_SS_Return	Feed	4 =	Vapour Liquid	-	-3.676e+	The state of the s	
2	9 Main TS	Kero SS Draw	Draw	HVSVS	v3.1.3 (Build 4	928.0	• -7330	407.0	of Children by the
3	Hyprotech Ltd.								25 of



Table 5. 29: feed/products of ADU

1		7744-14	ID.		Case Name:	C:\Program Files\Hyp	rolech\HYSYS 3.1.3\Sample	s\dyncrude3.hsc	
	HYPROT	TEAM LN Calgary.	Alberta		Unit Set:	Field-USGPM			
1	Circovette in	CANADA			Date/Time:	Sun May 03 14:58:57	2009		
+	diction in the	7 - 636 (4.295)						A STATE OF THE STA	
		Colu	nn Sul	b-Flowshe	eet: T-10	0 @Main (co	ontinued)		
1				FE	EDS / PRODUC	CTS			
1	10 Main TS					- to Toron		T	
I	11_Main TS								
1	12_Main TS								
4	13_Main TS	- 1							
1	14 Main TS								
7	15_Main TS	51.5	Farmi	-3.500e+007	-				
-		<pa_6></pa_6>	Energy Feed	-3.3000+007	Vapour				
9	16_Main TS	Diesel_SS_Return <pa_2></pa_2>	Energy	-3.500e+007	-	_	-4.266e+004	504.1	* ****
5		PA 2 Return	Feed	_	Liquid		-7.104e+004	364.6	
1		Diesel_SS_Draw	Draw		Liquid	1352	9117	511.5	
2	17_Main TS	PA 2 Draw	Draw		Liquid	1820	9117	511.5	-
3	18_Main TS								
4	19 Main TS								
5	20_Main TS								
3		<pa_5></pa_5>	Energy	-3.500e+007	Veneue				
4	21_Main TS	AGO_SS_Return	Feed	2 500-1007	Vapour		-4.546e+004	600.3	
1	21_//	<pa_3></pa_3>	Energy	-3.500e+007	Liquid		2.544004		******
1		PA_3_Return	Feed		Liquid	274.6	-8.514e+004 * -1.083e+004	481.4	
4	22_Main TS	PA 3_Draw	Draw		Liquid	1429	-1.083e+004	613.0	
4	22 Main TC	PA 3 Draw	Diany				1,5000,004	613.0	
3	23 Main TS 24 Main TS							A	-
4	25 Main TS								
5	26 Main TS								
8	27 Main TS								
7		Q-Trim	Energy	6.642e+007					
В	28Main TS	Atm Feed	Feed		Mixed	6231	-5.151e+004	612.6	
9	29 Main TS	Main Steam	Feed		Vapour	416.3	-4.352e+004	375.0	-
0	29Main 13	Residue	Draw		Liquid	1594	-1.600e+004	667.1	
1	1_Kero_SS	Kero_SS_Draw	Feed		Liquid Vapour	210.4	-5.048e+004	407.0	
2	1_K610_00	1	Draw		Vapour	210.1	-5338	430.2	
3	2_Kero_SS		the state of the s	the contract of		Departure.		MARIE DE LA COMPANION DE LA CO	
4	3_Kero_SS	Vara CC Energy	Energy	7.500e+006	•			1.00	-
5	Kero_SS_Reb	Kero_SS_Energy Kerosene	Draw		Liquid	718.1	-7261	459.6	
6		Diesel SS_Draw	Feed	1 12-	Liquid		-6.278e+004	• 511.5	-
8	1Diesel_SS	2	Draw		Vapour	383.1	-6195	• 504.1	
٩	2 Diesel_SS			olo					
0	rest Block man person	Diesel Steam	Feed		Vapour	166.5	-4.379e+004	300.0	
1	3_Diesel_SS	Diesel	Draw	<u></u>	Liquid	1135	* -9695	485.9	
2	4 400 55	AGO_SS_Draw	Feed		Liquid		-7.462e+004	• 613.0	
3	1_AGO_SS	3	Draw		Vapour	211.0	-6601	* 600.3	
4	2_AGO_SS		F-12	-	Vapour	138.8	1272		
5	3_AGO_SS	AGO Steam	Feed		Liquid	202.4	-4.379e+004	300.0	
В		AGO	Draw	1.55	Section 1995	202.4	-1.218e+004	568.8	
7					SETUP				
3				Co	lumn Flowsheet To	ppology	CANADA SALE AND AND AND		10.00
9	Tabal The Co	es: 40 *	Total Trav	/-Sections:		denser + Reboiler:	2 * Pump Around		
0	Total Theor, Stag	es: 40°	Side Rect	The second secon		our Bypasses:	2 * Pump Around	s.	
1	Side Strippers:	3	1 55		20.1		<u> </u>		
41				HYS	SYS v3.1.3 (Buil	d 4827)	A Company of the Comp	Page 26	



Table 5. 30: feed/products & dynamics of ADU

		TEAM LND	Case Nam	e: C:\Program File	s\Hyprotech\HYSYS 3,1.3\San	nples/dyncrude3.hsc		
	MYPROTECH	Calgary, Alberta	Unit Set:	Field-USGPM				
+	Control in the control	CANADA	Date/Time	Date/Time: Sun May 03 14:58:57 2009				
-		Column Sub	-Flowsheet: T	-100 @Main	(continued)			
9			Sub-Flow	sheet				
0		Feed Streams			Product Streams			
1	Internal Stream	External Stream	Transfer Basis	Internal Stream	External Stream	Transfer Basis		
2	Main Steam	Main Steam @Main	T-P Flash	Residue	Residue @Main	T-P Flash		
3	Q-Trim	Q-Trim @Main	None Reg'd T-P Flash	Atmos Cond Off Gas	Atmos Cond @Main	None Req'd		
4	Atm Feed	Atm Feed @Main	None Reg'd	Waste Water	Off Gas @Main	T-P Flash		
15	Kero_SS_Energy	Disast Cleam @Main	T-P Flash	Naphtha	Waste Water @Main	T-P Flash		
17	Diesel Steam	Diesel Steam @Main AGO Steam @Main	T-P Flash	Kerosene	Naphtha @Main	T-P Flash		
18	AGO Steam	AGO Steam (Eyvani	<none set=""></none>	Diesel	Kerosene @Main	T-P Flash		
9	atm residue vac duty		None Reg'd	AGO	Diesel @Main	T-P Flash		
9	vac duly		a an area which processes to all the area of plants and appeal and of the area of the area.	PA_1_Q	AGO @Main	T-P Flash None Reg'd		
21				PA_2_Q	-	None Req'd		
22		•		PA_3_Q	The state of the s	None Reg'd		
23				vac feed	***************************************	<none requ<="" td=""></none>		
4			20, 100	The state of the s		11010 2017		
25			VARIA	RIFS				
26			A REPORT					
27		Co	olumn Flowsheet Vars	Available as Parame	ters	Relative and		
28	Data Source		Variable	Componer	nt	Description		
29								
30			COMPONE	NT MAPS				
31			19.70					
32			Feed St		·			
33	Feed	l Name	In to SubFI	owSheet	Out of Su	bFlowSheet		
34	Main Steam							
35	Q-Trim							
36	Atm Feed		**************************************	-				
~~	Kero_SS_Energy							
37	Diesel Steam							
37 38	Diesel Steam AGO Steam							
37 38 39								
37 38 39 40	AGO Steam		Product	Stream	•			
37 38 39 40 41	AGO Steam atm residue vac duty		Product In to SubF					
37 38 39 40 41 42	AGO Steam atm residue vac duty	ict Name	Product In to SubF		Out of Su	bFlowSheet		
37 38 39 40 41 42 43	AGO Steam atm residue vac duty	ct Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43	AGO Steam atm residue vac duty	ict Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas	ict Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45	AGO Steam atm residue vac duty Produ Residue Atmos Cond	ict Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha	ict Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha Kerosene	nct Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha Kerosene Diesel	ct Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha Kerosene Diesel AGO	ct Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha Kerosene Diesel AGO PA_1_Q	ict Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50 51	AGO Steam atm residue vac duty Produ Residue Almos Cond Off Gas Waste Water Naphtha Kerosene Diesel AGO PA_1_Q PA_2_Q	ict Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 50 51 52	AGO Steam atm residue vac duty Produ Residue Almos Cond Off Gas Waste Water Naphtha Kerosene Dlesel AGO PA_1_Q PA_2_Q PA_3_Q PA_3_Q	act Name			Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53	AGO Steam atm residue vac duty Produ Residue Almos Cond Off Gas Waste Water Naphtha Kerosene Diesel AGO PA_1_Q PA_2_Q	ct Name	In to SubF	lowSheet	Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 53 55 56	AGO Steam atm residue vac duty Produ Residue Almos Cond Off Gas Waste Water Naphtha Kerosene Dlesel AGO PA_1_Q PA_2_Q PA_3_Q PA_3_Q	ct Name		lowSheet	Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 55 56	AGO Steam atm residue vac duty Produ Residue Almos Cond Off Gas Waste Water Naphtha Kerosene Dlesel AGO PA_1_Q PA_2_Q PA_3_Q PA_3_Q	ct Name	In to SubF	MICS	Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 55 56	AGO Steam atm residue vac duty Produ Residue Almos Cond Off Gas Waste Water Naphtha Kerosene Dlesel AGO PA_1_Q PA_2_Q PA_3_Q	ct Name	In to SubF	lowSheet	Out of Su	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 55 56 57 58 59	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha Kerosene Diesel AGO PA 1 Q PA 2 Q PA 3 Q vac feed	ct Name	DYNA Vessel Dynamic	MICS		bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 56 57 58 60	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha Kerosene Diesel AGO PA_1_Q PA_2_Q PA_3_Q vac feed Vessel		DYNA Vessel Dynamic	MICS Specifications	Kero_SS_Reb	bFlowSheet		
37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 53 56 57 58 59 60 61	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha Kerosene Diesel AGO PA_1_Q PA_2_Q PA_3_Q vac feed Vessel Diameter	(f)	DYNA Vessel Dynamic	MICS Specifications Idenser 13.12 13.12 1	Kero_SS_Reb 5.235	bFlowSheet		
7 8 9 9 0 0 1 1 1 2 2 3 3 4 4 4 5 5 6 6 6 7 7 8 8 8 9 9 9 9 9 9 9 9 9 9 9 9 9	AGO Steam atm residue vac duty Produ Residue Atmos Cond Off Gas Waste Water Naphtha Kerosene Diesel AGO PA_1_Q PA_2_Q PA_3_Q vac feed Vessel	(h)	DYNA Vessel Dynamic	MICS Specifications Idenser 13.12 •	Kero_SS_Reb	bFlowSheet Page 2		



