A REPORT ON
Modeling and Simulation of Crude Distillation Unit

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By
Abhishek Sonu

Under the supervision of
Mrs. P. Rose Havilah
Assistant professor, COE
UPES

UNIVERSITY OF PETROLEUM AND ENERGY STUDIES
DEHRADUN, UTTARAKHAND

UNIVERSITY OF PETROLEUM & ENERGY STUDIES
(ISO 9001:2000 Certified)

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.

P. Rose Havilah 11/5/09

Ms.P.Rose Havilah
Assistant Professor

rate Office:
arbons Education & Research Society
or, PHD House,
Institutional Area
Kranti Marg, New Delhi - 110 016 India
1-11-41730151-53 Fax : +91-11-41730154

Main Campus:
Energy Acres,
PO Bidholi Via Prem Nagar,
Dehradun - 248 007 (Uttarakhand), India
Ph.: +91-135-2102690-91, 2694201/ 203/ 208
Fax: +91-135-2694204

Regional Centre (NCR) :
SCO. 9-12, Sector-14,
Gurgaon 122 007
(Haryana), India
Ph: +91-124-4540 300
Fax: +91-124-4540 330

Regional Centre (Rajahmundry):
GIET, NH 5, Velugubanda,
Rajahmundry - 533 294,
East Godavari Dist., (Andhra Pradesh), Ind
Tel: +91-883-2484811/ 855
Fax: +91-883-2484822

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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:

$$\xi = \frac{y_{ij} - y_{i,j-1}}{y^*_{ij} - y_{i,j-1}} \qquad (1.1)$$

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).

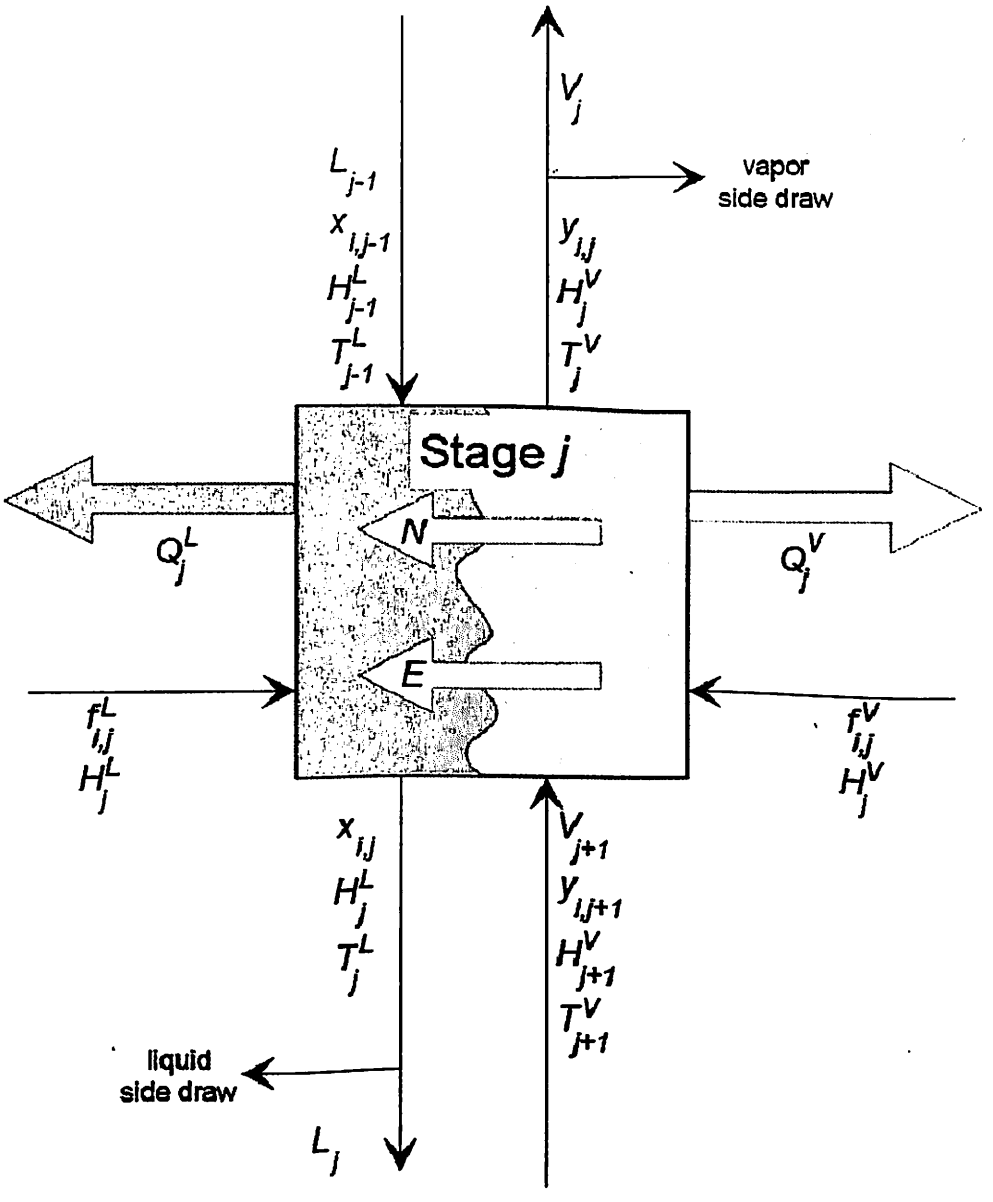


Figure 2.1: Schematic diagram of a nonequilibrium stage (Taylor and Krishna, 1993).

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:

$$\begin{aligned} M_{ij}^V &\equiv (1+r_j^V+\phi_j^V)V_j y_{ij} - V_{j+1}y_{i,j+1} - \phi_{j-1}^V V_{j-1}y_{i,j-1} - f_{ij}^V - \sum_{\nu=1}^{n} G_{ij\nu}^V + N_{ij} \\ &= 0 \qquad i=1,2,\ldots,c \end{aligned} \tag{2.1}$$

$$\begin{aligned} M_{ij}^L &\equiv (1+r_j^L+\phi_j^L)L_j x_{ij} - L_{j-1}x_{i,j-1} - \phi_{j+1}^L L_{j+1}x_{i,j+1} - f_{ij}^L - \sum_{\nu=1}^{n} G_{ij\nu}^L - N_{ij} \\ &= 0 \qquad i=1,2,\ldots,c \end{aligned} \tag{2.2}$$

where G_{ijv} is the interlinked flow rate for component i from stage v 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.

$$\begin{aligned} M_{tj}^V &\equiv (1+r_j^V+\phi_j^V)V_j - V_{j+1} - \phi_{j-1}^V V_{j-1} - F_j^V - \sum_{i=1}^{c}\sum_{\nu=1}^{n} G_{ij\nu}^V + N_{tj} \\ &= 0 \end{aligned} \tag{2.3}$$

$$\begin{aligned} M_{tj}^L &\equiv (1+r_j^L+\phi_j^L)L_j - L_{j-1} - \phi_{j+1}^L L_{j+1} - F_j^L - \sum_{i=1}^{c}\sum_{\nu=1}^{n} G_{ij\nu}^L - N_{tj} \\ &= 0 \end{aligned} \tag{2.4}$$

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:

$$R_{ij}^V \equiv N_{ij} - N_{ij}^V = 0 \qquad i=1,2\ldots,c-1 \tag{2.5}$$

$$R_{ij}^L \equiv N_{ij} - N_{ij}^L = 0 \qquad i=1,2\ldots,c-1 \tag{2.6}$$

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

$$N_{ij}^V = J_{ij}^V a_j^I + y_{ij} N_{tj} \quad (2.7)$$

$$N_{ij}^L = J_{ij}^L a_j^I + x_{ij} N_{tj} \quad (2.8)$$

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

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

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

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

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

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

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

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:

$$R_{ii}^P = \frac{z_i}{k_{ic}^P} + \sum_{k=1,k \neq i}^{c} \frac{z_k}{k_{ik}^P} \quad (2.13)$$

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

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:

$$E_j^V \equiv (1 + r_j^V + \phi_j^V)V_j H_j^V - V_{j+1}H_{j+1}^V - \phi_{j-1}^V V_{j-1}H_{j-1}^V - F_j^V H_j^{VF} - \sum_{\nu=1}^{n} G_{j\nu}^V H_{j\nu}^V + Q_j^V + e_j^V = 0 \quad (2.15)$$

$$E_j^L \equiv (1 + r_j^L + \phi_j^L)L_j H_j^L - L_{j-1}H_{j-1}^L - \phi_{j+1}^L L_{j+1}H_{j+1}^L - F_j^L H_j^{LF} - \sum_{\nu=1}^{n} G_{j\nu}^L H_{j\nu}^L + Q_j^L - e_j^L = 0 \quad (2.16)$$

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

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

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

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:

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

Where H^V and H^L are the vapor and liquid heat transfer coefficients respectively, and ' 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:

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

$$h^V = k\rho C_p \mathrm{Le}^{2/3} \quad (2.21)$$

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

$$h^L = k\rho C_p \sqrt{\mathrm{Le}} \qquad (2.22)$$

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:

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

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:

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

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

as well as at the interface:

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

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

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	c
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	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*):

$T^1 = T^2 = \ldots = T^\pi$	*Thermal Equilibrium - no heat flux across phases*
$P^1 = P^2 = \ldots = P^\pi$	*Mechanical Equilibrium - no phase displacement*
$\mu_i^1 = \mu_i^2 = \ldots = \mu_i^\pi$	*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,

$$Ki = yi/xi$$

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

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

$$f_i^V = \phi_i^V y_i P$$

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

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:

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

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:

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

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:

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

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:

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

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

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

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

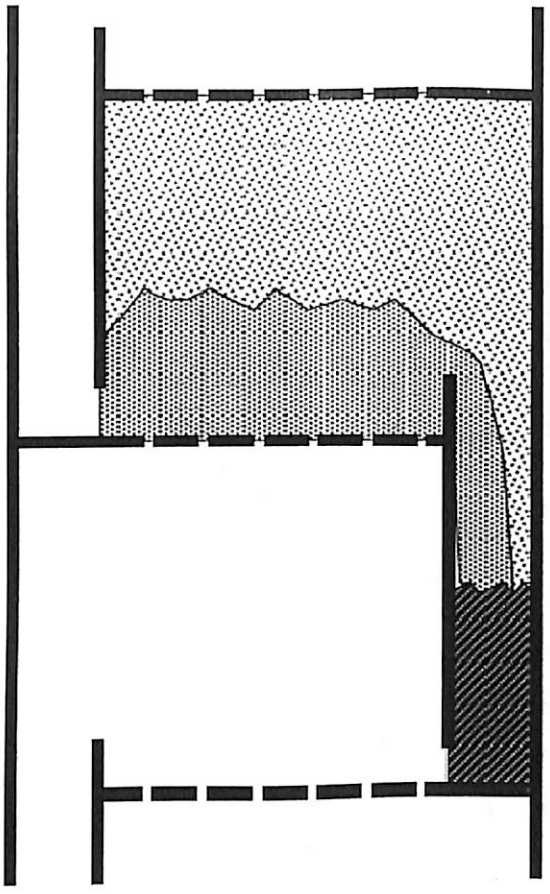


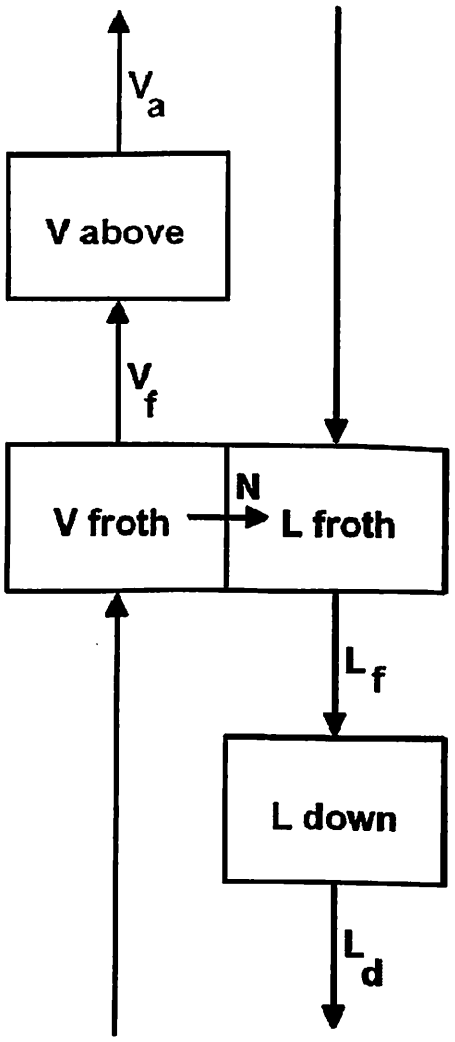
Figure 4.1: Schematic diagram of a general tray.