Table 5. 31: holdup details of ADU

	TEAM LND	Case Name: C:\Pi	ogram FilesWyprotechWYSYS 3.1.3	Samples/dyncrude3.hsc
HYPROTECH	Calgary, Alberta	Unit Set: Field	-USGPM	· <u> </u>
Contents instituted	CANADA	Date/Time: Sun	May 03 14:58:57 2009	
			APPRICATE AND AP	***
	Column Sub-Flo	wsheet: T-100 @	Main (continued)	
Volume.0 Liquid Volume Percent	(R3) (%)	1759 30.06 °	70.63 ° 50.05 °	
Lovel Calculator		Horizontal cylinder	Vertical cylinder	
Fraction Calculator		Use levels and nozzles	Use levels and nozzles	
Vessel Delta P	(psi)	9.000 *	0.0000 *	
Fixed Vessel P Spec	(psia)	19.70 Not Active	30.00 Not Active	
Fixed P Spec Active		er Equipment in Column Flo		
	. Our		walled	
	, ig - naggada ad terminin (maga a dada) dami, iday ke iginaw. Ada (yaka dadada ig birina da	PA_1_Cooler @COL1		
		PA_2_Cooler @COL1 PA_3_Cooler @COL1		
		VLV-100 @COL1		
manufacture of the second seco		VLV-101 @COL1		
Andread of the state of the sta		VLV-102 @COL1		
		vac heater @COL1		
		Holdup Details		
		Pressure	Volume	Bulk Liquid Volume
		(psia)	(83)	(113)
*Condenser		19.70	1759	528.5
1_Main TS	a . The statement of th	28.80 28.80	3123	512.8
2 Main TS	A STATE OF THE PROPERTY OF THE PARTY OF THE	28.90	3123	433.2 439,3
3 Main TS 4 Main TS	and section of comments we seem that a comment comment of and analysis and the	28.99	3123	441.1
5 Main TS		29.08	3123	441.1
6_Main TS		29.17	3123	439.9
7_Main TS	COMPANY TO CO. C. AND AND AND AND AND AND AND AND AND AND	29.27 29.36	3123	437.6
8 Main TS		29.45	3123	434.2 401.3
9_Main TS 10_Main TS		29.53	3123	399.2
11 Main TS		29.82	3123	398.1
12 Main TS		29.70	3123	396.8
13 Main TS		29.78	3123	394.8
14 Main TS		29.87 29.95	3123 ·	391.3
15 Main TS	Contract of the first include the contract of	30.03	3123	384.4 487.3
16 Main TS		30.13	3123 •	345.1
17 Main TS 18 Main TS		30.20	3123 •	335.3
19 Main TS		30.27	3123 •	329.4
20 Main TS		30.34	3123	324.1
21 Main TS		30.41	3123 ·	446.2
22 Main TS 23 Main TS		30.58	3123	344.3 337.9
23 Main TS		30.65	3123	331.3
24 Main TS 25 Main TS		30.72	3123 •	324.6
26 Main TS		30.79	3123	314.6
27_Main TS		30.85	3123	290.2
28 Main TS		30.92 31.00	3123	437.8
29 Main TS		29.53	19.97	369.2
1_Kero_SS	The second secon	29.75	19.97	3.800 3.863
2 Kero_SS 3_Kero_SS	*****	30.00	19.97	3.905
	4	30.00	70.63	35.35
The state of the s		HYSYS VS 13 BUILTER	70)	Exce 28 (



Table 5. 32: holdup details of ADU

ū		Case Name:	C:\Program Files\Hyprotech\:HYS\	(S 3.1 3)Samplestdammide3 her
3	HYPROTECH TEAM LND Calgary, Alberta	Unit Set;	Field-USGPM	o o, no o o o o o o o o o o o o o o o o
3	CANADA	Date/Time:	Sun May 03 14:58:57 2009	
5		<u> </u>		
7 8	Column Sub	-Flowsheet: T-100	@Main (continue	ed)
9		Pressure (psle)	Volume (ft3)	Bulk Liquid Volume (ft3)
111	1_Diesel_SS	30.20	124.8	24.19
12 13	2_Diesel_SS 3_Diesel_SS	30.43 30.62	124.8 124.8	23.89
14	1_AGO_SS	30.57	31.20	• 23.33 • 5.158
15	2_AGO_SS	30,71	31.20	5.063
16	3_AGO_SS	30.84	31.20	4.922
17 18		NOTES		
19				
20 21 22 23 24 25 26 27 28 29 30 31 32 33 33 34 35 36 37 38 39 40 41 42 44 45 46 46 46 46 46 46 46 46 46 46 46 46 46				
21	•			
23				
24				
25				
26				
28	•			
29				
30				
31				
32				
34				
35				
36				
37				
39				
40				
41				•
42				
1				
45				
46				
47				
49				
50 51 52 53 55 55 55 58 58 60 61				
52				
5				
55				
56				
57				
5				
60				
62	Drivprojecia (tal	WHYSYSYATIS (BUILD		
	Licensed to: TEAM LND		2. Last 5.00 Sept. 1889 (110)	* Specified by user



Vacuum Distillation Column

This column is used to fractionate the least volatile components of the crude, which are the bottoms of the atmospheric distillation column (TOPPED). The vacuum distillation model represents a column of 9 theoretical trays without a reboiler and a condenser. The column includes three pumparounds that are in charge of recirculating the liquid between trays 2 and 1, 6 and 5 and 9 and 8. The bottoms of atmospheric distillation are fed at tray 9 of the vacuum distillation column. The product streams of the column are located at the upper part (vac ovhd), at tray 2 (LVGO), at tray 5 (HVGO) and at tray 9 (vac residue). This column operates at 1.6 psia of pressure.

All the process flow diagram and simulated results are shown below:

PA_2_Return PA_2_Draw Case: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc Flowsheet: vac tower (COL1)

Figure 5. 3: PFD of vacuum tower



Table 5. 33: connection & monitor of vac tower

1					Case Na	me: C:	Program Files/Hy	protechVHYS	Y\$ 3.1,3\C	asos/vec t	ower.hsc	
2	LIMPOPER	TEAM LND			Unit Set:	Fie	M-USGPM					
4	HYPROTECI	CANADA			Date/Tin	ne: Su	n May 03 14:03:42	2 2009	1 1	√ 100		
5					<u> </u>		· · · · · · · · · · · · · · · · · · ·	·	···	<u> </u>		
7	, ***	Colun	ın Su	b-Flowsh	eet: '	vac tov	ver @Ma	in				
8	<u> </u>		.,		CONNE	CTIONS						
10		January (1944)			Inlot S	tream			11 4, 1	used by 10	. 1	13.5
12	STREAM N			Sta	qe			FROMU	NIT OPER	ATION		
13	vac steam			TS-1 TS-1			Heater				Vac	heater
14 15	vac feed	Conservation			Outlet	Stream	Land to the			4044		
16	STREAM N			Sta	ge			TOU	IT OPERA	TION		
17	vac ovhd			TS-1 TS-1								
18 19	vac residue TopStagePA_Q-Cook		PA	***************************************				* * * * *				
20	LVGO		2_	TS-1								
21	HVGO			TS-1								
22	PA_2_Q-Cooler		PA								************	
23	PA 3_Q-Cooler		PA									
24 25						IITOR						
26							2.47.60			1.12	iii	\$5. (\$4)
27		Specified V		Current Valu		Wt. Error 8.943e-005	Wt. Tol. 1.000e-002 *	Abs. 1	ol. USGPM •	Active	Estimate	Used
28	TopStagePA_Rate(Pa)		JSGPM *	1211 US 140.9 US		8.9436-005	1,000e-002		USGPM •	On	On .	On
29	LVGO Rate		JSGPM *	396.3 US		2.306e-005	1.000e-002 °		USGPM .	On	On On	On On
31	PA 1 Duty(Pa)	300.3		-9.152o+006 E	ttu/tur		1.000e-002 °	0.947	8 Btu/hr •	Off	On	Off
32	Draw Rate	57	32 lb/hr *		lb/hr	-6,637e-005	1.0006-002		205 lb/hr *	On	On	On
33	PA 2 Rate(Pa)		JSGPM .	. 1167 US		-4.260o-006	1.000e-002 °		USGPM *	On _	On	On
34	PA_2_Duty(Pa)	-3.000e+00		-3.000e+007 E		-1.48So-005	1.000s-002 °		USGPM •	On On	On	On
35	PA 3 Rate(Pa)	-1,137e+00	USGPM *	-1.137e+007 I		0.0000	1.000e-002 °		78 Bluffer *	On	On	On On
36 37	PA 3 Duty(Pa)	7,,,,,,,,			SF	ECS						
38			e esta interna	Col			Get Company	27 11 27 1 3	G M H A	11.70	and a street of	2217
39		And Alexander							رط <u>ت شاهد آف</u> رانه	والشاداء والسوادي		es desert
40					TopStage	PA_Rate(Pa	a) 					
41	Fixed / Ranged:	Phod	Printary	Altornate:	Primory	Lower Bo			Upper E	ound:		
43	Spec Type:	Flow Rate	Pumpan	ound:	PA 1	Flow Basi	les .	Volume				4
44					LVG	O Rate	•					
45 46	Fixed / Ranged:	Fhod		/ Alternate:	Primary		(ind:			dund:	Sept.	
47	Stream:	LVGO	Flow Ba	als: A decided to	Volume				1. A. S. S. S.	er a New	1.19.12	i_1, i_2, i_3
48					HVG	O Rate						
49		Fixed	Primari	/ Alternato:	Primary	Lower Bo	und:		. Upper	Sound	er ar i e	97
50	P(ADG) Cardy			ele:	Volume			ara i Ar		T_1U^{-1}		
51 52	- Sugank				PA 1_	Duty(Pa)						
53	1		Carrent et a	autoria.	Primary		trid:	1742355		Allen Acti		
54	Fixed / Ranged:	Fixed	Primary	7 Alematec	PA 1		American de la companya de la companya de la companya de la companya de la companya de la companya de la compa		r-	Boilind:		
55	Spec Type:	Dialy	1. Secretaria	market Carter and		w Rate				<u> </u>		3 - 1 - 2 1
58 57							A 1500 (1 10 17 1 10 1		1.33.			
56 58	Fixed / Ranged:	Fixed		/ Allernato:	Primary Mass		xind:		/ Upper			
59		visc ownd	Flow Bo	ske:			Jee	<i>i.</i>				
80					PA_2	_Rate(Pa)						
61		Fixed	Primary	/ Alternatio:	Primory				Upper	Bound:	ا المام معارض المام المام المام المام المام المام المام المام المام المام المام المام المام المام المام المام ا	لانتارا
63	Honorotech Ltd.		34 1		ysys va	1!3 (Build 4	827) Ya			J. A. Julia	d IReg	
	Licensed to: TEAM LND										* Specified	by user.



Table 5. 34: profiles of vac tower

					Case Name: C:\	Program Files\Hyprote	ch\HYSYS 3.1	,3\Cases\va	c tower.hsc			
3	HYPROTE	TEAM LND Calgary, Al			Unit Set: Field-USGPM							
4 5		CANADA			Date/Time: Sun May 03 14:03:42 2009							
5		Colum	ın Sub-	Flowsh	eet: vac tov	ver @Main	(contin	ued)				
3		University of the			vsheet: vac tower @Main (continued) Column Specification Parameters							
0	Beyfry in the 29 of the 17 out to			Coluir	PA_2_Rate(Pa)	GIS	SERVICE AREA	NC/201011	1000 - 280 - 124 -			
1	DECREE STREET		Pumparound:	A 2014 TOR 12	PA 2 Flow Basis	·	ume	M Marie 12	A CONTRACTOR			
3	Spec Type:	Flow Rate	Pumparound:		PA_2_Duty(Pa)	Vol	une		7 - 1 - 1 - 1 - 1 - 3 - 1 - 3 - 1 - 1 - 1			
14		F	Primary / Alter	note.	Primary Lower Bour	nd:	Linne	r Bound:				
6	Fixed / Ranged: Spec Type:		Pumparound:	nare.	PA_2		Oppe	i Bound.				
7	1 1		Harper		PA_3_Rate(Pa)		-					
18	Fixed / Ranged:	Fixed	Primary / Alter	nate:	Primary Lower Bou	nd:	- Uppe	er Bound:	10,000			
20	Spec Type:	Flow Rate	Pumparound:		PA 3 Flow Basis	C Vo	ume	NA ST				
21					PA_3_Duty(Pa)							
23	Fixed / Ranged:	Fixed	Primary / Alte	nate:	Primary Lower Bou	nd:	Орре	er Bound:				
24	Spec Type:	Duty	Pumparound:	11 11 11 11 11	PA_3		WIR RIN					
25 26					SUBCOOLING							
27				A SAME AND				adde de	has on the election			
28	Degrees of Subcoolin	g										
29 30	Subcool to				User Variables	Assessed to the second		1 - 4				
31					0001 101100							
-												
32			1827		PROFILES							
32 33					General Parameters		118357.29	Transfer				
32 33 34 35	Sub-Flow Sheet:			vac tow		l Stages:		The Asset	9			
-	Sub-Flow Sheet:				General Parameters or (COL1) Number of Profile Estimates Temperature	Net Liquid			9 Net Vapour			
32 33 34 35 36 37 38	Sub-Flow Sheet:				General Parameters or (COL1) Number of Profile Estimates Temperature (F)	kara Miji kati	7)		Net Vapour (lbmole/hr)			
32 33 34 35 36 37 38	Sub-Flow Sheet:		1_TS-1 2_TS-1		General Parameters or (COL1) Number of Profile Estimates Temperature	Net Liquid			Net Vapour (lbmole/hr) 241.6			
332 333 334 335 336 337 338 39	Sub-Flow Sheet:				General Parameters or (COL1) Number of Profile Estimates Temperature (F) 250.0 * 256.8 . 466.0	Net Liquid	2104 0.0000 277.0		Net Vapour (lbmole/hr) 241.6 250.3			
332 333 334 335 336 337 338 339 440 441	Sub-Flow Sheet:		2_TS-1 3_TS-1 4_TS-1	1	General Parameters or (COL1) Number of Profile Estimates remperature (F) 250.0 * 256.8	Net Liquid	2104 0.0000 277.0 261.4		Net Vapour (lbmole/hr) 241.6 250.3 381.2 762.4			
32 333 34 35 36 37 38 39 40 41 42 43	Sub-Flow Sheet:		2_TS-1 3_TS-1		General Parameters or (COL1) Number of Profile Estimates Temperature (F) 250.0 * 256.8 . 466.0	Net Liquid	2104 0.0000 277.0		Net Vapour (lbmole/hr) 241.6 250.3 381.2 762.4 746.8			
32 33 34 35 36 37 38 39 40 41 42 43	Sub-Flow Sheet:		2_TS-1 3_TS-1 4_TS-1 5_TS-1		General Parameters or (COL1) Number of Profile Estimates Temperature (F) 250.0 * 256.8	Net Liquid	2104 0.0000 277.0 261.4 1447 41.57 22.32		Net Vapour (lbmole/hr) 241.6 250.3 381.2 762.4 746.8 1064			
32 33 34 35 36 37 38 39 40	Sub-Flow Sheet:		2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1		General Parameters or (COL1) Number of Profile Estimates [F) 250.0 * 256.8 * 466.0 * 542.4 * 586.5 * 623.7 * 708.2 * 717.2	Net Liquid	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386		Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 1064 1003 983.6			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46	Sub-Flow Sheet:		2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1		General Parameters or (COL1) Number of Profile Estimates remperature (F) 250.0 * 256.8 . 466.0 542.4 585.5 623.7 708.2 717.2 700.0 *	Net Liquid	2104 0.0000 277.0 261.4 1447 41.57 22.32		Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 1064 1003 983.6			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47	Sub-Flow Sheet:		2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1		General Parameters or (COL1) Number of Profile Estimates remperature (F) 250.0 * 256.8	Net Liquid	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386					
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50			2_TS-1 3_TS-1 4_TS-1 5_TS-6_TS-7_TS-1 8_TS-1 9_TS-1		General Parameters or (COL1) Number of Profile Estimates remperature (F) 250.0 * 256.8 . 466.0 542.4 585.5 623.7 708.2 717.2 700.0 *	Net Liquid (ibmole/h	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386 853.4		Net Vapour ((bmole/hr)) 241.6 250.3 381.2 762.4 746.8 1064 1003 983.6 380.8			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51	Stages	Overall Efficience	2_TS-1 3_TS-1 4_TS-1 5_TS-6_TS-7_TS-1 8_TS-1 9_TS-1		General Parameters or (COL1) Number of Profile Estimates 250.0 * 256.8 . 466.0 . 542.4 . 585.5 . 623.7 . 708.2 . 717.2 . 700.0 * EFFICIENCIES Stage Efficiencies	Net Liquid	2104 0,0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000	Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 746.8 1064 1003 983.6 380.8			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52	Stages 1_TS-1	1.0	2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1 9_TS-1	1.000	Stage Efficiencies 1,000	Propens 1.000	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000	Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 746.6 1064 1003 983.6 380.6			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53	Stages	1.0 1.0 1.0	2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1 9_TS-1	1.000 1.000	Content Parameters	Propane 1.000 1.000	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000 1.000 1.000	Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 746.6 1064 1003 983.6 380.6			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54	Stages 1_TS-1 2_TS-1 3_TS-1 4_TS-1	1.0 1.0 1.0	2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1 9_TS-1 000 000 000	1.000 1.000 1.000	Stage Efficiencies 1,000	Propens 1.000	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000 1.000 1.000 1.000	Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 746.8 1003 983.6 380.6 1.00 1.00 1.00 1.00 1.00 1.00 1.00 1			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 50 51 52 53 54 55 56	Stages 1_TS-1 2_TS-1 3_TS-1 4_TS-1 5_TS-1	1.0 1.0 1.0 1.0	2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1 9_TS-1 9_TS-1 000 000 000	1.000 1.000	Content Parameters	Propane 1.000 1.000 1.000	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000 1.000 1.000 1.000	Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 746.6 1000 983.1 380.0 n-Butane 1.000 1.000 1.000 1.000			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 55 55 56 57	Stages 1 _ TS-1 2 _ TS-1 3 _ TS-1 4 _ TS-1 5 _ TS-1 6 _ TS-1	1.0 1.0 1.0 1.0 1.0	2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1 9_TS-1 000 000 000	1.000 1.000 1.000 1.000	Stage Efficiencies 1,000	Propane 1.000 1.000 1.000 1.000 1.000	2104 0,0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000 1.000 1.000 1.000	Net Vapour ((bmole/hr) 241.6 250.3 381.7 762.4 746.1 1006 1000 983.1 380.1 1.00 1.00 1.00 1.00 1.00 1.00 1.00			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54 55 55 56 57 58	Stages 1_TS-1 2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1	1.0 1.0 1.0 1.0 1.0 1.0 1.0	2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1 9_TS-1 000 000 000 000 000 000 000 000 000	1.000 1.000 1.000 1.000 1.000 1.000 1.000	Content Parameters	Propane 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000 1.000 1.000 1.000 1.000	Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 746.6 1000.3 983.1 380.0 1.00 1.00 1.00 1.00 1.00 1.00 1.00			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 55 56 57	Stages 1 _ TS-1 2 _ TS-1 3 _ TS-1 4 _ TS-1 5 _ TS-1 6 _ TS-1	1.0 1.0 1.0 1.0 1.0 1.0 1.0 1.0	2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1 9_TS-1 000 000 000 000 000 000 000 000 000 0	1,000 1,000 1,000 1,000 1,000 1,000	Content Parameters	Propane 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000	2104 0,0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000	Net Vapour ((bmole/hr) 241.6 250.3 381.2 762.4 746.6 1064 1003 983.6 380.6 n-Butane 1.00 1.00 1.00 1.00 1.00 1.00 1.00 1.			
32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53 54 55 55 56 57 58 59	Stages 1. TS-1 2. TS-1 3. TS-1 4. TS-1 5. TS-1 6. TS-1 7. TS-1 8. TS-1	1.0 1.0 1.0 1.0 1.0 1.0 1.0 1.0 1.0 0.0 1.0 1	2_TS-1 3_TS-1 4_TS-1 5_TS-1 6_TS-1 7_TS-1 8_TS-1 9_TS-1 000 000 000 000 000 000 000 000 000 0	1.000 1.000 1.000 1.000 1.000 1.000 1.000	Content Parameters	Propane 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000	2104 0.0000 277.0 261.4 1447 41.57 22.32 2386 853.4	1.000 1.000 1.000 1.000 1.000 1.000 1.000 1.000	Net Vapour (lbmole/hr) 241.6 250.3 381.2 762.4 746.8 1064 1003 983.6 380.8			