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

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

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

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:

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

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

$$\frac{dU_{ij}^{Va}}{dt} = y_{ij}^{f} V_j^f + F_{ij}^{Va} - y_{ij}^{a} V_j^a \tag{4.5}$$

$$\frac{dU_{ij}^{Ld}}{dt} = x_{ij}^{f} L_j^f + F_{ij}^{Ld} - x_{ij}^{d} L_j^d \tag{4.6}$$

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

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

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

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

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

The energy balances for each holdup are:

$$\frac{dE_j^{Vf}}{dt} = \frac{E_{j+1}^{Va}}{U_{t,j+1}^{Va}} V_{j+1}^{c} + \sum_{i=1}^{c} H_{ij}^{FVf} F_{ij}^{Vf} - \frac{E_j^{Vf}}{U_{tj}^{Vf}} V_j^f - \epsilon_j^{Vf} + Q_j^{Vf} \tag{4.11}$$

$$\frac{dE_j^{Lf}}{dt} = \frac{E_{j-1}^{Ld}}{U_{tj-1}^{Ld}} L_{j-1}^{d} + \sum_{i=1}^{c} H_{ij}^{FLf} F_{ij}^{Lf} - \frac{E_j^{Lf}}{U_{tj}^{Lf}} L_j^{f} + \epsilon_j^{Lf} + Q_j^{Lf} \quad (4.12)$$

$$\frac{dE_j^{Vf}}{dt} = \frac{E_j^{Vf}}{U_{tj}^{Vf}} V_j^{f} + \sum_{i=1}^{c} H_{ij}^{FVf} F_{ij}^{Vf} - \frac{E_j^{Vf}}{U_{tj}^{Vf}} V_j^{f} + Q_j^{Vf} \quad (4.13)$$

$$\frac{dE_j^{Ld}}{dt} = \frac{E_j^{Ld}}{U_{tj}^{Ld}} L_j^{d} + \sum_{i=1}^{c} H_{ij}^{FLd} F_{ij}^{Ld} - \frac{E_j^{Ld}}{U_{tj}^{Ld}} L_j^{d} + Q_j^{Ld} \quad (4.14)$$

Where □ is the energy transport to/from the interface (see below). $H_{ij}^{F\,Pw}$ 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

$$0 = \sum_{i=1}^{c} (H_{ij}^{Vf} U_{ij}^{Vf}) - E_j^{Vf} \quad (4.15)$$

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

$$0 = \sum_{i=1}^{c} (H_{ij}^{Vd} U_{ij}^{Vd}) - E_j^{Vd} \quad (4.17)$$

$$0 = \sum_{i=1}^{c} (H_{ij}^{Ld} U_{ij}^{Ld}) - E_j^{Ld} \quad (4.18)$$

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:

$$\epsilon_j^{Vf} = \sum_{i=1}^{c} N_{ij}^{Vf} H_{ij}^{Vf} + h_j^{Vf} a_j (T_j^{Vf} - T_j^{I}) \quad (4.19)$$

$$\epsilon_j^{Lf} = \sum_{i=1}^{c} N_{ij}^{Lf} H_{ij}^{Lf} - h_j^{Lf} a_j (T_j^{I} - T_j^{Lf}) \quad (4.20)$$

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

$$0 = \epsilon_j^{Vf} - \epsilon_j^{Lf} \quad (4.21)$$

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

$$0 = \sum_{i=1}^{c} y_{ij}^{If} - 1 \quad (4.22)$$

$$0 = \sum_{i=1}^{c} x_{ij}^{If} - 1 \quad (4.23)$$

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

$$0 = K_{ij}^{lf} x_{ij}^{lf} \quad y_{ij}^{lf} \qquad (4.24)$$

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:

$$(0) = (N_j) - N_{tj}(y_j^{Vf}) - c_j^{Vf} a_j [R_j^{Vf}]^{-1} [\Gamma_j^{Vf}]((y_j^{Vf}) \quad (y_j^{lf})) \qquad (4.25)$$

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

Where N_{ij} 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}):

$$0 = p_1 - p_{spec} \qquad (4.27)$$

$$0 = p_j - p_{j-1} - \Delta p_{j-1} \qquad (4.28)$$

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:

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

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

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

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

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^{Ld}_{tj} should be a function of, L^d_j 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

$$0 = V_j^f - V_j^a \qquad (4.33)$$

To use a constant molar vapor holdup above the froth (usually the change in U^{Va}_{tj} 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	+
$E_j^{Vf}, E_j^{Lf}, E_j^{Va}, E_j^{Ld}$	(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	–
$V_j^f, L_j^f, V_j^a, L_j^d$	(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^f	(4.21)	1	–
N_{ij}	(4.25), (4.26)	c	–
x_{ij}^{lf}	(4.26), (4.23), (4.24)	c	–
y_{ij}^{lf}	(4.24), (4.22)	c	–
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:

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

Where the liquid mole fraction is computed by:

$$x_{ib} = \frac{U_{ib}^L}{U_{tb}^L} \qquad (4.35)$$

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

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

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

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

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:

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

The energy holdup and energy relation are:

$$\frac{dE_b}{dt} = \frac{E_n^{L,d}}{U_n^{L,d}} L_n^d - \frac{E_b}{U_{tb}^{L}} L_b \qquad (4.39)$$

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:

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

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

Variable(s)	Equation(s)	Number	Differential
U_{ij}	(4.62)	c	+
U_{tj}	(4.63)	1	–
E_j	(4.65)	1	+
y_{ij}	(4.67)	c	–
V_j	(4.68)	1	–
L_j	(4.64)	1	–
T_j	(4.66)	1	–
P_j	(4.69)	1	–

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

$$0 = h_j^{cf} A_j^f c_j^{Lf} - U_{tj} \tag{4.64}$$

$$\frac{dE_j}{dt} = (\sum_{i=1}^{c} y_{i,j+1} H_{i,j+1}^V) V_{j+1} + (\frac{E_{j-1}}{U_{j-1}}) L_{j-1} + \sum_{i=1}^{c} H_{ij}^F F_{ij} + Q_j$$
$$-(\sum_{i=1}^{c} y_{ij} H_{ij}^V) V_j - (\frac{E_j}{U_j}) L_j \tag{4.65}$$

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

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

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

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

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):

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

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

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

$$0 = (h_j^f - h_j^{cl})A_j^f c_j^{Vf} - U_{ij}^V \quad (4.73)$$

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

$$\frac{dE_j^V}{dt} + \frac{dE_j^L}{dt} = (\frac{E_{j+1}^V}{U_{i,j+1}^V})V_{j+1} + (\frac{E_{j-1}^L}{U_{i,j-1}^L})L_j + \sum_{i=1}^{c} H_{ij}^F F_{ij} + Q_j$$

$$-(\frac{E_j^V}{U_j^V})V_j - (\frac{E_j^L}{U_j^L})L_j \quad (4.75)$$

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

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

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

$$0 = p_j - p_{j-1} - \Delta P_{j-1} \quad (4.79)$$

(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

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

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

HYPROTECH HYDROCARBON INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name:	C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set:	Field-USGPM
Date/Time:	Sun May 03 02:57:56 2009

Assay: Assay-1

INPUT DATA

Data Type: TBP

Assay Basis:	Liquid Volume	TBP Distillation Conditions:	Atmospheric

Light Ends Status : Input Composition

Percentage of Light Ends in Assay:	1 *	Light Ends Basis:	Liquid Volume %

Light Ends Composition Table

Light Ends	Composition
Methane	6.600e-003 *
Ethane	2.250e-002 *
Propane	0.3200 *
I-Butane	0.2400 *
n-Butane	0.8200 *
H2O	0.0000 *

Bulk Properties: Used

Molecular Weight		—	Viscosity1 Temperature	(F)	100.0 *
Mass Density	(lb/ft3)	54.82 *	Viscosity1	(cP)	—
Watson Uopk		—	Viscosity2 Temperature	(F)	210.0 *
			Viscosity2	(cP)	—

Distillation Table

Assay Percent	Temperature (F)
0 *	15.00 *
5 *	90.00 *
9 *	165.0 *
15 *	240.0 *
20 *	310.0 *
30 *	435.0 *
40 *	524.0 *
50 *	620.0 *
60 *	740.0 *
70 *	885.0 *
76 *	969.0 *
80 *	1015 *
85 *	1050 *

Input Data Status

Molecular Weight : Not Used	Density : Not Used	Viscosity : Not Used

WORKING CURVES

Point #	Moles	Cum. Moles	NBP	MWT	Mass Density (lb/ft3)	Viscosity 1 (cP)	Viscosity 2 (cP)
0	0.0000	0.0000	31.10 F	56.87	36.50	0.1425	8.290e-002
1	1.000e-002	1.000e-002	48.77 F	59.69	43.10	0.2614	0.1361
2	1.000e-002	2.000e-002	64.86 F	62.72	43.56	0.2788	0.1469
3	1.000e-002	3.000e-002	81.66 F	65.90	44.02	0.2977	0.1586
4	1.000e-002	4.000e-002	98.48 F	69.53	44.47	0.3081	0.1670
5	1.000e-002	5.000e-002	115.2 F	73.42	44.91	0.3135	0.1735
6	1.000e-002	6.000e-002	131.8 F	77.36	45.34	0.3023	0.1732
7	1.000e-002	7.000e-002	148.1 F	81.30	45.75	0.3077	0.1868

* Specified by user.

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Table 5. 2: working curves

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name:	C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set:	Field-USGPM
Date/Time:	Sun May 03 02:57:56 2009

Assay: Assay-1 (continued)

WORKING CURVES

Point #	Moles	Cum. Moles	NBP	MWT	Mass Density (lb/ft3)	Viscosity 1 (cP)	Viscosity 2 (cP)
8	1.000e-002	8.000e-002	164.0 F	85.11	46.14	0.3268	0.1971
9	1.000e-002	9.000e-002	177.5 F	87.93	46.48	0.3456	0.2070
10	1.000e-002	1.000e-001	187.4 F	90.30	46.72	0.3605	0.2148
11	2.500e-002	0.1250	214.9 F	98.72	47.37	0.4081	0.2390
12	2.500e-002	0.1500	247.6 F	108.1	48.12	0.4769	0.2727
13	2.500e-002	0.1750	278.9 F	117.3	48.82	0.5562	0.3103
14	2.500e-002	0.2000	310.2 F	127.1	49.50	0.6495	0.3534
15	2.500e-002	0.2250	341.9 F	137.7	50.17	0.7601	0.4031
16	2.500e-002	0.2500	373.5 F	148.8	50.82	0.9258	0.4680
17	2.500e-002	0.2750	404.3 F	160.3	51.44	1.135	0.5392
18	2.500e-002	0.3000	432.9 F	171.4	52.00	1.381	0.6172
19	2.500e-002	0.3250	457.3 F	181.4	52.47	1.646	0.6953
20	2.500e-002	0.3500	479.2 F	190.9	52.89	1.936	0.7753
21	2.500e-002	0.3750	500.4 F	200.5	53.28	2.277	0.8635
22	2.500e-002	0.4000	521.4 F	210.4	53.67	2.692	0.9634
23	2.500e-002	0.4250	543.4 F	221.2	54.06	3.228	1.083
24	2.500e-002	0.4500	566.3 F	232.9	54.47	3.934	1.228
25	2.500e-002	0.4750	590.1 F	245.3	54.89	4.873	1.404
26	2.500e-002	0.5000	615.1 F	258.7	55.32	6.175	1.624
27	2.500e-002	0.5250	641.9 F	273.4	55.78	8.072	1.908
28	2.500e-002	0.5500	670.4 F	288.9	56.26	10.93	2.280
29	2.500e-002	0.5750	700.3 F	304.6	56.75	15.37	2.773
30	2.500e-002	0.6000	731.8 F	321.7	57.26	22.63	3.444
31	2.500e-002	0.6250	765.7 F	343.5	57.80	35.57	4.409
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
34	2.500e-002	0.7000	873.6 F	417.2	59.45	204.4	10.85
35	2.500e-002	0.7250	909.6 F	440.9	59.98	416.2	15.36
36	2.500e-002	0.7500	944.4 F	463.7	60.48	891.2	22.09
37	2.500e-002	0.7750	978.7 F	485.1	60.94	1945	31.76
38	2.500e-002	0.8000	1005 F	504.4	61.34	4145	44.83
39	2.500e-002	0.8250	1027 F	519.3	61.65	7739	59.29
40	2.500e-002	0.8500	1044 F	531.5	61.88	1.283e+004	74.13
41	1.500e-002	0.8650	1059 F	542.1	62.08	2.034e+004	90.66
42	1.500e-002	0.8800	1072 F	552.2	62.26	3.027e+004	110.0
43	1.500e-002	0.8950	1086 F	562.5	62.44	3.876e+004	134.1
44	1.500e-002	0.9100	1099 F	573.0	62.62	5.004e+004	164.5
45	1.500e-002	0.9250	1113 F	583.8	62.80	6.515e+004	203.1
46	1.500e-002	0.9400	1126 F	594.9	62.98	8.596e+004	252.4
47	1.500e-002	0.9550	1140 F	606.5	63.16	1.136e+005	316.3
48	1.500e-002	0.9700	1154 F	618.6	63.34	1.536e+005	401.9
49	1.500e-002	0.9850	1161 F	624.2	63.44	1.798e+005	455.4
50	1.500e-002	1.000	1174 F	634.7	63.61	2.443e+005	580.5

* Specified by user.