Table 5. 35: operation of vac tower

		TEAM LND		Case Nan	ne: C:\	Program Files\Hyprote	:h\HYSYS 3.1.3\Cases\v	ac tower.hsc
1	MYPROTE	Calgary, Albert	ta	Unit Set:	Fie	eld-USGPM		• .
ł	Circerate funds	CANADA		Date/Time	ec Su	n May 03 14:03:42 200		
t		Column	Sub-Flows	heet: v	ac tov	ver @Main	(continued)	
L		Column	Oub-i long		•	··· Ginaii	1 4. 1	
ŀ			Americans	Stage Effi		NBP[0]947*	NBP[0]1009°	ASDOMANOSA
+	Stages	Overall Efficiency 1,000	NBP[0]830° 1.000	146F[GGGG	1.000	1.000	1.000	NBP[0]1062* 1.000
1	4_TS-1	1.000	1,000		1.000	1.000	1.000	1,000
t	5_TS-1 6_TS-1	1.000	1.000		1.000	1.000	1,000	1.000
١	7_TS-1	1.000	1.000		1.000	1,000	1.000	1.000
ı	8_TS-1	1.000	1.000]	1.000	1.000	1,000	1.000
3	9_TS-1	1.000	1.000		1.000	1.000	1.000	1.000
·	Stages	Overall Efficiency	NBP[0]1124*					
į	1_TS-1	1.000	1,000	 				
2	2_TS-1	1.000	1.000					
1	3_TS-1	1.000	1.000					
1	4_TS-1	1.000	1.000				i	ļ
1	5_TS-1	1,000	1.000					ļ
4	6_TS-1	1.000	1.000	+				
4	7_TS-1	1.000	1.000					
1	8_TS-1	1.000	1.000				··· ··· ··· · · · · · · · · · · · · ·	
4	9_TS-1	1,000	1.000		······································			L
4				SOL	VER			
4	The same of the same of the	eggrandelse state	Column Solving	Algorithm	Modified HY	SIM inside Out		AND THE STREETS
4	**************************************	Solving Opti		,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,,			ration Parameters	*** * * * * * * * * * * * * * * * * * *
4	Navious Books	Solving Ope		1000 *	Accelerate	K Value & H Model Pa		Of
4	Maximum Iterations: Equilibrium Error Tok	Manage.		1.000e-05				U
2 3	Heat/Spec Error Tole			5.000e-004				
4	Save Solutions as Init			On		•		
5	Super Critical Handili			Simple K				
8	Trace Level:			Low		· · · · · · · · · · · · · · · · · · ·		
7	Init from Ideal K's:			Off		Dam	ping Parameters	
8	initi	al Estimate Genera	tor Parameters		Azeotrope			O
9	Iterative IEG (Good fo	r Chemicals):		Off	Fixed Dan	nping Factor:		
0								
1					L			
2				SIDE ST	RIPPERS			
3		<u> </u>						
4				SIDE RE	CTIFIERS			
5								
6				PUMP A	ROUNDS			
7		TO NOT BEEN TO THE REAL PROPERTY.		ump Arous	id Summ	rv III	e 2 dell'andres e 15	्रिक्ट्रेस्ट्रेस अस्टर्केस स्
8	7. 11 7. 7. 2. 2.	100.00				Produc		
9		j D	raw Stage	Retur	n Stage	(lbmo		Condenser Duty (Btu/hr)
0	64.4		2 TS-1		1_T	S-1	2095	-9.152e+00
	PA1		6_TS-1		5_T	S-1	1344	-3.000e+00
2	PA_2		9_TS-1		8_T:	S-1	1372	-1.137e+00
3	PA_3			VAD DV	PASSES			-1.1010100
4	1			7AF 01				
٥				DA	TING			
8	İ	-						
8				Trav 9	ections			
9				,,ay o				
<u>2</u>			isa i		71171			1
31	Tray Diameter		(f) 4.921					
			(ff) 0.1640	•				
12	Weir Height		V-7	YSYS V3 1		271		



Table 5. 36: properties of streams of vac tower

Test Test	sc	3.1.3 Cases lvac tower had	ch/HYSYS 3.1,3	ram Files\Hyprote	C:	Case Name				ī	
Column Sub-Flowsheet: Vac tower @Main (continued)				ISGPM	Fie	Unit Set:			PROTECH	_	
Column Sub-Flowsheet: vac tower @Main (continued)		A CONTRACTOR OF THE PARTY OF TH			Si	Date(Timer			CIPECULE INTOVATION	4	
Column Sub-Flowsheet: vac tower @Main (continued)							_				
10 Troy Space		inued)	(continu	ub-Flows	felr Length (ft) 3.9						
11 Tisy Volume			······································			•	3.937	(h)	Length	_	
Disable Heat Loss Calestations No No								(ft)	Space	_	
None										_	
Reting Calculations								ns		_	
Tray Fold Up										_	
10 Vessel 10 Vessel								(03)		-	
17 18 Vessel 19 Diameter 20 Length 21 Volume 22 Volume 23 Volume 24 Volume 25 Boot Length 25 Boot Length 26 Boot Diameter 26 Boot Length 27 Volume 27 Volume 28 Hold Up 29 Tessure Profile 28 Hold Up 29 Tessure Profile 29 PA 1 Cooler PA 2 Cooler PA 3 Cooler PA 4 Cooler PA 4 Cooler PA 5 Cooler PA 5 Cooler PA 5 Cooler PA 5 Cooler PA 6 Cooler PA 6 Cooler PA 6 Cooler PA 7 Cooler PA 8 Cooler PA 8 Cooler PA 8 Cooler PA 8 Cooler PA 8 Cooler PA 8 Cooler PA 8 Cooler PA 9 Cooler PA						Vocasi			Ср	_	
10 Diameter										_	
20 Langth	1.47	198 A N. N. N. N.	aleste per eget	200	Marine .	1. 名為 16年	野部 经验的	Burn in the second	。	18	
										_	
22 Orientation										_	
Vessel has a Boot										_	
24 Boot Ength										_	
25 Boot Length										_	
Color										_	
PA 1_Cooler										_	
PA 1 Cooler				sheet	lumn F	sulpment in C	Other E			27	
Pressure Profile										28	
Pressure (psis)				3 Cooler			PA 2 Cooler		PA 1_Cooler	_	
Pressure (pita)					rofile	Pressure l				_	
33	5.7 2.45	Processon Flores Avenue	Press		sia)	Pressure	71 7112 72	a arreste casa.	- W. T. W. T. P. T. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y. T. Y.	_	
2	- 1 - 1 - 1 - 1 - 1 - 1 - 1 - 1 - 1 - 1			•	_		2 (44-32)			-	
35 3 TS-1 1.650 psis 2.500e-002 psi 36					a	1.625 p				_	
1.700 psia 2.500e-002 psi 3		2.500e-002 psi	2.5		a	1.650 p	<u> </u>			_	
1.725 paia 2.500e-002 pai 39 7 73-1 1.750 paia 2.500e-002 pai 40 8 73-1 1.775 paia 2.500e-002 pai 41 9 78-1 1.500 paia 2.500e-002 pai 41 9 78-1 1.500 paia		2.500e-002 psi	2.5					-1	4_TS	36	
1.750 psia 2.500e-002 psi										37	
1,775 psia 2,500e-002 psi										_	
1.000 psia 1.0		The second secon		 						_	
Pressure Solving Options		200000012 DBI								_	
A				<u>-</u> -					3_10	_	
Add Pressure Tolerance 1,000e-004 * Pressure Drop Tolerance 1,000e-004 * Damping Factor 1,000 * Max Press Rerations					g Optio	ressure Solvi	F			_	
PROPERTIES	100 *	Max Press Iterations	1.000 * Max	actor	Dampi	1.0006-004	Drop Tolerance	00e-004 Pressur	sure Tolerance 1.0	-	
Properties Vaccifeed			<u></u>		MES	PROPER				_	
AB		Verteivere et en	STORY AND A	Neder to the first			Taking and the same			46	
10 10 10 10 10 10 10 10	Septem to	元72世纪由自由的 的首	maria de la Caracte	turid Dheen	A ICECI.	Vancor Die		100 - 0.4 Tipin2	TEQUENCE:	47	
Solution Solution					_					48	
100 100								<u></u>		_	
52 Molar Flow (Bmolefit) 1694 679.9 914.5 52 Molar Flow (Bnhr) 6.981e+005 2.321e+005 4.660e+005 53 Mass Flow (USGPM) 1445 496.2 949.7 54 Skf Ideal Liq Vol Flow (USGPM) -2.192e+005 -1.515e+005 -2.695e+005 55 Molar Enthalpy (Btu/fbmole) -2.192e+005 -443.7 579.0											
53 Mass Flow (b/hr) 6.981e+005 2.321e+005 4.660e+005 54 Ski kleat Liq Vol Flow (USGPM) 1445 496.2 948.7 55 Molar Enthalpy (Btu/fbmole) -2.192e+005 -1.515e+005 -2.695e+005 560.6 443.7 560.0				914.5	_					_	
54 Std Ideal Liq Vol Flow (USGPM) 1445 498.2 948.7 55 Molar Enthalpy (Stuffbrode) -2.192e+005 -1.515e+005 -2.695e+005 57.00					_					_	
SO Motor Entrapy (Business)								(USGPM)			
56 Mass Entitatoy									r Enthalpy	55	
313.0										58	
3/ Mode Europy (Edutation /					_						
30 Miss clauf (Signal)			<u> </u>		_						
59 Heat Flow (82/hr) -3.4554-008 -1.0564-003 -2.4656+008 60 Molar Density ((bmole/h3) 3.9886-004 1.7056-004 8.7356-002										_	
61 Mass Density (ID/R3) 0.1746 5.822-002 44.51					-002	5.822				_	
60 Std Ideal Lip Nass Density (I)-W30 60.24 59.33 61.24				81.24			60.24	(ID/ft3)		_	
(S) Thyprotect (48): 13. (Bysys void (8 (Build (4827))	do 5 864	Pag				YSYS VO.1.8	1786 1			-	



Table 5. 37: properties of streams of vac tower

			Case Name:	C:\Program Files\Hyprote	ch/HYSYS 3.1.3/Casesh	no tower.hec
HYPROTECH	TEAM LND Calgary, Albe	rta	Unit Set:	Field-USGPM		
THE PERSON NAMED IN COLUMN TO PERSON NAMED I	CANADA		Date/Time:	Sun May 03 14:03:42 200	9	16. 36. 34.
	and the second	A. L. Flance	hoote was to	ower Ollein		
				ower @Main		
			Vapour Phase	Liquid Phase		
		Overall 60.17	57,87	61.92		
Liq Mass Density @Std Conc		313.8	228.4	377.3		
TOTAL TITLE OF THE PARTY OF THE	tu/ibmole-F)	0.7187	0,6690	0.7404		
Mass Heat Capacity	(Btu/lb-F)	0.7157	1.564e-002	6.6346-002		
Thermal Conductivity	(Btu/hr-ft-F)		7.234e-003	0.2704		
Viscosity	(dP)			12.43		
Surface Tension	(dyne/cm)	437.9	341.4	509,8		
Molecular Weight	+		0.0008	1,930a-003	********	
Z Factor		Pro	perties : vac ste	ami .	the state of the state of	CONTRACT AND
		Overall	Vapour Phase			
		1,0000	1,0000			
Vapour/Phase Fraction		302.0	302.0			
Temperature:	[]	21.00	21.00			
Pressure:	(psia)	220.3	220.3			
Motar Flow	(ibmole/lu)	3968	3968			
Mass Flow	(ib/hr)	7,941	7,941			
Std ideal Liq Vol Flov/	(USGPM)	-1,018e+005	-1.018e+006			
Molar Enthalpy	(Btu/fbmole)	-1,0188-003	-6652			
Mass Enthalpy	(Btu/lb)		43.59			
Motar Entropy (E	stu/tbmole-F)	43.59	2.419			
Mass Entropy	(Btu/lb-F)	2.419	-2.243e+007			
Heat Flow	(Btu/fur)	-2.243e+007	2,592e-003			
Molar Density	(Ibrnole/R3)	2.592e-003	4,5706-002			
Mass Density	(tb/ft3)	4.670e-002	62.30			
Std Ideal Liq Mass Density	(16/83)	62.30	63,33		-	
Liq Mass Density @Std Con	rd (15/f13)	63.33 8.317	8,317	1		
Motar Heat Capacity (Btu/Ibmole-F)		0,4617			
Mass Heat Capacity	(Btu/lb-F)	0.4617	1,6640-002	<u> </u>		
Thermal Conductivity	(Btu/hr-ft-ff)	1.6646-002	1,407e-002			
Viscosity	(cP)	1.407e-002		1		
Surface Tension	(dyne/cm)		18.02			
Molecular Wigiahi		18.02	0.0040			
Z Factor		0.9912		CONTRACTOR OF THE PARTY.	经济的企业 的企业的	Paristration of the Control
	NT TO THE	AND THE RESIDENCE	Vapour Phase	Liquid Phase		1
		Overall	0.0000			
Vapour/Phase Fraction		0.0000	709.7			
Temperature:		709.7	1,800			
Pressure:	(psia)	1.800	00000			
Molar Flow	(fbmole/hr)	853.A	0.0000			
Mass Flow	(ID/N)	4.450e+005	0.0000			
Std Ideal Liq Vol Flow	(USGPM)	903.5	-1.338e+005			
Molar Enthalpy	(Btu/fbmole)	-2.885e+005	-753.2			
Mass Enthalov	(Btu/b)	-583.3	182.3			
3 Motar Entropy (Btu/lbmole-F)	439.9	1.026			
4 Mass Entropy	(Btu/b-F)	0,8437	0.0000		<u> </u>	+
5 Heat Flow	(Btu/hr)	-2.482e+008				+
6 Molar Density	(ibmole/R3)	8.767e-002	1.4386-004		 	
7 Mass Density	(15/53)	45.72	2.555e-002			
		61.41	59.41		 	
8 Std Ideal Liq Mass Density		62.22	58.74			
O I to Mana Danish ACH Ca		379.1	115.1			
9 Liq Mass Density @Std Co						
O Molar Heat Capacity ((Btu/fbmole-F)	0.7269	9,8478			
	(Btu/fbmole-F) (Btu/fb-F) (Btu/fb-fl-F)	6.935-002	0.6476 1.904e-002 Y/SYS v3U1/3/(Build	6.935e-002		Page 6 of



Table 5. 38: properties of streams of vac tower

I		_	.Case Name: C	C:\Program Files\Hyprote	chi HYSYS 3.1.3 Cases was tower hiss
	HYPROTECH Calgary	LND y, Alberta	Unit Set:	Field-USGPM	
	CANAL		. Date/Time:	Sun May 03 14:03:42 200	9
		ımn Sub-Flows			(continued)
		Pro			
		Overall	Vispour Phase	Liquid Phase	
	Viscosity (c		1.296e-002	0.3440	
ĺ	Surface Tension (dyne/cr		177.7	521,5	
I	Molecular Weight	521.5	0.9972	1.6386-003	
l	Z Factor	1,636e-003 Pr			
I		Overall	Vapour Phase	Liquid Phase	
I		0,0000	0.0000	1.0000	
I	Vapour/Phase Fraction	050.0	256.8	256.8	
ı	Temperature.	4 678	1.625	1.625	
	Pressure: (ps	042.0	0.0000	243.8	
١	project 1 fort	0.005-1004	0.0000	6.325e+004	
ł	M2551 1044	400	0.0000	140.9	
1	Old local Eng Var.	2.402005	-1.027e+005	-2.193e+005	
4	thou diviny)	007.2	-3587	-845.3	
		404.5	52.14	121.5	
5		0.4000	1.820	0.4682	
B	mass Limopy	2017 1007	0.0000	-5.347e+007	
7	TRUST ON		2.1156-004	0.1964	
<u>B</u>			6.059e-003	50.97	
9	Mass Density (Br Std Ideal Liq Mass Density (Br		57.52	55.97	
0	Liq Mass Density @Std Cond (th/		61.01	55.94	
1	Molar Heat Capacity (Blu/Shmole		12.85	138.5	
<u>12</u> 13	mount in the second		0.4487	0.5261	
4	Maria Control		1,418e-002	8.2376-002	
<u>-</u>		cP) 1,099	1.109e-002	1.099	
<u>~</u>	- A Chronic	m) 23.95		23.95	
<u>~</u>		259.5	28,64	259.5	
9	7 Cooler	1,0766-003	0.9991	1.0766-003	
<u>~</u>	HE - FRIEDRICH STATES IN THE PERSON OF	P. P.			anaan heere oo daara
0		Overall	Vapour Phase	Liquid Phese	
11		0.0000	0.0000	1,0000	
12		(F) 585.5	586.5	583.5	
K	Č Co	cia) 1,700	1.700	1,700	
<u> </u>	Molar Flora (Ibrnole		0.0000	475.9	
	(8)	/hr) 1.881e+005	0.0000	1,881e+005	
ï	CISC		0.0000	396.3	
17	On the		-1.422e+005	-2.538e+005	
4	Mass Enthalpy (Bit.		-671.5	-842.1	
4	To the second		185.3	292.0	
ì	Mass Entropy (Btuff		0.8748	0.7387 -1,208e+006	
51	Heat Flow (Bto	vtw) -1.208e+008	1.5246-004	0.1157	
ŝ	Molar Density (Ibmole	300 000		45.72	
ī	3 Mass Density (B)	M3) 45.72	27.22	59.19	
5	A DO MAKE INCIDENT DOLLAR	M3) 59.19		58.97	
3	Lia Mass Density (Std Cond (b)	A33) 58.97	1000	268.6	
9	8 Molar Heat Capacity (Btuffbrnoi			0,6796	
		- 0 0004 002	1.000	6,980-002	
5		0.4022		0.1932	
-	7	(cP) 0.1932		15.06	
5		10.00		. 10.00	
5	O Sústace Tension (dynes	(cm) 15.06			
5	O Surface Tension (dyne)	(em) 15.06 395.3 1.310e-003	211.8	395,3 1,310e-003	