Table 5. 3: component properties

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 04:11:12 2009

Blend: Blend-1

DATA

Oil Flow Information

Oil	Flow Units	Oil Flow Rate
Assay-1	Liquid Vol	—

Cut Ranges: User Points

Number of Cuts: 30

Bulk Data

Molecular Weight:	—	Mass Density:	—	Watson Uopk:	—		
Viscosity 1 Temp:	100.00 F *	Viscosity 1:	—	Viscosity 2 Temp:	210.00 F *	Viscosity 2:	—

CORRELATIONS

Default Set

Low End T (F)	High End T (F)	Tc	Pc	W	Cp
-459.67	2192.00	Lee_Kesler	Lee_Kesler	Lee_Kesler	Lee_Kesler

TABLES

Component Properties: Blend-1

Comp Name	NBP (F)	Mol Wt.	Density (lb/ft3)	Viscosity 1 (cP)	Viscosity 2 (cP)
NBP_52	51.88	58.88	41.08	0.25831	0.12462
NBP_79	79.08	63.88	43.72	0.28562	0.15104
NBP_111	111.32	70.84	44.59	0.31007	0.16897
NBP_143	143.46	78.29	45.43	0.30344	0.17622
NBP_178	177.64	86.76	46.33	0.33670	0.20234
NBP_208	208.15	92.99	46.93	0.37490	0.22223
NBP_240	239.84	101.15	47.56	0.42365	0.24673
NBP_272	271.61	110.84	48.32	0.49673	0.28226
NBP_304	303.55	120.24	49.02	0.58161	0.32219
NBP_336	335.68	130.41	49.71	0.68004	0.36749
NBP_368	367.82	141.35	50.39	0.80707	0.42237
NBP_400	400.06	153.12	51.05	0.99571	0.49271
NBP_433	432.70	166.11	51.73	1.2527	0.57726
NBP_464	464.48	179.86	52.40	1.5959	0.68092
NBP_496	496.09	193.97	53.01	2.0310	0.80043
NBP_528	527.77	208.60	53.59	2.5891	0.94159
NBP_560	559.95	224.08	54.16	3.3704	1.1135
NBP_592	591.91	240.22	54.72	4.4384	1.3249
NBP_624	623.92	256.79	55.26	5.9288	1.5840
NBP_656	656.08	273.78	55.79	8.0822	1.9090
NBP_688	688.17	290.84	56.31	11.294	2.3236
NBP_720	720.16	307.21	56.82	16.161	2.8532
NBP_752	752.29	324.34	57.32	23.725	3.5356
NBP_784	784.46	344.68	57.82	36.110	4.4445
NBP_832	832.24	377.89	58.55	73.226	6.4545
NBP_896	895.94	420.58	59.51	221.62	11.297
NBP_960	960.31	483.86	60.48	877.81	21.929
NBP_1027	1027.49	513.12	61.51	5790.1	52.097

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* Specified by user.

Table 5. 4: component properties

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 04:11:12 2009

Blend: Blend-1 (continued)

TABLES

Component Properties: Blend-1

Comp Name	NBP (F)	Mol Wt.	Density (lb/ft3)	Viscosity 1 (cP)	Viscosity 2 (cP)
NBP_1085	1085.44	557.90	62.36	34418	120.98
NBP_1153	1152.70	612.85	63.25	1.3030e+005	352.69

Comp Name	Critical Temp (F)	Critical Press (psia)	Accentric Fac.	Watson K
NBP_52	357.75	566.97	0.13703	12.130
NBP_79	391.15	554.07	0.16622	11.595
NBP_111	426.48	534.80	0.21071	11.591
NBP_143	463.23	523.29	0.24964	11.587
NBP_178	503.45	507.26	0.28193	11.573
NBP_208	537.38	480.05	0.30882	11.603
NBP_240	571.40	450.45	0.33824	11.629
NBP_272	605.78	426.25	0.36763	11.618
NBP_304	639.44	402.95	0.39869	11.614
NBP_336	672.61	381.00	0.43111	11.612
NBP_368	705.25	360.68	0.46445	11.609
NBP_400	737.56	341.73	0.49874	11.604
NBP_433	769.92	324.14	0.53402	11.595
NBP_464	801.19	308.38	0.56881	11.582
NBP_496	831.60	293.11	0.60470	11.577
NBP_528	861.54	278.42	0.64174	11.576
NBP_560	891.49	264.22	0.68028	11.578
NBP_592	920.96	250.98	0.71921	11.579
NBP_624	950.10	238.39	0.75899	11.581
NBP_656	979.07	226.43	0.79964	11.583
NBP_688	1007.8	215.21	0.84070	11.584
NBP_720	1036.1	204.82	0.88225	11.586
NBP_752	1064.3	194.49	0.92469	11.588
NBP_784	1092.4	184.92	0.96356	11.589
NBP_832	1133.9	171.78	1.0218	11.588
NBP_896	1188.6	155.80	1.0973	11.586
NBP_960	1243.6	141.36	1.1706	11.579
NBP_1027	1301.1	128.24	1.2429	11.561
NBP_1085	1349.9	117.84	1.3042	11.551
NBP_1153	1405.4	106.98	1.3748	11.550

Component Breakdown: Blend-1

Comp Name	Liquid Vol %	LiqVol % Cum	Vol Flow (USGPM)	Mass Flow (lb/hr)	Molar Flow (lbmole/hr)
Methane	0.0065	0.0065	---	---	---
Ethane	0.0225	0.0290	---	---	---
Propane	0.3200	0.3490	---	---	---
I-Butane	0.2400	0.5890	---	---	---
n-Butane	0.8200	1.4090	---	---	---
H2O	0.0000	1.4090	---	---	---
NBP_52	1.3669	2.7759	---	---	---
NBP_79	1.8927	4.6685	---	---	---
NBP_111	1.8923	6.5608	---	---	---
NBP_143	1.9428	8.5034	---	---	---

* Specified by user.

Table 5. 5: component break down

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name:	C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set:	Field-USGPM
Date/Time:	Sun May 03 04:11:12 2009

Blend: Blend-1 (continued)

TABLES

Component Breakdown: Blend-1

Comp Name	Liquid Vol %	LiqVol % Cum	Vol Flow (USGPM)	Mass Flow (lb/hr)	Molar Flow (lbmole/hr)
NBP_178	2.5971	11.1006	—	—	—
NBP_208	2.0168	13.1173	—	—	—
NBP_240	2.4670	15.5844	—	—	—
NBP_272	2.5318	18.1162	—	—	—
NBP_304	2.5191	20.6353	—	—	—
NBP_336	2.4866	23.1219	—	—	—
NBP_368	2.5083	25.6302	—	—	—
NBP_400	2.6168	28.2470	—	—	—
NBP_433	2.9853	31.2324	—	—	—
NBP_464	3.5364	34.7687	—	—	—
NBP_496	3.7526	38.5213	—	—	—
NBP_528	3.6375	42.1588	—	—	—
NBP_560	3.4150	45.5738	—	—	—
NBP_592	3.2346	48.8084	—	—	—
NBP_624	2.9852	51.7936	—	—	—
NBP_656	2.7726	54.5662	—	—	—
NBP_688	2.6323	57.1985	—	—	—
NBP_720	2.4833	59.6819	—	—	—
NBP_752	2.3150	61.9969	—	—	—
NBP_784	2.2138	64.2107	—	—	—
NBP_832	4.3232	68.5339	—	—	—
NBP_896	4.3608	72.8947	—	—	—
NBP_960	4.8190	77.7138	—	—	—
NBP_1027	7.6512	85.3649	—	—	—
NBP_1085	7.0358	92.4007	—	—	—
NBP_1153	7.5993	100.0000	—	—	—
		Total :	—	—	—

Molar Compositions

	Blend-1
Methane	0.0003
Ethane	0.0006
Propane	0.0088
i-Butane	0.0054
n-Butane	0.0192
H2O	0.0000
NBP_52	0.0357
NBP_79	0.0485
NBP_111	0.0447
NBP_143	0.0422
NBP_178	0.0519
NBP_208	0.0381
NBP_240	0.0434
NBP_272	0.0414
NBP_304	0.0384
NBP_336	0.0354
NBP_368	0.0334
NBP_400	0.0326

Table 5. 6: molar composition

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 04:11:12 2009

Blend: Blend-1 (continued)

TABLES

Molar Compositions

	Blend-1
NBP_433	0.0348
NBP_464	0.0385
NBP_496	0.0384
NBP_528	0.0350
NBP_560	0.0309
NBP_592	0.0276
NBP_624	0.0240
NBP_656	0.0211
NBP_688	0.0191
NBP_720	0.0172
NBP_752	0.0153
NBP_784	0.0139
NBP_832	0.0251
NBP_896	0.0231
NBP_960	0.0235
NBP_1027	0.0343
NBP_1085	0.0294
NBP_1153	0.0293

Oil Properties: Blend-1

Basis: Liquid Volume

Percent	Critical Temp (F)	Critical Pressure (psia)	Accentric Factor	Molecular Weight
0.00	-166.91	665.65	-9.7924e-003	12.624
1.00	305.74	550.67	0.20094	58.125
2.00	354.10	565.55	0.14244	58.737
3.50	388.11	556.05	0.16175	63.045
5.00	415.14	540.77	0.19603	68.344
7.50	462.63	523.47	0.24907	78.167
10.00	506.52	505.23	0.28434	87.324
12.50	543.35	474.75	0.31388	94.311
15.00	580.59	443.84	0.34603	103.59
17.50	614.51	420.14	0.37548	113.08
20.00	647.74	397.35	0.40665	122.71
25.00	713.28	355.83	0.47287	144.18
30.00	772.63	322.73	0.53699	167.26
35.00	818.22	299.81	0.58866	187.68
40.00	858.78	279.77	0.63823	207.21
45.00	901.38	259.72	0.69321	229.41
50.00	947.19	239.62	0.75497	255.11
55.00	998.26	218.87	0.82698	285.21
60.00	1054.3	198.05	0.90976	317.96
65.00	1116.4	177.19	0.99723	363.67
70.00	1179.8	158.28	1.0851	413.67
75.00	1240.2	142.20	1.1662	461.12
80.00	1288.5	131.00	1.2274	501.95
85.00	1325.1	123.06	1.2729	534.58
90.00	1357.2	116.04	1.3134	564.98
92.50	1374.7	112.26	1.3356	582.17
95.00	1394.8	108.12	1.3610	602.05

* Specified by user.

Table 5. 7: oil properties

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 04:11:12 2009

Blend: Blend-1 (continued)

TABLES

Oil Properties: Blend-1

Basis liquid Volume

Percent	Critical Temp (F)	Critical Pressure (psia)	Accentric Factor	Molecular Weight
96.50	1407.7	105.55	1.3777	615.10
98.00	1419.0	103.46	1.3922	626.37
99.00	1426.6	102.11	1.4018	633.87
100.00	1434.2	100.79	1.4115	641.38

Percent	Density (lb/ft3)	Viscosity 1 (cP)	Viscosity 2 (cP)
0.00	17.809	6.1706e-003	7.7137e-002
1.00	36.340	0.13571	7.5967e-002
2.00	40.589	0.23781	0.11965
3.50	43.469	0.28201	0.15017
5.00	44.343	0.30346	0.16459
7.50	45.325	0.30355	0.17608
10.00	46.295	0.33971	0.20418
12.50	46.944	0.38248	0.22609
15.00	47.656	0.44096	0.25526
17.50	48.404	0.51739	0.29209
20.00	49.098	0.60437	0.33282
25.00	50.450	0.84778	0.43838
30.00	51.686	1.2784	0.58535
35.00	52.638	1.8249	0.74528
40.00	53.432	2.5389	0.92730
45.00	54.240	3.6887	1.1794
50.00	55.093	5.7545	1.5555
55.00	56.025	10.076	2.1742
60.00	57.030	20.612	3.2705
65.00	58.127	53.665	5.4873
70.00	59.240	182.31	10.254
75.00	60.294	798.87	20.973
80.00	61.163	3690.7	42.431
85.00	61.810	14207	77.441
90.00	62.350	43372	138.81
92.50	62.634	70128	194.08
95.00	62.953	1.0785e+005	285.93
96.50	63.160	1.3760e+005	368.49
98.00	63.344	1.8078e+005	458.85
99.00	63.466	2.1685e+005	531.10
100.00	63.588	2.6012e+005	614.72

Oil BP Temperatures: Blend-1

Basis liquid Volume

Percent	TBP (F)	ASTM D86 (F)	D86 Crack Reduced (F)	D1160 Vac. (F)	D1160 ATM (F)	D2887 (F)
0.00	-3.1676	86.282	86.282	-125.20	8.9307	-23.883
1.00	31.122	108.74	108.74	-101.73	40.314	-10.048
2.00	50.475	121.95	121.95	-88.247	58.266	3.4240
3.50	75.461	139.58	139.58	-70.571	81.703	23.096
5.00	100.84	158.09	158.09	-52.335	105.77	42.325
7.50	142.94	190.00	190.00	-21.501	146.22	74.077
10.00	180.34	219.52	219.52	6.4667	182.64	107.26

* Specified by user.

Table 5. 8: oil BP temperature

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 04:11:12 2009

Blend: Blend-1 (continued)

TABLES

Oil BP Temperatures: Blend-1

Basisliquid Volume

Percent	TBP (F)	ASTM D86 (F)	D86 Crack Reduced (F)	D1160 Vac. (F)	D1160 ATM (F)	D2887 (F)
12.50	213.68	246.67	246.67	31.822	215.43	140.44
15.00	248.32	275.60	275.60	58.541	249.77	172.03
17.50	279.81	302.48	302.48	83.151	281.20	202.50
20.00	311.54	330.03	330.03	108.20	312.99	232.66
25.00	375.79	387.05	387.05	159.61	377.62	293.66
30.00	435.44	440.72	440.72	207.93	437.63	358.76
35.00	482.10	482.52	482.52	245.99	484.39	422.90
40.00	524.81	520.93	513.84	280.87	526.88	477.45
45.00	570.63	562.45	551.72	318.03	571.75	530.47
50.00	620.71	608.16	591.58	358.98	620.73	587.93
55.00	677.49	660.47	634.09	406.77	677.29	646.17
60.00	740.81	719.14	676.98	461.08	740.79	701.87
65.00	812.06	785.21	718.70	522.88	812.06	759.95
70.00	885.56	853.19	754.68	587.57	885.56	824.52
75.00	956.30	918.19	783.39	650.77	956.31	889.49
80.00	1012.8	969.52	802.83	701.88	1012.7	951.81
85.00	1055.8	1008.2	815.91	741.19	1055.7	1008.8
90.00	1095.4	1043.4	826.60	777.76	1095.4	1066.6
92.50	1118.2	1063.5	832.65	798.97	1118.2	1102.4
95.00	1141.5	1083.8	838.32	820.70	1141.5	1143.8
96.50	1155.5	1095.9	841.60	833.62	1155.5	1170.9
98.00	1169.4	1107.9	844.77	846.88	1169.4	1199.5
99.00	1178.6	1115.8	846.81	855.52	1178.6	1219.3
100.00	1187.6	1123.5	848.79	864.08	1187.6	1239.5

Oil User Properties: Blend-1

Basisliquid Volume

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				
55.00				
60.00				
65.00				
70.00				

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* Specified by user.

Figure 5. 1: TBP curve

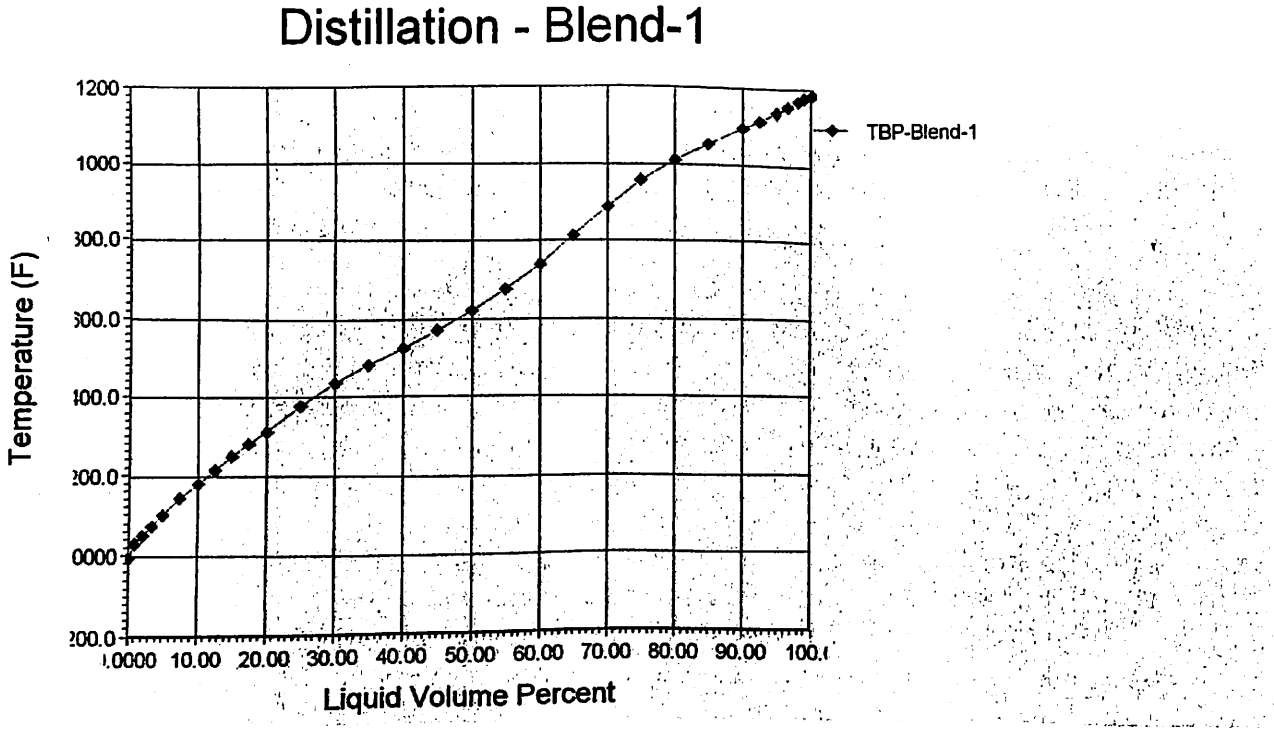


Table 5. 9: Oil distribustion table

Oil distribustion table-HYSYS			
Name	Begin T [F]	End T [F]	Fraction
Off Gas	42.30155119	49.99998901	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

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.

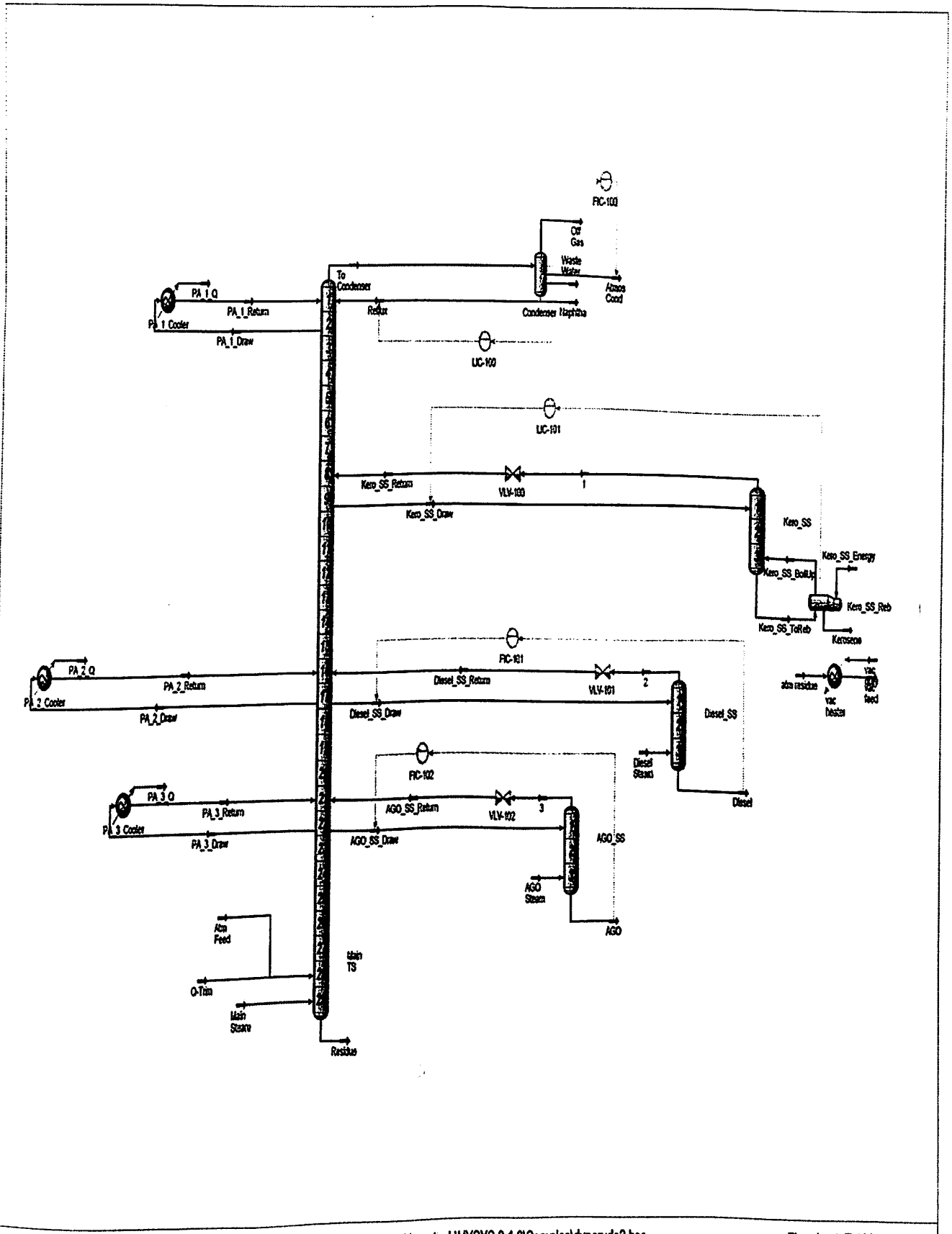
The column of atmospheric distillation is the unit where the most part of the crude fractioning is produced. The model is composed of a column with 29 theoretical trays, with a partial condenser and without a reboiler. It has 3 side columns of 3 theoretical trays each, which are of kerosene with a reboiler, 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



Sun May 03 14:54:16 2009 Case: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc Flowsheet: T-100 (COL1)

Table 5. 10: connection of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION
Main Steam	29__Main TS	
Q-Trim	28__Main TS	
Atm Feed	28__Main TS	Valve VLV-100
Kero_SS_Energy	Kero_SS_Reb	
Diesel Steam	3__Diesel_SS	
AGO Steam	3__AGO_SS	
atm residue		
vac duty		

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION
Residue	29__Main TS	
Atmos Cond	Condenser	
Off Gas	Condenser	
Waste Water	Condenser	
Naphtha	Condenser	
Kerosene	Kero_SS_Reb	
Diesel	3__Diesel_SS	
AGO	3__AGO_SS	
PA_1_Q	PA_1	
PA_2_Q	PA_2	
PA_3_Q	PA_3	
vac feed		

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error	Wt. Tol.	Abs. Tol.	Active	Estimate	Used
Kero_SS Prod Flow	271.2 USGPM *	271.2 USGPM	---	1.000e-002 *	4.403 USGPM *	On	On	On
Diesel_SS Prod Flow	561.5 USGPM *	562.2 USGPM	---	1.000e-002 *	4.403 USGPM *	On	On	On
AGO_SS Prod Flow	131.2 USGPM *	131.3 USGPM	---	1.000e-002 *	4.403 USGPM *	On	On	On
PA_1_Rate(Pa)	1458 USGPM *	1458 USGPM	---	1.000e-002 *	4.403 USGPM *	On	On	On
PA_1_Duty(Pa)	-5.500e+007 Btu/hr *	---	---	1.000e-002 *	0.9478 Btu/hr *	On	On	On
PA_2_Rate(Pa)	875.0 USGPM *	874.8 USGPM	---	1.000e-002 *	4.403 USGPM *	On	On	On
PA_2_Duty(Pa)	-3.500e+007 Btu/hr *	---	---	1.000e-002 *	0.9478 Btu/hr *	On	On	On
PA_3_Rate(Pa)	875.0 USGPM *	874.8 USGPM	---	1.000e-002 *	4.403 USGPM *	On	On	On
PA_3_Duty(Pa)	-3.500e+007 Btu/hr *	---	---	1.000e-002 *	0.9478 Btu/hr *	On	On	On
Naphtha Prod Rate	670.8 USGPM *	671.0 USGPM	---	1.000e-002 *	4.403 USGPM *	On	On	On
Liquid Flow	102.1 USGPM *	92.19 USGPM	---	1.000e-002 *	4.403 USGPM *	On	On	On
Kero Reb Duty	7.500e+006 Btu/hr *	7.500e+006 Btu/hr	---	1.000e-002 *	9.478 Btu/hr *	On	On	On
Vap Prod Flow	0.0000 lbmole/hr *	0.0000 lbmole/hr	---	1.000e-002 *	2.205 lbmole/hr *	On	On	On
Reflux Ratio	1.000 *	76.37	---	1.000e-002 *	1.000e-002 *	Off	On	Off

SPECS

Column Specification Parameters

Kero_SS Prod Flow

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stream:	Kerosene	Flow Basis:	Volume				

Diesel_SS Prod Flow

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---

* Specified by user.