Table 5. 39: properties of streams of vac tower

-				Case Name:	C:\Program Files\H	typrotech\HYS	SYS 3.1.3\Cas	es/vac to	wer.hsc
1	HYPROTECH	TEAM LND Calgary, Alberta		Unit Set:	Field-USGPM				
	TIPRO IEGI	CANADA		Date/Time:	Sun May 03 14:03:	42 2009			
1			A STATE OF THE STATE OF	344					
1		Column S	Sub-Flows	heet: vac t	ower @Ma	ain (co	ntinuec	i)	
-				sheet: vac tower @Main (continued)					
4			Overall	vac ovh Vapour Phase	Ĭ	SALES OF FRANCE			20 A Promis
1	Vapour/Phase Fraction		1.0000	1.0000	1				
2	Temperature:	(F)	225.3	225.3					
3	Pressure:	(psia)	1.600	1.600	ļ			-	
4	Molar Flow	(lbmole/hr)	241.6	241.6 5732				+	
5	Mass Flow	(lb/hr)	5732 °	12.15					
6	Std Ideal Liq Vol Flow	(USGPM) (Btu/lbmole)	-1.025e+005	-1.025e+005					
8	Molar Enthalpy Mass Enthalpy	(Btu/lb)	-4318	-4318					
9	Molar Entropy (E	Stu/Ibmole-F)	49.84	49.84	<u></u>				
0	Mass Entropy	(Btu/lb-F)	2.101	2.101 -2.475e+007		_		-	
21	Heat Flow	(Btu/hr)	-2.475e+007 2.179e-004	2.179e-004				_	
22	Molar Density	(lbmole/ft3)	5.169e-003	5.169e-003				_	
23	Mass Density Std Ideal Liq Mass Density	(lb/ft3)	58.83	58.83			CONTRACTOR		
24	Lig Mass Density @Std Con		61.77	61.77					
26		Btu/lbmole-F)	10.56	10.56 0.4450				-	
27	Mass Heat Capacity	(Btu/lb-F)	0.4450	1.370e-002	-			_	
28	Thermal Conductivity	(Btu/hr-ft-F)	1.370e-002 1.017e-002	1.017e-002					
29	Viscosity	(cP)	1,0176-002		1				
30	Surface Tension Molecular Weight	(dyne/cm)	23.73	23.73					
31	Molecular Weight	(dyne/cm)	23.73 0.9991	23.73 0.9991					
	The state of the s	(dyne/cm)							
31 32 33 34	Molecular Weight Z Factor	(dyne/cm)		0.9991	The	e composition	option is selec	cted	
31 32 33 34 35	Molecular Weight	(dyne.din)		0.9991 SUMMARY		composition	option is selec	cted	
31 32 33 34 35 36	Molecular Weight Z Factor	(dyneicin)	0,9991	0.9991 SUMMARY Molar Feed Compositi		e composition	option is selec	cted	
31 32 33 34 35 36	Molecular Weight Z Factor	vac feed 1.594400e+03	0,9991	0.9991 SUMMARY Molar Feed Compositi		e composition	option is selec	cted	
31 32 33 34 35 36 37 38	Molecular Weight 2 Factor Flow Basis:	vac feed 1.594400e+03	0.9991 vac s 220.	0.9991 SUMMARY Molar Feed Compositi		e composition	option is selec	cted	
31 32 33 34 35 36 37 38 39	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane	vac feed 1.594400e+03 0.0000	0,9991	0.9991 SUMMARY Molar Feed Compositi		o composition	option is selec	cted	
31 32 33 34 35 36 37 38 39 40	Molecular Weight Z Factor Flow Basis: Flow Rate (Ibmole/hr) Methane Ethane	vac feed 1.594400e+03	0.9991 vac s 220.	0.9991 SUMMARY Molar Feed Compositi		e composition	option is selec	cted	
31 32 33 34 35 36 37 38 39 40 41 42	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane Elhane Propane	vac feed 1.594400e+03 0.0000 0.0000	0,9991 vac 5 220: 	SUMMARY Molar Feed Compositi team 2752		e composition	option is selec	cted	
31 32 33 34 35 36 37 38 40 41 42 43	Molecular Weight Z Factor Flow Basis: Flow Rate (Ibmole/hr) Methane Ethane	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000	0,991 vac s 220.	O.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is selec	cted	32032
31 32 33 34 35 36 37 38 39 40	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane Elthane Propane i-Butane	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000	0,991 vac s 220.	0.9991 SUMMARY Molar Feed Compositi team 27- 2000 2000 2000 2000 2000 2000 2000		o composition	option is selec	cted	
31 32 33 34 35 36 37 38 39 40 41 42 43 44	Molecular Weight Z Factor Flow Basis: Flow Rate (Ibmole/hr) Methane Ethane Propane i-Butane n-Butane H2O NBP[0]49*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0007	0,991 vac s 220 0,0 0,0 0,0 0,0 0,0 0,0 0,0 0,0 0,0	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is selec	cted	
31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane Elhane Propane i-Butane n-Butane H2O NBP[0]49* NBP[0]79*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0007 0.0001	0,991 vac s 220.	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is selec	cled	
31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane Elhane Propane i-Butane n-Butane H2O NBP[0]49* NBP[0]79* NBP[0]111*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0007	0,9991 vac s 220. 0,0 0,0 0,0 1,0 0,0 0,0 0,0	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is select	cted	
31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48	Molecular Weight Z Factor Flow Basis: Flow Rate (Ibmole/hr) Methane Ethane Propane i-Butane n-Butane H2O NBP[0]49* NBP[0]79* NBP[0]111* NBP[0]144*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001	0,9991 vac s 220. 0.0 0.0 0.0 0.0 1.0 0.0 0.0 0	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is select	cted	
31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane Elhane Propane i-Butane n-Butane H2O NBP[0]49* NBP[0]79* NBP[0]11* NBP[0]114* NBP[0]176*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0002 0.0002	0,9991 vac s 220. 0.0 0.0 0.0 0.0 0.0 0.0 0.	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is select	cted	
31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane Ethane Propane i-Butane n-Butane H2O NBP[0]49 NBP[0]79* NBP[0]11* NBP[0]11* NBP[0]176* NBP[0]176*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0002 0.0002 0.0003 0.0004	0,9991 vac s 220. 0,0 0,0 0,0 0,0 0,0 0,0 0,0 0,0 0,0	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is select	cted	
31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50 51 52 53	Molecular Weight Z Factor Flow Basis: Flow Rate (Ibmole/hr) Methane Ethane Propane i-Butane n-Butane h2O NBP[0]49* NBP[0]11* NBP[0]145* NBP[0]145* NBP[0]145* NBP[0]145* NBP[0]145* NBP[0]208* NBP[0]240* NBP[0]202*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0002 0.0002 0.0003 0.0004 0.0006	0,9991 Vac s 220. 0,0 0,0 0,0 1,0 0,0 0,0 0,0 0	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is selec	cted	
31 33 33 33 33 33 33 33 33 40 40 41 41 42 43 44 44 44 45 55 55 55 55 55 55 55 55 55	Molecular Weight Z Factor Flow Basis: Flow Rate (Ibmole/hr) Methane Elhane Propane i-Butane n-Butane H2O NBP[0]49* NBP[0]79* NBP[0]11* NBP[0]11* NBP[0]144* NBP[0]176* NBP[0]176* NBP[0]208* NBP[0]208* NBP[0]2040* NBP[0]272* NBP[0]304*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0002 0.0002 0.0003 0.0004 0.0006 0.0008	0,9991 Vac 5 220. 0,0 0,0 0,0 0,0 0,0 0,0 0,	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is solec	cted	
31333333333333333333333333333333333333	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane Ethane Propane i-Butane n-Butane H2O NBP[0]49* NBP[0]79* NBP[0]11* NBP[0]144* NBP[0]144* NBP[0]176* NBP[0]208* NBP[0]208* NBP[0]240* NBP[0]272* NBP[0]336*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0002 0.0002 0.0003 0.0004 0.0006	0,9991 Vac 5 220. 0,0 0,0 0,0 1,0 0,0 0,0 0,0	0.9991 SUMMARY Molar Feed Compositi Islam 27752		o composition	option is solec	cted	
31333333333333333333333333333333333333	Molecular Weight Z Factor Flow Basis: Flow Rate (ibmole/hr) Methane Ethane Propane i-Butane n-Butane h2O NBP[0]49* NBP[0]11* NBP[0]144* NBP[0]144* NBP[0]145* NBP[0]240* NBP[0]272* NBP[0]272* NBP[0]336* NBP[0]368*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0002 0.0002 0.0003 0.0004 0.0006 0.0008 0.0008	0,9991 Vac 5 220.	0.9991 SUMMARY Molar Feed Compositi team 27752		o composition	option is soler	cted	
31333333333333333333333333333333333333	Molecular Weight Z Factor Flow Basis: Flow Rate (Ibmole/hr) Methane Elthane Propane i-Butane n-Butane n-Butane h2O NBP[0]49* NBP[0]11* NBP[0]144* NBP[0]176* NBP[0]144* NBP[0]208* NBP[0]208* NBP[0]240* NBP[0]272* NBP[0]306* NBP[0]366* NBP[0]368*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0002 0.0002 0.0003 0.0004 0.0006 0.0008 0.0001	0,9991 Vac 5 220. 0,0 0,0 0,0 0,0 0,0 0,0 0,	0.9991 SUMMARY Molar Feed Compositi team 2752		o composition	option is soler	cted	
31333333333333333333333333333333333333	Molecular Weight Z Factor Flow Basia: Flow Rate (ibmole/hr) Methane Elhane Propane i-Butane n-Butane H2O NBP[0]49* NBP[0]11* NBP[0]144* NBP[0]176* NBP[0]176* NBP[0]28* NBP[0]240* NBP[0]240* NBP[0]364* NBP[0]365* NBP[0]365* NBP[0]365* NBP[0]365* NBP[0]403*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0002 0.0002 0.0003 0.0004 0.0006 0.0008 0.0011 0.0016 0.0002 0.0002	0,9991 Vac 5 220. 0.0 0.0 0.0 0.0 0.0 0.0 0	0.9991 SUMMARY Molar Feed Compositi team 2752 2		o composition	option is solec	cted	
31 32 33 34 35 36 37 38 39 40 41 42 43 44 45 46 47 48 49 50	Molecular Weight Z Factor Flow Basia: Flow Rate (ibmole/hr) Methane Elhane Propane i-Butane n-Butane H2O NBP[0]49* NBP[0]11* NBP[0]176* NBP[0]176* NBP[0]176* NBP[0]28* NBP[0]240* NBP[0]240* NBP[0]364* NBP[0]365* NBP[0]365* NBP[0]365* NBP[0]368* NBP[0]403* NBP[0]403* NBP[0]403* NBP[0]404*	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0001 0.0002 0.0002 0.0008 0.0001 0.0001 0.0008 0.0001 0.0008 0.0001 0.0008 0.0008 0.0008 0.0008 0.0008 0.0008 0.0008 0.0008 0.0008 0.0008 0.0008	0,9991 Vac s 220. 0,0 0,0 0,0 0,0 0,0 0,0 0,	0.9991 SUMMARY Molar Feed Compositi team 2752 2- 2000 2000 2000 2000 2000 2000 2000		o composition	option is solec	cted	
313 333 334 335 336 337 338 339 440 441 442 443 444 444 445 550 550 550 550 550 550 550	Molecular Weight Z Factor Flow Basis: Flow Rate (Ibmole/hr) Methane Elthane Propane I-Butane H2O NBP[0]49° NBP[0]11° NBP[0]11° NBP[0]144° NBP[0]144° NBP[0]208° NBP[0]240° NBP[0]306° NBP[0]306° NBP[0]306° NBP[0]306° NBP[0]400° NBP[0]403° NBP[0]400° NBP[0]403° NBP[0]404° NBP[0]404° NBP[0]406° NBP[0]406° NBP[0]406° NBP[0]406° NBP[0]406° NBP[0]406°	vac feed 1.594400e+03 0.0000 0.0000 0.0000 0.0000 0.0001 0.0001 0.0002 0.0002 0.0003 0.0004 0.0006 0.0008 0.0011 0.0016 0.0002 0.0002	0,9991 Vac s 220. 0,0 0,0 0,0 0,0 0,0 0,0 0,	0.9991 SUMMARY Molar Feed Compositi team 2752 2		o composition	option is solec	cted	



Table 5. 40: feed composition ad flow rate of vac tower

1		TEAM LND		ase Name: C.\Program Files\Hyprotech\HYSYS 3,1,3\Cases\vac tov.er.hs	1 (10)				
1	HYPROTECH	Calgary, Alberta	U	nt Set: Field-USGPM	1411				
1	CIFCEVEL ILABORATION	CANADA	D	Date/Time: Sun May 03 14:03:42 2009					
t	NAMES AND ADDRESS.		Marine		400				
]		Column Sub	-Flowshee	t: vac tower @Main (continued)					
4	1. A. F	THE REPORT OF THE PARTY OF THE		DUMMARY					
0		· ·		SUMMARY					
1	NI KADBAKATO TEGOT D	vac feed	vac steam	Design Color of the At Supple Control of the Atlantic Color	SEGM				
2	NBP[0]592*	0.0182	0.0000						
3	NBP[0]624*	0.0221	0.0000						
4	NBP[0]656*	0.0265 0.0316	0.0000						
5	NBP[0]688*	0.0366	0.0000						
7	NBP[0]720* NBP[0]752*	0.0410	0.0000						
a	NBP[0]784*	0.0465	0.0000						
9	NBP[0]830*	0.0959	0.0000						
0	NBP[0]888*	0.0929	0.0000						
1	NBP[0]947*	0.0924	0.0000						
22	NBP[0]1009*	0.1181	0.0000						
23	NBP[0]1062*	0.1614	0.0000						
4	NBP[0]1124*	0.1586		lolar The composition option is selected					
5	Flow Basis:	19.45.75.75.75.75.66.00		Feed Flows	(SINV)				
6		yac feed	vac steam						
8	Flow Rate (Ibmole/hr)	1.594400e+03	220.2752		-				
29	r jow rease (ionocura)								
30	Methane (lbmole/hr)	0.0001	0.0000						
31	Ethane (lbmole/hr)	0.0003	0.0000						
32	Propane (lbmole/hr)	0.0088	0.0000						
33	i-Butane (lbmole/hr)	0.0093	0.0000	,	********				
34	n-Butane (ibmole/hr)	0.0381	220.2752						
35	H2O (lbmole/hr)	10.6450 0.1117	0.0000		***				
36	NBP[0]49* (lbmole/hr)	0.1657	0.0000						
37	NBP[0]79* (lbmole/hr)	0.2086	0.0000						
38	NBP[0]111* (lbmole/hr) NBP[0]144* (lbmole/hr)	0.2687	0.0000						
40	NBP[0]176* (lbmole/hr)	0.3615	0.0000						
41	NBP[0]208* (lbmole/hr)	0.5205	0.0000						
42	NBP[0]240* (lbmole/hr)	0.7168	0.0000						
43	NBP[0]272* (lbmole/hr)	0.9808	0.0000						
44	NEP[0]304* (lbmole/hr)	1.3246	0.0000						
45	NBP[0]336* (lbmole/hr)	1,7891	0.0000						
46	NBP[0]368* (lbmole/hr)	2.4806 3.5796	0.0000		-				
47	NBP[0]400* (lbmole/hr)	5,6996	0.0000						
48	NBP[0]433* (ibmole/hr)	9.2951	0.0000						
49	NBP[0]464* (ibmole/hr)	13.5894	0.0000						
50 51	NBP[0]496* (lbmole/hr) NBP[0]528* (lbmole/hr)	17.9760	0.0000						
52	NBP[0]560* (lbmole/hr)	22.9605	0.0000						
53		29.0422	0,0000						
54	NBP[0]624* (lbmole/hr)	35.2451	0.0000						
55	NBP[0]656* (lbmole/hr)	42.2106	0.0000						
56	NBP[0]688* (lbmole/hr)	50.3988	0.0000						
57	NBP[0]720* (ibmole/hr)	58.2932	0.0000		*********				
58	NBP[0]752* (lbmole/hr)	65.3380 74.2190	0.0000						
59	MANAGEN WAS CORES OF THE PERSON OF THE PERSO	152.8487	0.0000	Management of the second of th					
60	And notice the property of the	148.1192	0.0000						
	NBP[0]888* (lbmole/hr)	And the second s	0.0000						
61	NBP[0]947* (lbmole/hr)	147.3452	THE RESERVE AND ADDRESS OF THE PARTY NAMED IN COLUMN TWO IS NOT THE OWNER.	y3.1.3 (Build 4827) Pa	ige 9 of				

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Table 5. 41: product composition ad flow rate of vac tower

T			Case Nar	ne: C:\Program File	s\Hyprotech\HYSYS 3,1,3\Car	ses/vac tower.hsc
١,	HYPROTECH	TEAM LND Calgary, Alberta	Unit Set:	Field-USGPM		
1		CANADA	Date/Tim	e: Sun May 03 14:0	3:42 2009	
1		and the second state of the second	Management of the second			
1		Column Suh	-Flowsheet: v	ac tower @N	Main (continue	d) (t
1		Columnicas		9.		Parties and the state of
1	The second secon		SUMM	ARY		
5		The state of the s	vac steam	and Colored Services	v two traces are to trace to the page to the co	TO A SHIP OF THE RELEASE TO
4	When the Land of t	vac feed	0.0000		S Management of the same of th	
-	NBP[0]1009* (lbmole/hr)	188,3519 257,3472	0.0000			
	NBP[0]1062* (lbmole/hr)	257.3472	0.0000			
5	NBP[0]1124* (lbmole/hr)	252.5100	Prod			
5	Flow Basis:		Molar		he composition option is select	eted
7 10	MARKET STATES	College College College College	Product Cor	HVGO	rless	
в	In the Line of the State of the	vac ovhd	LVGO	475.9398	vac residue 853,3920	SC SCHOOL AT SETUDIOUS FORM
9 1	Flow Rate (lbmole/hr)	241,5725	243.7709		853,3920	
0	24.0° (17) [2.0° (0.0000	0.0000	0,0000	0.0000	
1	Methane	0.000	0.0000	0.0000	0.0000	
2	Ethane	0.0000	0.0000	0.0000	0.0000	
3	Propane	0.0000	0.0000	0.0000	0.0000	
4	i-Butane	0.0002	0.0000	0.0000	0.0000	
5	n-Butane	0.9538	0,0006	0.0002	0.0003	
7	H2O	0.0005	0.0000	0.0000	0.0000	
8	NBP[0]49* NBP[0]79*	0.0007	0.0000	0.0000	0.0000	
9	NBP[0]111*	0.0008	0.0000	0.0000	0.0000	
10	NBP[0]144*	0.0011	0.0000	0.0000	0.0000	
31	NBP[0]176*	0.0014	0.0000	0.0000	0.0000	
32	NBP[0]208*	0.0020	0.0001	0.0000	0.0000	
33	NBP[0]240*	0.0027	0.0006	0.0000	0.0000	
34	NBP[0]272*	0.0035	0.0013	0.0000	0.0000	
15	NBP[0]304*	0.0042 0.0047	0.0027	0.0000	0.0000	
36	NBP[0]336*	0.0048	0.0053	0.0000	0.0000	-
37	NBP[0]368*	0.0047	0.0099	0.0001	0,0000	
38	NBP[0]400*	0.0043	0.0187	0.0002	0,0000	
40	NBP[0]433*	0.0038	0.0335	0.0004	0.0000	
11	NBP[0]464* NBP[0]496*	0.0029	0.0513	0.0008	0.0000	***************************************
42	NBP[0]528*	0.0018	0.0689	0.0015 0.0027	0.0001	
43	NBP[0]560*	0.0010	0.0876	0.0049	0.0002	***
44	NEP[0]592*	0.0006	0.1063	0.0088	0.0004	
45	NBP[0]624*	0.0003	0.1385	0.0162	0.0009	***************************************
16	NBP[0]656*	0.0001	0.1404	0.0308	0.0017	
47	NBP[0]688*	0.0000	0.1150	0.0575	0.0034	The state of the s
48	NBP[0]720*	0.0000	0.0632	0.0937	0.0062	
49	NBP[0]752*	0.0000	0.0226	0.1242	0.0113	
50	NBP[0]784*	0.0000	0.0053	0.2440	0.0415	
51	NBP[0]830*	0.0000	0.0002	0.1795	0.0734	
52	NBP[0]888*	0.0000	0.0000	0.1112	0.1107 0.1823	
53	NBP[0]947* NBP[0]1009*	0.0000	0.0000	0,0689 0.0421	0.1823	
55	NBP[0]1062*	0.0000	0.0000	0.0124	0.2894	
56	NBP[0]1124*	0.0000	Molar		The composition option is sele	ected
57	Flow Basis:	The second second second		t Flows		A CONTRACTOR OF THE
58	OF THE PROPERTY OF	as orbid	LVGO	HVGO	vac residue	
59		vac ovhd 241.5725	243.7709	475.9398	853.3920	*
60	Flow Rate (lbmole/hr)	241.5125				
61	the state of the s	0.0001	0.0000	0.0000	0.0000	•

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Table 5. 42: product composition ad flow rate of vac tower

+	W.F.	CANADA	Date/T	ime; Sun May 03 14:0	3:42 2009	
		Column Sub	-Flowsheet:	vac tower @N	lain (continu	ed)
+			TOTAL DESCRIPTION	IMARY		
-	and a continue to the top to the first of the	L.	LVGO	HVGO	The second section of the second	R. Dieser von Stellen von
	Ethane (lbmole/hr)	0.0003	0.0000	0.0000 -	vac residue 0.0000	MERCANISM SOCIETA SERVICIO COLO
	Propane (lbmole/hr)	0.0088	0.0000	0.0000 -	0.0000	•
	i-Butane (Ibmole/hr)	0.0092	0.0000	0.0000 -	0.0000	•
	n-Butane (lbmole/hr)	0.0379	0.0002	0.0000 •	0.0000	•
100	H2O (lbmole/hr)	230.4175	0.1582	0.0737	0.2707	•
	NBP[0]49* (lbmole/hr)	0.1108	0.0008	0.0001	0.0000	•1
	NBP[0]79* (lbmole/hr)	0.1639	0.0016	0.0002	0.0000	•
	NBP[0]111* (lbmole/hr)	0.2053	0.0030 0.0058	0.0003 -	0.0000	
ł	NBP[0]144" (lbmole/hr)	0.2625 ° 0.3490 °	0.0058	0.0005	0.0000	-
ŀ	NBP[0]176* (lbmole/hr)	0.3490	0.0274	0.0013	0.0000	
ł	NBP[0]208* (lbmole/hr) NBP[0]240* (lbmole/hr)	0.6523	0.0623	0.0022	0.0000	•
	NBP[0]272* (lbmole/hr)	0.8354	0.1415	0.0039	0.0001	•
Ī	NBP[0]304" (lbmole/hr)	1.0067	0.3111	0.0067	0.0001	•
i	NBP[0]336* (/bmole/hr)	1.1257 •	0.6515	0.0116 •	0.0003	•
1000	NBP[0]368* (lbmole/hr)	1.1698	1.2892	0.0210	0.0006	•
I	NBP[0]400* (lbmole/hr)	1,1277	2.4105	0.0402	0.0013	•
L	NBP[0]433* (lbmole/hr)	1.0483	4.5618	0.0863	0.0032	•
ı	NBP[0]464* (lbmole/hr)	0.9295 *	8,1670 12,4997	0.1905	0.0081	-
ł	NBP[0]496* (lbmole/hr)	0.6895 ° 0.4352 °	16.7987	0.7021	0.0187	
ł	NBP[0]528" (lbmole/hr)	0.2490	21.3530	1.2743 •	0.0843	•
ł	NBP[0]560* (lbmole/hr) NBP[0]592* (lbmole/hr)	0.1347	26.4004	2.3293 •	0.1778	•
t	NBP[0]624* (lbmole/hr)	0.0660	30.6290	4.1867 -	0.3634	•
i	NBP[0]656* (lbmole/hr)	0.0295	33.7507	7.6919	0.7384	
Ī	NBP[0]688* (lbmole/hr)	0.0118 •	34.2345	14.6611	1,4915	•
	NBP[0]720* (lbmole/hr)	0.0037 -	28.0310	27.3714 •	2.8872	•
ď	NBP[0]752* (lbmole/hr)	0.0007 •	15.4181	44.5938	5.3254	•
N	NBP[0]784* (lbmole/hr)	0,0001	5.5167	59,0902	9.6120	•
5	NBP[0]830* (lbmole/hr)	0.0000	1.2886 0.0458	116.1238	35,4363	•
	NBP[0]888* (lbmole/hr)	0.0000	0.0458	85.4415	62.6320	
	NBP[0]947* (lbmole/hr)	0.0000	0.0000	52.9088 32.8010	94.4353 155.5508	
	NBP[0]1009* (ibmole/hr)	0.0000	0.0000	20.0321	237.3151	•
	NBP[0]1062* (lbmole/hr)	0.0000	0.0000	5.9110	246.9994	•
H	NBP[0]1124* (lbmole/hr) Flow Basis:		Molar	Th	e composition option is se	elected
1000	HOW Besis.	THE PROPERTY OF STREET	Product	Recoveries		AME THAT I THAT I GOT
i		vac ovfid	LVGO	HVGO	vac residue	
100	Flow Rate (lbmole/hr)	241.5725	243.7709	475.9398	853.3920	
			0.0303			
	Methane (%)	99.9483	0.0303 0.0918	0.0212	0.0001	
	Ethane (%)	99.8721 99.7342	0.2117	0.0359	0.0002	
	Propane (%)	99,5311	0.3947	0.0536	0.0003	The second secon
	i-Butane (%)	99,4260	0.4922	0.0737	0.0007	
	n-Butane (%)	99.7823	0.0685	0.0319	0.0007	
	H2O (%)	99.1980	0,6965	0.1045	0.0010	
10	NBP[0]49* (%) NBP[0]79* (%)	98.9162	0.9626	0.1200	0.0010	
H	NBP[0]111* (%)	98.4358	1.4214	0.1412	0.0012	
w		The same of the sa	2.1602	0.1692	0.0020	
	NBP[0]144* (%)	97.6686		0.1002	0.0020	