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Table 5. 11: Specs of ADU

HYPROTECH LIFECYCLE INNOVATION	TEAM LND Calgary, Alberta CANADA			Case Name:	C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc		
				Unit Set:	Field-USGPM		
				Date/Time:	Sun May 03 14:58:57 2009		

Column Sub-Flowsheet: T-100 @Main (continued)

Column Specification Parameters

Diesel_SS Prod Flow							
Stream:	Diesel	Flow Basis:	Volume				
AGO_SS Prod Flow							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stream:	AGO	Flow Basis:	Volume				
PA_1_Rate(Pa)							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Flow Rate	Pumparound:	PA_1	Flow Basis:	Volume		
PA_1_Duty(Pa)							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Duty	Pumparound:	PA_1				
PA_2_Rate(Pa)							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Flow Rate	Pumparound:	PA_2	Flow Basis:	Volume		
PA_2_Duty(Pa)							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Duty	Pumparound:	PA_2				
PA_3_Rate(Pa)							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Flow Rate	Pumparound:	PA_3	Flow Basis:	Volume		
PA_3_Duty(Pa)							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Duty	Pumparound:	PA_3				
Naphtha Prod Rate							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stream:	Naphtha	Flow Basis:	Volume				
Liquid Flow							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stage:	27 Main TS	Flow Basis:	Volume	Liquid Specification:	Light		
Kero Reb Duty							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Energy Stream:	Kero_SS_Energy						
Vap Prod Flow							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stage:	Condenser	Flow Basis:	Molar				
Reflux Ratio							
Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stage:	Condenser	Flow Basis:	Molar	Liquid Specification:	Light		

* Specified by user.

Table 5. 12: profile of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

SUBCOOLING

	Condenser	
Degrees of Subcooling	0.0000 *	
Subcool to	—	

User Variables

PROFILES

General Parameters

Sub-Flow Sheet: T-100 (COL1) | Number of Stages: 29 *

Profile Estimates

	Temperature (F)	Net Liquid (USGPM)	Net Vapour (USGPM)
Condenser	100.0 *	486.7	2.138e-020
1_Main TS	250.0 *	2698	1183
2_Main TS	315.3	1475	1938
3_Main TS	341.3	1527	2171
4_Main TS	354.5	1538	2223
5_Main TS	363.7	1529	2232
6_Main TS	371.7	1510	2225
7_Main TS	380.3	1478	2206
8_Main TS	390.9	1429	2172
9_Main TS	404.0	1032	2054
10_Main TS	420.9	1002	1999
11_Main TS	434.6	984.3	1969
12_Main TS	444.6	967.9	1952
13_Main TS	452.4	945.9	1935
14_Main TS	459.6	910.0	1913
15_Main TS	467.9	839.9	1877
16_Main TS	481.1	2080	1807
17_Main TS	508.9	477.6	2079
18_Main TS	549.8	394.4	2000
19_Main TS	569.3	347.9	1917
20_Main TS	578.5	309.4	1871
21_Main TS	584.2	1486	1832
22_Main TS	611.4	451.9	2092
23_Main TS	633.1	395.0	2101
24_Main TS	643.3	343.8	2044
25_Main TS	649.8	297.0	1993
26_Main TS	655.0	234.2	1946
27_Main TS	661.5	102.1	1883
28_Main TS	674.8	1360	1751
29_Main TS	600.0 *	1283	92.28
1_Kero_SS	427.0	379.2	70.97
2_Kero_SS	437.1	395.3	108.0
3_Kero_SS	445.0	405.3	124.0
Kero_SS_Reb	455.5	271.3	134.0
1_Diesel_SS	501.6	620.5	94.19
2_Diesel_SS	495.6	600.0	65.05
3_Diesel_SS	483.3	581.5	44.58
1_AGO_SS	588.8	154.1	41.99
2_AGO_SS	588.3	144.7	27.81
3_AGO_SS	587.4	131.3	18.45

* Specified by user.

Table 5. 13: operation summary of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Stage Efficiencies

Stages	Overall Efficiency	NBP[0]1124*				
23 Main TS	1.000	1.000				
24 Main TS	1.000	1.000				
25 Main TS	1.000	1.000				
26 Main TS	1.000	1.000				
27 Main TS	1.000	1.000				
28 Main TS	1.000	1.000				
29 Main TS	1.000	1.000				
1 Kero_SS	1.000	1.000				
2 Kero_SS	1.000	1.000				
3 Kero_SS	1.000	1.000				
Kero_SS_Reb	1.000	1.000				
1 Diesel_SS	1.000	1.000				
2 Diesel_SS	1.000	1.000				
3 Diesel_SS	1.000	1.000				
1 AGO_SS	1.000	1.000				
2 AGO_SS	1.000	1.000				
3 AGO_SS	1.000	1.000				

SOLVER

Column Solving Algorithm: HYSIM Inside-Out

Solving Options		Acceleration Parameters	
Maximum Iterations:	10000	Accelerate K Value & H Model Parameters:	Off
Equilibrium Error Tolerance:	1.000e-05		
Heat/Spec Error Tolerance:	5.000e-004		
Save Solutions as Initial Estimate:	On		
Super Critical Handling Model:	Simple K		
Trace Level:	Low		
Init from Ideal K's:	Off	Damping Parameters	
Initial Estimate Generator Parameters		Azeotrope Check:	Off
Iterative IEG (Good for Chemicals):	Off	Fixed Damping Factor:	1

SIDE STRIPPERS

Side Stripper Summary

	# Stages	Liquid Draw Stage	Vapour Return Stage	Product Flow (lbmole/hr)	Reboiler Duty (Btu/hr)
Kero_SS	3 *	9 Main TS		718.1	7.500e+006
Diesel_SS	3 *	17 Main TS		1135	
AGO_SS	3 *	22 Main TS		202.4	

SIDE RECTIFIERS

PUMP AROUNDS

Pump Around Summary

	Draw Stage	Return Stage	Product Flow (lbmole/hr)	Condenser Duty (Btu/hr)
PA_1	2 Main TS	1 Main TS	—	-6.500e+007
PA_2	17 Main TS	16 Main TS	—	-3.500e+007
PA_3	22 Main TS	21 Main TS	—	-3.500e+007
PA_4	1 Kero_SS	8 Main TS	—	0.0000
PA_5	1 AGO_SS	21 Main TS	—	0.0000

* Specified by user.

Table 5. 14: rating of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Pump Around Summary

	Draw Stage	Return Stage	Product Flow (lbmole/hr)	Condenser Duty (Btu/hr)
PA_6	1 Diesel_SS	16 Main TS	—	0.0000

VAP BYPASSES

RATING

Tray Sections

Tray Section		Main TS	Kero_SS	Diesel_SS	AGO_SS
Tray Diameter	(ft)	44.95 *	3.937 *	9.843 *	4.921 *
Weir Height	(ft)	0.1640 *	0.1640 *	0.1640 *	0.1640 *
Weir Length	(ft)	32.81 *	5.577 *	8.858 *	3.937 *
Tray Space	(ft)	1.969 *	1.640 *	1.640 *	1.640 *
Tray Volume	(ft3)	3123	19.97	124.8	31.20
Disable Heat Loss Calculations		No	No	No	No
Heat Model		None	None	None	None
Rating Calculations		No	No	No	No
Tray Hold Up	(ft3)	260.3	1.997	12.48	3.120

Vessels

Vessel		Condenser	Kero_SS_Reb		
Diameter	(ft)	13.12 *	5.235		
Length	(ft)	13.12 *	3.281		
Volume	(ft3)	1759	70.63 *		
Orientation		Horizontal	Vertical		
Vessel has a Boot		Yes	No		
Boot Diameter	(ft)	2.187 *	—		
Boot Length	(ft)	4.374 *	—		
Hold Up	(ft3)	528.5	35.35		

Other Equipment In Column Flowsheet

PA_1_Cooler	PA_2_Cooler	PA_3_Cooler	VLV-100
VLV-101	VLV-102	vac heater	

Pressure Profile

	Pressure (psia)	Pressure Drop (psi)
Condenser	28.70 psia	—
1 Main TS	37.70 psia *	—
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	—	—
15 Main TS	—	—

* Specified by user.

Table 5. 15: properties atm feed of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

16_Main TS	—	—
17_Main TS	—	—
18_Main TS	—	—
19_Main TS	—	—
20_Main TS	—	—
21_Main TS	—	—
22_Main TS	—	—
23_Main TS	—	—
24_Main TS	—	—
25_Main TS	—	—
26_Main TS	—	—
27_Main TS	—	—
28_Main TS	—	—
29_Main TS	30.97 psia *	—
1_Kero_SS	—	—
2_Kero_SS	—	—
3_Kero_SS	29.58 psia *	—
Kero_SS_Reb	29.58 psia	—
1_Diesel_SS	—	—
2_Diesel_SS	—	—
3_Diesel_SS	30.15 psia *	—
1_AGO_SS	—	—
2_AGO_SS	—	—
3_AGO_SS	30.56 psia *	—

Pressure Solving Options

Pressure Tolerance	1.000e-004 *	Pressure Drop Tolerance	1.000e-004 *	Damping Factor	1.000 *	Max Press Iterations	100 *

PROPERTIES

Properties : Atm Feed

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction		0.6972	0.6972	0.3028	0.0000
Temperature:	(F)	612.6	612.6	612.6	612.6
Pressure:	(psia)	30.92	30.92	30.92	30.92
Molar Flow	(lbmole/hr)	6231	4344	1887	0.0000
Mass Flow	(lb/hr)	1.285e+006	5.643e+005	7.205e+005	0.0000
Std Ideal Liq Vol Flow	(USGPM)	2916	1403	1513	0.0000
Molar Enthalpy	(Btu/lbmole)	-1.198e+005	-6.857e+004	-2.378e+005	-1.278e+005
Mass Enthalpy	(Btu/lb)	-581.0	-527.8	-622.7	-619.6
Molar Entropy	(Btu/lbmole-F)	162.9	107.0	290.9	162.9
Mass Entropy	(Btu/lb-F)	0.7899	0.8238	0.7617	0.7899
Heat Flow	(Btu/hr)	-7.465e+008	-2.979e+008	-4.487e+008	0.0000
Molar Density	(lbmole/ft3)	3.956e-003	2.786e-003	0.1174	0.1849
Mass Density	(lb/ft3)	0.8157	0.3620	44.83	34.01
Std Ideal Liq Mass Density	(lb/ft3)	54.93	50.15	59.35	54.93
Liq Mass Density @Std Cond	(lb/ft3)	54.70	50.71	58.98	54.70
Molar Heat Capacity	(Btu/lbmole-F)	136.4	80.91	264.3	151.7
Mass Heat Capacity	(Btu/lb-F)	0.6616	0.8228	0.6921	0.7356
Thermal Conductivity	(Btu/hr-ft-F)	—	2.042e-002	8.110e-002	2.070e-002
Viscosity	(cP)	—	1.155e-002	0.3041	0.1323
Surface Tension	(dyne/cm)	206.2	—	12.81	5.741
Molecular Weight		—	129.9	381.9	206.2
Z Factor			0.9843	2.289e-002	1.629e-002

* Specified by user.

Table 5. 16: properties of stream of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Properties : Main Steam

	Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction	1.0000	1.0000	0.0000	0.0000
Temperature: (F)	375.0	375.0	375.0	375.0
Pressure: (psia)	31.01	31.01	31.01	31.01
Molar Flow (lbmole/hr)	416.3	416.3	0.0000	0.0000
Mass Flow (lb/hr)	7500	7500	0.0000	0.0000
Std Ideal Liq Vol Flow (USGPM)	15.01	15.01	0.0000	0.0000
Molar Enthalpy (Btu/lbmole)	-1.012e+005	-1.012e+005	-1.170e+005	-1.170e+005
Mass Enthalpy (Btu/lb)	-5619	-5619	-6493	-6493
Molar Entropy (Btu/lbmole-F)	43.57	43.57	43.57	43.57
Mass Entropy (Btu/lb-F)	2.418	2.418	2.418	2.418
Heat Flow (Btu/hr)	-4.214e+007	-4.214e+007	0.0000	0.0000
Molar Density (lbmole/ft3)	3.497e-003	3.497e-003	2.993	2.993
Mass Density (lb/ft3)	6.300e-002	6.300e-002	53.93	53.93
Std Ideal Liq Mass Density (lb/ft3)	62.30	62.30	62.30	62.30
Liq Mass Density @Std Cond (lb/ft3)	63.33	63.33	63.33	63.33
Molar Heat Capacity (Btu/lbmole-F)	8.428	8.428	20.75	20.75
Mass Heat Capacity (Btu/lb-F)	0.4679	0.4679	1.152	1.152
Thermal Conductivity (Btu/hr-ft-F)	1.877e-002	1.877e-002	0.3878	0.3878
Viscosity (cP)	1.572e-002	1.572e-002	0.1407	0.1407
Surface Tension (dyne/cm)	—	—	39.94	39.94
Molecular Weight	18.02	18.02	18.02	18.02
Z Factor	0.9899	0.9899	1.156e-003	1.156e-003

Properties : Diesel Steam

	Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction	1.0000	1.0000	0.0000	0.0000
Temperature: (F)	300.0	300.0	300.0	300.0
Pressure: (psia)	30.62	30.62	30.62	30.62
Molar Flow (lbmole/hr)	166.5	166.5	0.0000	0.0000
Mass Flow (lb/hr)	3000	3000	0.0000	0.0000
Std Ideal Liq Vol Flow (USGPM)	6.003	6.003	0.0000	0.0000
Molar Enthalpy (Btu/lbmole)	-1.019e+005	-1.019e+005	-1.185e+005	-1.185e+005
Mass Enthalpy (Btu/lb)	-5654	-5654	-6577	-6577
Molar Entropy (Btu/lbmole-F)	42.80	42.80	42.80	42.80
Mass Entropy (Btu/lb-F)	2.376	2.376	2.376	2.376
Heat Flow (Btu/hr)	-1.696e+007	-1.696e+007	0.0000	0.0000
Molar Density (lbmole/ft3)	3.805e-003	3.805e-003	3.135	3.135
Mass Density (lb/ft3)	6.855e-002	6.855e-002	56.48	56.48
Std Ideal Liq Mass Density (lb/ft3)	62.30	62.30	62.30	62.30
Liq Mass Density @Std Cond (lb/ft3)	63.33	63.33	63.33	63.33
Molar Heat Capacity (Btu/lbmole-F)	8.348	8.348	19.65	19.65
Mass Heat Capacity (Btu/lb-F)	0.4634	0.4634	1.091	1.091
Thermal Conductivity (Btu/hr-ft-F)	1.668e-002	1.668e-002	0.3968	0.3968
Viscosity (cP)	1.399e-002	1.399e-002	0.1825	0.1825
Surface Tension (dyne/cm)	—	—	48.91	48.91
Molecular Weight	18.02	18.02	18.02	18.02
Z Factor	0.9870	0.9870	1.198e-003	1.198e-003

Properties : AGO Steam

	Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction	1.0000	1.0000	0.0000	0.0000
Temperature: (F)	300.0	300.0	300.0	300.0
Pressure: (psia)	30.84	30.84	30.84	30.84
Molar Flow (lbmole/hr)	138.8	138.8	0.0000	0.0000

* Specified by user.

Table 5. 17: properties of stream of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Properties : AGO Steam

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Mass Flow	(lb/hr)	2500	2500	0.0000	0.0000
Std Ideal Liq Vol Flow	(USGPM)	5.003	5.003	0.0000	0.0000
Molar Enthalpy	(Btu/lbmole)	-1.019e+005	-1.019e+005	-1.185e+005	-1.185e+005
Mass Enthalpy	(Btu/lb)	-5654	-5654	-6577	-6577
Molar Entropy	(Btu/lbmole-F)	42.79	42.79	42.79	42.79
Mass Entropy	(Btu/lb-F)	2.375	2.375	2.375	2.375
Heat Flow	(Btu/hr)	-1.413e+007	-1.413e+007	0.0000	0.0000
Molar Density	(lbmole/ft3)	3.833e-003	3.833e-003	3.135	3.135
Mass Density	(lb/ft3)	6.905e-002	6.905e-002	56.48	56.48
Std Ideal Liq Mass Density	(lb/ft3)	62.30	62.30	62.30	62.30
Liq Mass Density @Std Cond	(lb/ft3)	63.33	63.33	63.33	63.33
Molar Heat Capacity	(Btu/lbmole-F)	8.349	8.349	19.65	19.65
Mass Heat Capacity	(Btu/lb-F)	0.4635	0.4635	1.091	1.091
Thermal Conductivity	(Btu/hr-ft-F)	1.669e-002	1.669e-002	0.3968	0.3968
Viscosity	(cP)	1.399e-002	1.399e-002	0.1825	0.1825
Surface Tension	(dyne/cm)	—	—	48.91	48.91
Molecular Weight		18.02	18.02	18.02	18.02
Z Factor		0.9869	0.9869	1.207e-003	1.207e-003

Properties : Naphtha

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction		0.0000	0.0000	1.0000	0.0000
Temperature:	(F)	114.0	114.0	114.0	114.0
Pressure:	(psia)	19.70	19.70	19.70	19.70
Molar Flow	(lbmole/hr)	2759	0.0000	2759	0.0000
Mass Flow	(lb/hr)	2.464e+005	0.0000	2.464e+005	0.0000
Std Ideal Liq Vol Flow	(USGPM)	671.0	0.0000	670.9	0.0000
Molar Enthalpy	(Btu/lbmole)	-8.366e+004	-5.354e+004	-8.365e+004	-1.220e+005
Mass Enthalpy	(Btu/lb)	-936.9	-980.4	-936.7	-6773
Molar Entropy	(Btu/lbmole-F)	31.32	47.40	31.32	31.32
Mass Entropy	(Btu/lb-F)	0.3507	0.8679	0.3507	1.738
Heat Flow	(Btu/hr)	-2.308e+008	0.0000	-2.308e+008	0.0000
Molar Density	(lbmole/ft3)	0.4993	3.296e-003	0.4993	3.437
Mass Density	(lb/ft3)	44.59	0.1800	44.59	61.91
Std Ideal Liq Mass Density	(lb/ft3)	45.78	37.64	45.78	62.30
Liq Mass Density @Std Cond	(lb/ft3)	46.35	38.55	46.35	63.33
Molar Heat Capacity	(Btu/lbmole-F)	44.27	23.21	44.27	18.58
Mass Heat Capacity	(Btu/lb-F)	0.4958	0.4250	0.4958	1.031
Thermal Conductivity	(Btu/hr-ft-F)	6.587e-002	1.028e-002	6.587e-002	0.3688
Viscosity	(cP)	0.3528	8.008e-003	0.3528	0.5880
Surface Tension	(dyne/cm)	18.00	—	18.00	68.52
Molecular Weight		89.30	54.61	89.30	18.02
Z Factor		6.407e-003	0.9708	6.407e-003	9.309e-004

Properties : Residue

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction		0.0000	0.0000	1.0000	0.0000
Temperature:	(F)	667.1	667.1	667.1	667.1
Pressure:	(psia)	30.97	30.97	30.97	30.97
Molar Flow	(lbmole/hr)	1594	0.0000	1594	0.0000
Mass Flow	(lb/hr)	6.981e+005	0.0000	6.981e+005	0.0000
Std Ideal Liq Vol Flow	(USGPM)	1445	0.0000	1445	0.0000
Molar Enthalpy	(Btu/lbmole)	-2.564e+005	-9.948e+004	-2.564e+005	-9.948e+004
Mass Enthalpy	(Btu/lb)	-585.5	-1385	-585.5	-1385

Table 5. 18: properties of stream of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Properties : Residue

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase	
Molar Entropy	(Btu/lbmole-F)	352.2	352.2	352.2	352.2	
Mass Entropy	(Btu/lb-F)	0.8043	4.905	0.8043	4.905	
Heat Flow	(Btu/hr)	-4.088e+008	0.0000	-4.088e+008	0.0000	
Molar Density	(lbmole/ft3)	0.1023	2.590e-003	0.1023	2.590e-003	
Mass Density	(lb/ft3)	44.79	0.1860	44.79	0.1860	
Std Ideal Liq Mass Density	(lb/ft3)	60.24	55.71	60.24	55.71	
Liq Mass Density @Std Cond	(lb/ft3)	60.17	58.35	60.17	58.35	
Molar Heat Capacity	(Btu/lbmole-F)	312.0	43.90	312.0	43.90	
Mass Heat Capacity	(Btu/lb-F)	0.7125	0.6115	0.7125	0.6115	
Thermal Conductivity	(Btu/hr-ft-F)	6.623e-002	2.367e-002	6.623e-002	0.1439	
Viscosity	(cP)	0.2821	1.789e-002	0.2821	2.621e-003	
Surface Tension	(dyne/cm)	12.92	—	12.92	3.724	
Molecular Weight		437.9	71.80	437.9	71.80	
Z Factor		2.504e-002	0.9889	2.504e-002	0.9889	

Properties : Kerosene

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase	
Vapour/Phase Fraction		0.0000	0.0000	1.0000	0.0000	
Temperature:	(F)	459.6	459.6	459.6	459.6	
Pressure:	(psia)	30.00	30.00	30.00	30.00	
Molar Flow	(lbmole/hr)	718.1	0.0000	718.1	0.0000	
Mass Flow	(lb/hr)	1.140e+005	0.0000	1.140e+005	0.0000	
Std Ideal Liq Vol Flow	(USGPM)	271.2	0.0000	271.2	0.0000	
Molar Enthalpy	(Btu/lbmole)	-1.163e+005	-9.277e+004	-1.163e+005	-1.163e+005	
Mass Enthalpy	(Btu/lb)	-732.9	-619.1	-732.9	-732.9	
Molar Entropy	(Btu/lbmole-F)	82.97	82.97	82.97	82.97	
Mass Entropy	(Btu/lb-F)	0.5228	0.5537	0.5228	0.5228	
Heat Flow	(Btu/hr)	-8.352e+007	0.0000	-8.352e+007	0.0000	
Molar Density	(lbmole/ft3)	0.2566	3.296e-003	0.2566	0.2573	
Mass Density	(lb/ft3)	40.72	0.4939	40.72	40.83	
Std Ideal Liq Mass Density	(lb/ft3)	52.39	51.93	52.39	52.39	
Liq Mass Density @Std Cond	(lb/ft3)	52.62	52.18	52.62	52.62	
Molar Heat Capacity	(Btu/lbmole-F)	103.2	81.66	103.2	103.2	
Mass Heat Capacity	(Btu/lb-F)	0.6505	0.5450	0.6505	0.6505	
Thermal Conductivity	(Btu/hr-ft-F)	5.682e-002	1.418e-002	5.682e-002	5.682e-002	
Viscosity	(cP)	0.1797	8.496e-003	0.1797	0.1800	
Surface Tension	(dyne/cm)	10.47	—	10.47	10.47	
Molecular Weight		158.7	149.8	158.7	158.7	
Z Factor		1.185e-002	0.9227	1.185e-002	1.182e-002	

Properties : Diesel

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase	
Vapour/Phase Fraction		0.0000	0.0000	1.0000	0.0000	
Temperature:	(F)	485.9	485.9	485.9	485.9	
Pressure:	(psia)	30.15	30.15	30.15	30.15	
Molar Flow	(lbmole/hr)	1135	0.0000	1135	0.0000	
Mass Flow	(lb/hr)	2.466e+005	0.0000	2.466e+005	0.0000	
Std Ideal Liq Vol Flow	(USGPM)	562.2	0.0000	562.2	0.0000	
Molar Enthalpy	(Btu/lbmole)	-1.553e+005	-1.040e+005	-1.553e+005	-1.145e+005	
Mass Enthalpy	(Btu/lb)	-715.1	-1369	-715.1	-6357	
Molar Entropy	(Btu/lbmole-F)	130.4	130.4	130.4	130.4	
Mass Entropy	(Btu/lb-F)	0.6003	1.716	0.6003	7.237	
Heat Flow	(Btu/hr)	-1.764e+008	0.0000	-1.764e+008	0.0000	
Molar Density	(lbmole/ft3)	0.1987	3.078e-003	0.1987	2.221	

Table 5. 19: properties of stream of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Properties : Diesel

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Mass Density	(lb/ft3)	43.15	0.2339	43.15	40.01
Std Ideal Liq Mass Density	(lb/ft3)	54.70	54.68	54.70	62.30
Liq Mass Density @Std Cond	(lb/ft3)	54.78	57.34	54.78	63.33
Molar Heat Capacity	(Btu/lbmole-F)	140.7	41.35	140.7	24.20
Mass Heat Capacity	(Btu/lb-F)	0.6476	0.5442	0.6476	1.343
Thermal Conductivity	(Btu/hr-ft-F)	6.345e-002	1.777e-002	6.345e-002	0.3545
Viscosity	(cP)	0.2180	1.353e-002	0.2180	0.1061
Surface Tension	(dyne/cm)	12.99	—	12.99	25.61
Molecular Weight		217.2	75.99	217.2	18.02
Z Factor		1.496e-002	0.9654	1.496e-002	1.338e-003