Table 5. 43: column profile of vac tower

		TEAM LND	Unit Set:	Case Name: C:Program Files/Hyprotech/HYSYS 3.1.3/Cases/wac tower.hac Unit Set: Field-USGPM							
	HYPROTECH	Calgary, Alberta CANADA				03 14:03:42 2009					
4				Date/Time	: Su	May	03 14:03:4	2 2009		<u> </u>	<u>:, tean.' </u>
		Calimon	Sub-Flowsh	reet: V	ac tow	/er	@Ma	in (c	ontinu	ed)	
		Colmini	3ub-1 101131				<u>e</u>				
				SUMM	ARY						
	ner and shirt	Colores and Color	LVGC	o a single	UP LO HV	GO!	CTT I	\ \va	o residue	はなご	THE THREET
	NESP(0)208* (%)	94,4803	5.268	8				*****	0.0036		
	NBP[0]240* (%)	90.9962	8.688			103 928			0.0051 0.0073		
1.	The fact of the beautiful to	85.1770	14.422 23.48			021	•• • • • •		0.0105		
12	NBP(0)304* (%)	76.0001 62.9205	36.414		0.6	493			0.0155		
S.	NBP(0)387 (%)	47.1574	51.97			485			0.0233		
	NEP (OHOO! (%)	31.5023	67.33 80.03			230 142			0.0355 0.0553		
٠	NBP(0)4331 (%)	18,3926	87.86			497			0.0866		
	NEPO464 (%)	5.0738	91.98			073			0.1376		
100	NBPUSZ8 (%)	2.4207	93.45			1060 5498			0.2223		****
Į.	N8890560: (%)	1.0847	92.99)204			0.6122		the same to be recovered about a spirit before the second about the second
Z.	NBPOESZ (%)	0.4639 0.1873	86.90			8787			1.0312		
Y.	NEP(0)824*(%)	0.0700	79.95			2226			1.7494		
	NBPIDEST(%)	0.0233	67.92 48.08			0902 9546			2.9594 4.9530		······································
1	ANSE(0)7201(94)	0,0083	23.59			2510			8.1505		
3	NBP(0/54*(%)	0,0001	7.43		79.	6160			12.9508		
2	(MEPORSO (%)	0.0000	0.84			9730			23.1839		
Ē	NEP(0)888*.(%)	0.0000	0.03			.6842 .9080			42.2849 64.0912		
	NEPPOPATY(%)	0.0000	0.00			.4148			82.5852		
	NEPROTOSE (%)	0.0000	0.00		7.	7841			92.2159		
	NEPPOTIZE (%)	0.0000	0.00	00	2.	3372			97.6628		
Ī				COLUMN	PROFILES	;					
4		8.708 Reb	oil Ratio:	2.916	The F	lows (Option is Se	elected	Flow B		Mol
Ę	Reflux Ratio:	ATTACAMENT TO SERVICE OF	tie i	Columb Pro	nes For		ores esta	erecords	rita de la companya de la companya de la companya de la companya de la companya de la companya de la companya d	erente:	La Company
Ī	PART TO THE PART OF THE PART O	rigaratus (F)	Primarit (pain)	21	04	السروانية	TAND TOOLES	CH PI SO I	2095		241.5
1.	1_TS-1	225.3 256.8	1,600	0.0	************	<u> </u>	250.3				2339
-	2_TS-1 3_TS-1	468.0	1.650		7.0	381.2		}			
-	4_TS-1	542.4	1.675	261,4 1447			762.4 748.8		1344		47- 4
	5_TS-1	585.5	1.700		41.57		1064				475.9 1344
4	6_TS-1	623.7 708.2	1.750	22.32			1003				
-	7_TS-1 8_TS-1	717.2	1.775	23	86	 	983.6		2967 220.3		
	9 TS-1	709.7	1,800	Column Pro	files Ener	dV.	380.8				2226
		Tôm	enter (z)	Light Erous	. Carlo mo	77					teef Loan (Builte)
	1_TS-1	William Co.	225.3	-2.232	e+005			1.0250+1	005		
1	2_TS-1	256.8			-2.193e+005 -2.260e+005		-1.027e+005 -1.213e+005				
1	3_TS-1	466.0 542.4			-2.294e+005		-1.428e+008				
4	4_TS-1	585.5			-2.538e+005		-1.4226+005				
7	5_TS-1 6_TS-1	623.7		-2.552e+005			-1.51Se+005				
9	7_T6-1	708.2		-2.788e+005			-1.472e+005				
0 8 TS-1		717.2		-2.811e+005 -2.885e+005			-1.456e+005 -1.338e+006				
4	9 TS-1										
2	Hypotheeli (St)	er an extended to the		YSYS VA	3 (Bind 4	327)					Page (2 of



Chapter 6 Conclusions

A rigorous nonequilibrium model has been developed where each phase in the froth and disengagement zone is considered as a separate, variable, completely mixed holdup and only mechanical equilibrium is assumed (equal pressure over the tray). Mass transfer occurs between the vapor and liquid in the dispersion on the tray. The nonequilibrium model includes tray sizing parameters and mass transfer models and it is observed that these have a direct and significant influence on the column dynamics. Thus, the nonequilibrium model has the potential to include tray sizing parameters as part of the column design, control, and optimization. Without efficiencies the model is predictive, no estimates were needed to describe the performance of an existing industrial column, just tray design layout and operational specifications. Tray layout specifications are not required for a nonequilibrium simulation, they can be generated by using the design-mode during a steady-state simulation.

Nonequilibrium simulations show that the (back-calculated) component Murphree tray efficiencies for multicomponent systems are unequal and can become completely different given a small change in specifications. If the component efficiencies are unequal and they change with respect to specifications then they cannot be used in dynamic column simulations since no model is available to rigorously compute the efficiencies. The difference between equilibrium with a constant efficiency and nonequilibrium simulation transients can be pronounced, both qualitatively and quantitatively. These differences are due to both mass and heat transfer limitations which the equilibrium model ignores. There are also significant differences in dynamic behavior of columns at higher pressures simulated with models that include or ignore the vapor holdup above the froth. In general, it also takes longer to reach a new steady-state for the nonequilibrium column simulation compared to the equilibrium simulation.

Mass transfer models are developed that account for cross flow effects on large distillation trays while avoiding the pitfalls that can strike while employing overall mass transfer coefficients in the calculation of interphase mass transfer rates. New plug and dispersion flow models for the nonequilibrium column model are proposed and found to predict tray efficiencies in general agreement with FRI experimental data on large scale equipment. Several methods of evaluating binary mass transfer coefficients in distillation have been evaluated. Of the methods tested, that of Chan and Fair (1984) provides the best predictions of column performance. Additional evidence that the Maxwell-Stefan equations should be used in the calculation of mass transfer

rates in distillation is provided by comparing the predictions of the nonequilibrium model with the new flow and mass transfer models with a simpler model based on all components having an equal facility for mass transfer. Column designs obtained with the simple model can be very different (in terms of numbers of stages, optimal feed, sidestream and controller locations) from those obtained with the more rigorous approach. However, there is little difference in the dynamic behavior of columns. This is in contrast to the use of different mass transfer coefficient correlations which have differing dynamic characteristics. Dynamic studies might provide insight on the fundamental basis of the mass transfer coefficient correlations as well as that for interfacial areas. Tray layout parameters may have a profound effect on the dynamic performance of the column. Effects of different tray layout parameters on the mass transfer can only be modelled with a nonequilibrium model. Equilibrium model



simulations will show no difference in performance other than those affecting the hydrodynamics of the tray layout. The free area ratio influences both the mass transfer and the hydrodynamics on the tray and is one of the most important parameters. Future work on the dynamic nonequilibrium column model must include the extension to packed columns. which is rather simple if accurate and correctly behaving holdup correlations for structured and random packings are available. Recent advances in modelling packed columns have given rise to models of a more fundamental basis which also include the pressure drop over the packing. As mentioned above, more fundamental research is needed to determine better methods to estimate mass transfer coefficients in the vapor and liquid phase, as well as the total interfacial area available for mass transfer. This remains an important area for improvement. Dynamic nonequilibrium simulations of experimental data might actually help to discriminate between various models and provide the actual relations between vapor and liquid resistances. The influence of entrainment and weeping flows on the tray hydrodynamics and the mass transfer parameters is little understood, and not yet incorporated in a proper manner in the nonequilibrium model. To do so will allow the nonequilibrium model simulation of columns that operate outside the normal operation conditions. This is important for the simulation of the start-up and shut-down of column processes. More models for the condenser and reboiler should be developed, including equations that model control units normally utilized in column operations. These models can become much more advanced and include the heat transfer process taking place or even the mass transfer process of condensing or evaporation. Extension of the models will allow more specifications. Another interesting subject is the dynamic interaction of multiple interlinked columns, simulated with nonequilibrium models, for example those encountered in extractive distillation. Possibly dynamic nonequilibrium liquid-liquid column simulations are possible by extending the current steady-state nonequilibrium implementation of these operations. The design mode can be further enhanced by using an optimizing algorithm that would optimize the tray layout for minimum cost or pressure drop, maximum flexibility or mass transfer, or a combination of these.



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