Properties : AGO

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction		0.0000	0.0000	1.0000	0.0000
Temperature:	(F)	568.8	568.8	568.8	568.8
Pressure:	(psia)	30.56	30.56	30.56	30.56
Molar Flow	(lbmole/hr)	202.4	0.0000	202.4	0.0000
Mass Flow	(lb/hr)	6.002e+004	0.0000	6.002e+004	0.0000
Std Ideal Liq Vol Flow	(USGPM)	131.3	0.0000	131.3	0.0000
Molar Enthalpy	(Btu/lbmole)	-1.951e+005	-1.058e+005	-1.951e+005	-1.123e+005
Mass Enthalpy	(Btu/lb)	-657.9	-2017	-657.9	-6238
Molar Entropy	(Btu/lbmole-F)	304.5	304.5	304.5	304.5
Mass Entropy	(Btu/lb-F)	1.027	5.814	1.027	16.90
Heat Flow	(Btu/hr)	-3.948e+007	0.0000	-3.948e+007	0.0000
Molar Density	(lbmole/ft3)	0.1471	2.820e-003	0.1471	1.873
Mass Density	(lb/ft3)	43.61	0.1477	43.61	33.75
Std Ideal Liq Mass Density	(lb/ft3)	56.97	57.62	56.97	62.30
Liq Mass Density @Std Cond	(lb/ft3)	56.88	60.95	56.88	63.33
Molar Heat Capacity	(Btu/lbmole-F)	201.3	29.49	201.3	35.69
Mass Heat Capacity	(Btu/lb-F)	0.6788	0.5830	0.6788	1.981
Thermal Conductivity	(Btu/hr-ft-F)	6.532e-002	2.191e-002	6.532e-002	0.3144
Viscosity	(cP)	0.1292	1.776e-002	0.1292	9.061e-002
Surface Tension	(dyne/cm)	13.15	—	13.15	14.68
Molecular Weight		296.5	52.38	296.5	18.02
Z Factor		1.882e-002	0.9817	1.882e-002	1.478e-003

Properties : Off Gas

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction		1.0000	1.0000	0.0000	0.0000
Temperature:	(F)	114.0	114.0	114.0	114.0
Pressure:	(psia)	19.70	19.70	19.70	19.70
Molar Flow	(lbmole/hr)	7.308	7.308	0.0000	0.0000
Mass Flow	(lb/hr)	399.4	399.1	0.0000	0.0000
Std Ideal Liq Vol Flow	(USGPM)	1.322	1.322	0.0000	0.0000
Molar Enthalpy	(Btu/lbmole)	-5.358e+004	-5.354e+004	-5.365e+004	-1.220e+005
Mass Enthalpy	(Btu/lb)	-981.2	-980.4	-938.7	-6773
Molar Entropy	(Btu/lbmole-F)	46.94	46.94	28.92	46.94
Mass Entropy	(Btu/lb-F)	0.8596	0.8596	0.3239	2.606
Heat Flow	(Btu/hr)	-3.916e+005	-3.913e+005	0.0000	0.0000
Molar Density	(lbmole/ft3)	3.296e-003	3.296e-003	0.4993	3.437
Mass Density	(lb/ft3)	0.1800	0.1800	44.59	61.91
Std Ideal Liq Mass Density	(lb/ft3)	37.64	37.64	46.78	62.30
Liq Mass Density @Std Cond	(lb/ft3)	38.55	38.55	46.35	63.33
Molar Heat Capacity	(Btu/lbmole-F)	23.21	23.21	44.27	18.58

* Specified by user.

Table 5. 20: properties of stream of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Properties : Off Gas

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Mass Heat Capacity	(Btu/lb-F)	0.4250	0.4250	0.4958	1.031
Thermal Conductivity	(Btu/hr-ft-F)	1.028e-002	1.028e-002	6.587e-002	0.3688
Viscosity	(cP)	8.007e-003	8.007e-003	0.3528	0.5880
Surface Tension	(dyne/cm)	---	---	18.00	68.52
Molecular Weight		54.61	54.61	89.30	18.02
Z Factor		0.9708	0.9708	6.407e-003	9.309e-004

Properties : Waste Water

		Overall	Vapour Phase	Liquid Phase	Aqueous Phase
Vapour/Phase Fraction		0.0000	0.0000	0.0000	1.0000
Temperature:	(F)	114.0	114.0	114.0	114.0
Pressure:	(psia)	19.70	19.70	19.70	19.70
Molar Flow	(lbmole/hr)	698.9	0.0000	0.0000	698.9
Mass Flow	(lb/hr)	1.259e+004	0.0000	0.0000	1.259e+004
Std Ideal Liq Vol Flow	(USGPM)	25.19	0.0000	0.0000	25.19
Molar Enthalpy	(Btu/lbmole)	-1.220e+005	-5.354e+004	-8.365e+004	-1.220e+005
Mass Enthalpy	(Btu/lb)	-6773	-980.4	-936.7	-6773
Molar Entropy	(Btu/lbmole-F)	14.69	14.69	14.69	14.69
Mass Entropy	(Btu/lb-F)	0.8154	0.2690	0.1645	0.8154
Heat Flow	(Btu/hr)	-8.527e+007	0.0000	0.0000	-8.527e+007
Molar Density	(lbmole/ft3)	3.437	3.296e-003	0.4993	3.437
Mass Density	(lb/ft3)	61.91	0.1800	44.59	61.91
Std Ideal Liq Mass Density	(lb/ft3)	62.30	37.64	45.78	62.30
Liq Mass Density @Std Cond	(lb/ft3)	63.33	38.55	46.35	63.33
Molar Heat Capacity	(Btu/lbmole-F)	18.58	23.21	44.27	18.58
Mass Heat Capacity	(Btu/lb-F)	1.031	0.4250	0.4958	1.031
Thermal Conductivity	(Btu/hr-ft-F)	0.3688	1.028e-002	6.587e-002	0.3688
Viscosity	(cP)	0.5880	8.008e-003	0.3528	0.5880
Surface Tension	(dyne/cm)	68.52	---	18.00	68.52
Molecular Weight		18.02	54.61	89.30	18.02
Z Factor		9.309e-004	0.9708	6.407e-003	9.309e-004

SUMMARY

Flow Basis: Molar — The composition option is selected

Feed Composition

	Atm Feed	Main Steam	Diesel Steam	AGO Steam
Flow Rate (lbmole/hr)	6.230685e+03	416.3174	166.5270	138.7725
	---	---	---	---
Methane	0.0003	0.0000	0.0000	0.0000
Ethane	0.0006	0.0000	0.0000	0.0000
Propane	0.0086	0.0000	0.0000	0.0000
i-Butane	0.0054	0.0000	0.0000	0.0000
n-Butane	0.0193	0.0000	0.0000	0.0000
H2O	0.0000	1.0000	1.0000	1.0000
NBP[0]49*	0.0364	0.0000	0.0000	0.0000
NBP[0]79*	0.0436	0.0000	0.0000	0.0000
NBP[0]111*	0.0427	0.0000	0.0000	0.0000
NBP[0]144*	0.0416	0.0000	0.0000	0.0000
NBP[0]176*	0.0438	0.0000	0.0000	0.0000
NBP[0]208*	0.0451	0.0000	0.0000	0.0000
NBP[0]240*	0.0437	0.0000	0.0000	0.0000
NBP[0]272*	0.0418	0.0000	0.0000	0.0000
NBP[0]304*	0.0391	0.0000	0.0000	0.0000

Table 5. 21: feed composition of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

SUMMARY

	Atm Feed	Main Steam	Diesel Steam	AGO Steam
NBP[0]336*	0.0362	0.0000	0.0000	0.0000
NBP[0]368*	0.0342	0.0000	0.0000	0.0000
NBP[0]400*	0.0334	0.0000	0.0000	0.0000
NBP[0]433*	0.0356	0.0000	0.0000	0.0000
NBP[0]464*	0.0394	0.0000	0.0000	0.0000
NBP[0]496*	0.0393	0.0000	0.0000	0.0000
NBP[0]528*	0.0356	0.0000	0.0000	0.0000
NBP[0]560*	0.0314	0.0000	0.0000	0.0000
NBP[0]592*	0.0279	0.0000	0.0000	0.0000
NBP[0]624*	0.0242	0.0000	0.0000	0.0000
NBP[0]656*	0.0213	0.0000	0.0000	0.0000
NBP[0]688*	0.0192	0.0000	0.0000	0.0000
NBP[0]720*	0.0172	0.0000	0.0000	0.0000
NBP[0]752*	0.0153	0.0000	0.0000	0.0000
NBP[0]784*	0.0138	0.0000	0.0000	0.0000
NBP[0]830*	0.0229	0.0000	0.0000	0.0000
NBP[0]888*	0.0212	0.0000	0.0000	0.0000
NBP[0]947*	0.0209	0.0000	0.0000	0.0000
NBP[0]1009*	0.0267	0.0000	0.0000	0.0000
NBP[0]1062*	0.0364	0.0000	0.0000	0.0000
NBP[0]1124*	0.0358	0.0000	0.0000	0.0000

Flow Basis: Molar — The composition option is selected

Feed Flows

	Atm Feed	Main Steam	Diesel Steam	AGO Steam
Flow Rate (lbmole/hr)	6.230685e+03	416.3174	166.5270	138.7725
	—	—	—	—
Methane (lbmole/hr)	1.7716	0.0000	0.0000	0.0000
Ethane (lbmole/hr)	3.8868	0.0000	0.0000	0.0000
Propane (lbmole/hr)	53.6976	0.0000	0.0000	0.0000
i-Butane (lbmole/hr)	33.8880	0.0000	0.0000	0.0000
n-Butane (lbmole/hr)	120.1634	0.0000	0.0000	0.0000
H2O (lbmole/hr)	0.0000	416.3174	166.5270	138.7725
NBP[0]49* (lbmole/hr)	226.6922	0.0000	0.0000	0.0000
NBP[0]79* (lbmole/hr)	271.5848	0.0000	0.0000	0.0000
NBP[0]111* (lbmole/hr)	266.1628	0.0000	0.0000	0.0000
NBP[0]144* (lbmole/hr)	259.2997	0.0000	0.0000	0.0000
NBP[0]176* (lbmole/hr)	273.0171	0.0000	0.0000	0.0000
NBP[0]208* (lbmole/hr)	280.7032	0.0000	0.0000	0.0000
NBP[0]240* (lbmole/hr)	272.5778	0.0000	0.0000	0.0000
NBP[0]272* (lbmole/hr)	260.4584	0.0000	0.0000	0.0000
NBP[0]304* (lbmole/hr)	243.5817	0.0000	0.0000	0.0000
NBP[0]336* (lbmole/hr)	225.5531	0.0000	0.0000	0.0000
NBP[0]368* (lbmole/hr)	213.0128	0.0000	0.0000	0.0000
NBP[0]400* (lbmole/hr)	208.0350	0.0000	0.0000	0.0000
NBP[0]433* (lbmole/hr)	222.1032	0.0000	0.0000	0.0000
NBP[0]464* (lbmole/hr)	245.5658	0.0000	0.0000	0.0000
NBP[0]496* (lbmole/hr)	244.7402	0.0000	0.0000	0.0000
NBP[0]528* (lbmole/hr)	221.9375	0.0000	0.0000	0.0000
NBP[0]560* (lbmole/hr)	195.5413	0.0000	0.0000	0.0000
NBP[0]592* (lbmole/hr)	173.6917	0.0000	0.0000	0.0000
NBP[0]624* (lbmole/hr)	150.9855	0.0000	0.0000	0.0000

* Specified by user.

Table 5. 22: product composition of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

SUMMARY

	Atm Feed	Main Steam	Diesel Steam	AGO Steam	
NBP[0]656* (lbmole/hr)	132.5358	0.0000	0.0000	0.0000	
NBP[0]688* (lbmole/hr)	119.3544	0.0000	0.0000	0.0000	
NBP[0]720* (lbmole/hr)	107.1919	0.0000	0.0000	0.0000	
NBP[0]752* (lbmole/hr)	95.1681	0.0000	0.0000	0.0000	
NBP[0]784* (lbmole/hr)	86.1128	0.0000	0.0000	0.0000	
NBP[0]830* (lbmole/hr)	142.9682	0.0000	0.0000	0.0000	
NBP[0]888* (lbmole/hr)	131.8298	0.0000	0.0000	0.0000	
NBP[0]947* (lbmole/hr)	130.3381	0.0000	0.0000	0.0000	
NBP[0]1009* (lbmole/hr)	166.3298	0.0000	0.0000	0.0000	
NBP[0]1062* (lbmole/hr)	227.1031	0.0000	0.0000	0.0000	
NBP[0]1124* (lbmole/hr)	223.1021	0.0000	0.0000	0.0000	

Products

Flow Basis: Molar — The composition option is selected

Product Compositions

	Off Gas	Naphtha	Waste Water	Residue	Kerosene
Flow Rate (lbmole/hr)	7.3082	2.758898e+03	698.8582	1.594396e+03	718.0960
	---	---	---	---	---
Methane	0.0512	0.0003	0.0000	0.0000	0.0000
Ethane	0.0347	0.0010	0.0000	0.0000	0.0000
Propane	0.1788	0.0173	0.0000	0.0000	0.0000
i-Butane	0.0517	0.0117	0.0000	0.0000	0.0000
n-Butane	0.1332	0.0419	0.0000	0.0000	0.0000
H2O	0.0713	0.0012	1.0000	0.0067	0.0000
NBP[0]49*	0.1775	0.0800	0.0000	0.0001	0.0000
NBP[0]79*	0.1350	0.0968	0.0000	0.0001	0.0000
NBP[0]111*	0.0758	0.0956	0.0000	0.0001	0.0000
NBP[0]144*	0.0414	0.0935	0.0000	0.0002	0.0000
NBP[0]176*	0.0240	0.0987	0.0000	0.0002	0.0001
NBP[0]208*	0.0133	0.1016	0.0000	0.0003	0.0005
NBP[0]240*	0.0067	0.0985	0.0000	0.0004	0.0018
NBP[0]272*	0.0032	0.0929	0.0000	0.0006	0.0055
NBP[0]304*	0.0014	0.0829	0.0000	0.0008	0.0149
NBP[0]336*	0.0005	0.0655	0.0000	0.0011	0.0475
NBP[0]368*	0.0001	0.0197	0.0000	0.0016	0.1970
NBP[0]400*	0.0000	0.0008	0.0000	0.0022	0.2533
NBP[0]433*	0.0000	0.0000	0.0000	0.0036	0.2452
NBP[0]464*	0.0000	0.0000	0.0000	0.0058	0.1893
NBP[0]496*	0.0000	0.0000	0.0000	0.0085	0.0424
NBP[0]528*	0.0000	0.0000	0.0000	0.0113	0.0023
NBP[0]560*	0.0000	0.0000	0.0000	0.0144	0.0001
NBP[0]592*	0.0000	0.0000	0.0000	0.0182	0.0000
NBP[0]624*	0.0000	0.0000	0.0000	0.0221	0.0000
NBP[0]656*	0.0000	0.0000	0.0000	0.0265	0.0000
NBP[0]688*	0.0000	0.0000	0.0000	0.0316	0.0000
NBP[0]720*	0.0000	0.0000	0.0000	0.0366	0.0000
NBP[0]752*	0.0000	0.0000	0.0000	0.0410	0.0000
NBP[0]784*	0.0000	0.0000	0.0000	0.0465	0.0000
NBP[0]830*	0.0000	0.0000	0.0000	0.0959	0.0000
NBP[0]888*	0.0000	0.0000	0.0000	0.0929	0.0000
NBP[0]947*	0.0000	0.0000	0.0000	0.0924	0.0000
NBP[0]1009*	0.0000	0.0000	0.0000	0.1181	0.0000

* Specified by user.

Table 5. 23: product composition of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

SUMMARY

	Off Gas	Naphtha	Waste Water	Residue	Kerosene
NBP[0]1062*	0.0000	0.0000	0.0000	0.1614	0.0000
NBP[0]1124*	0.0000	0.0000	0.0000	0.1586	0.0000
	Diesel	AGO			
Flow Rate (lbmole/hr)	1.135494e+03	202.4290			
	---	---			
Methane	0.0000	0.0000			
Ethane	0.0000	0.0000			
Propane	0.0000	0.0000			
i-Butane	0.0000	0.0000			
n-Butane	0.0000	0.0000			
H2O	0.0062	0.0077			
NBP[0]49*	0.0000	0.0000			
NBP[0]79*	0.0000	0.0000			
NBP[0]111*	0.0000	0.0000			
NBP[0]144*	0.0001	0.0000			
NBP[0]176*	0.0001	0.0000			
NBP[0]208*	0.0003	0.0000			
NBP[0]240*	0.0006	0.0000			
NBP[0]272*	0.0013	0.0000			
NBP[0]304*	0.0027	0.0000			
NBP[0]336*	0.0051	0.0001			
NBP[0]368*	0.0093	0.0002			
NBP[0]400*	0.0170	0.0005			
NBP[0]433*	0.0354	0.0013			
NBP[0]464*	0.0887	0.0035			
NBP[0]496*	0.1748	0.0080			
NBP[0]528*	0.1769	0.0155			
NBP[0]560*	0.1504	0.0272			
NBP[0]592*	0.1239	0.0453			
NBP[0]624*	0.0943	0.0716			
NBP[0]656*	0.0648	0.1142			
NBP[0]688*	0.0353	0.1803			
NBP[0]720*	0.0109	0.2212			
NBP[0]752*	0.0017	0.1818			
NBP[0]784*	0.0002	0.1003			
NBP[0]830*	0.0000	0.0211			
NBP[0]888*	0.0000	0.0003			
NBP[0]947*	0.0000	0.0000			
NBP[0]1009*	0.0000	0.0000			
NBP[0]1062*	0.0000	0.0000			
NBP[0]1124*	0.0000	0.0000			
Flow Basis:		Molar	The composition option is selected		

Product Flows

	Off Gas	Naphtha	Waste Water	Residue	Kerosene
Flow Rate (lbmole/hr)	7.3082 *	2.758898e+03 *	698.8582 *	1.594396e+03 *	718.0960 *
	---	---	---	---	---
Methane (lbmole/hr)	0.3744 *	0.8192 *	0.0000 *	0.0001 *	0.0000 *
Ethane (lbmole/hr)	0.2537 *	2.8162 *	0.0000 *	0.0003 *	0.0000 *
Propane (lbmole/hr)	1.3070 *	47.7343 *	0.0000 *	0.0088 *	0.0000 *
i-Butane (lbmole/hr)	0.3778 *	32.1524 *	0.0000 *	0.0093 *	0.0000 *
n-Butane (lbmole/hr)	0.9737 *	115.7150 *	0.0000 *	0.0381 *	0.0001 *

Table 5. 24: product flow of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

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 *
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 *
NBP[0]336* (lbmole/hr)	0.0039 *	180.7228 *	0.0000 *	1.7891 *	34.0955 *
NBP[0]368* (lbmole/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 *
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 *
NBP[0]624* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	35.2450 *	0.0000 *
NBP[0]656* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	42.2105 *	0.0000 *
NBP[0]688* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	50.3987 *	0.0000 *
NBP[0]720* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	58.2931 *	0.0000 *
NBP[0]752* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	65.3378 *	0.0000 *
NBP[0]784* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	74.2189 *	0.0000 *
NBP[0]830* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	152.8483 *	0.0000 *
NBP[0]888* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	148.1189 *	0.0000 *
NBP[0]947* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	147.3449 *	0.0000 *
NBP[0]1009* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	188.3514 *	0.0000 *
NBP[0]1062* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	257.3466 *	0.0000 *
NBP[0]1124* (lbmole/hr)	0.0000 *	0.0000 *	0.0000 *	252.9099 *	0.0000 *
	Diesel	AGO			
Flow Rate (lbmole/hr)	1.135494e+03 *	202.4290 *			
	---	---			
Methane (lbmole/hr)	0.0000 *	0.0000 *			
Ethane (lbmole/hr)	0.0000 *	0.0000 *			
Propane (lbmole/hr)	0.0001 *	0.0000 *			
i-Butane (lbmole/hr)	0.0002 *	0.0000 *			
n-Butane (lbmole/hr)	0.0011 *	0.0000 *			
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 *			
NBP[0]176* (lbmole/hr)	0.1253 *	0.0002 *			
NBP[0]208* (lbmole/hr)	0.3071 *	0.0005 *			
NBP[0]240* (lbmole/hr)	0.7026 *	0.0012 *			
NBP[0]272* (lbmole/hr)	1.5272 *	0.0028 *			
NBP[0]304* (lbmole/hr)	3.0827 *	0.0065 *			
NBP[0]336* (lbmole/hr)	5.8238 *	0.0152 *			
NBP[0]368* (lbmole/hr)	10.6005 *	0.0370 *			

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* Specified by user.

Table 5. 25: product recoveries of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

SUMMARY

	Diesel	AGO
NBP[0]400* (lbmole/hr)	19.3232 *	0.0936 *
NBP[0]433* (lbmole/hr)	40.2346 *	0.2606 *
NBP[0]464* (lbmole/hr)	100.6787 *	0.7044 *
NBP[0]496* (lbmole/hr)	198.4789 *	1.6126 *
NBP[0]528* (lbmole/hr)	200.9207 *	3.1284 *
NBP[0]560* (lbmole/hr)	170.7404 *	5.5132 *
NBP[0]592* (lbmole/hr)	140.6510 *	9.1767 *
NBP[0]624* (lbmole/hr)	107.0700 *	14.4953 *
NBP[0]656* (lbmole/hr)	73.5300 *	23.1158 *
NBP[0]688* (lbmole/hr)	40.0959 *	36.4920 *
NBP[0]720* (lbmole/hr)	12.3762 *	44.7791 *
NBP[0]752* (lbmole/hr)	1.9266 *	36.7959 *
NBP[0]784* (lbmole/hr)	0.1725 *	20.3028 *
NBP[0]830* (lbmole/hr)	0.0023 *	4.2747 *
NBP[0]888* (lbmole/hr)	0.0000 *	0.0564 *
NBP[0]947* (lbmole/hr)	0.0000 *	0.0005 *
NBP[0]1009* (lbmole/hr)	0.0000 *	0.0000 *
NBP[0]1062* (lbmole/hr)	0.0000 *	0.0000 *
NBP[0]1124* (lbmole/hr)	0.0000 *	0.0000 *

Flow Basis: Molar — The composition option is selected

Product Recoveries

	Off Gas	Naphtha	Waste Water	Residue	Kerosene
Flow Rate (lbmole/hr)	7.3082	2.758898e+03	698.8582	1.594396e+03	718.0960
	---	---	---	---	---
Methane (%)	21.1318	46.2430	0.0000	0.0036	0.0000
Ethane (%)	6.5263	72.4541	0.0000	0.0085	0.0000
Propane (%)	2.4340	88.8946	0.0000	0.0164	0.0000
i-Butane (%)	1.1148	94.8784	0.0000	0.0273	0.0001
n-Butane (%)	0.8103	96.2980	0.0000	0.0317	0.0001
H2O (%)	0.0722	0.4410	96.8462	1.4752	0.0000
NBP[0]49* (%)	0.5722	97.4056	0.0000	0.0493	0.0004
NBP[0]79* (%)	0.3633	98.3747	0.0000	0.0610	0.0011
NBP[0]111* (%)	0.2081	99.0808	0.0000	0.0784	0.0035
NBP[0]144* (%)	0.1166	99.4861	0.0000	0.1036	0.0115
NBP[0]176* (%)	0.0644	99.7361	0.0000	0.1324	0.0356
NBP[0]208* (%)	0.0346	99.8769	0.0000	0.1854	0.1313
NBP[0]240* (%)	0.0180	99.7106	0.0000	0.2630	0.4695
NBP[0]272* (%)	0.0090	98.3943	0.0000	0.3766	1.5210
NBP[0]304* (%)	0.0043	93.9102	0.0000	0.5438	4.3961
NBP[0]336* (%)	0.0017	80.1243	0.0000	0.7932	15.1164
NBP[0]368* (%)	0.0003	25.5515	0.0000	1.1645	66.4094
NBP[0]400* (%)	0.0000	1.0332	0.0000	1.7207	87.4195
NBP[0]433* (%)	0.0000	0.0202	0.0000	2.5662	79.2881
NBP[0]464* (%)	0.0000	0.0003	0.0000	3.7852	55.3596
NBP[0]496* (%)	0.0000	0.0000	0.0000	5.5526	12.4467
NBP[0]528* (%)	0.0000	0.0000	0.0000	8.0995	0.7457
NBP[0]560* (%)	0.0000	0.0000	0.0000	11.7420	0.0279
NBP[0]592* (%)	0.0000	0.0000	0.0000	16.7205	0.0009
NBP[0]624* (%)	0.0000	0.0000	0.0000	23.3433	0.0000
NBP[0]656* (%)	0.0000	0.0000	0.0000	31.8484	0.0000
NBP[0]688* (%)	0.0000	0.0000	0.0000	42.2261	0.0000

* Specified by user.

Table 5. 26: product recoveries of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

SUMMARY

	Off Gas	Naphtha	Waste Water	Residue	Kerosene
NBP[0]720* (%)	0.0000	0.0000	0.0000	54.3820	0.0000
NBP[0]752* (%)	0.0000	0.0000	0.0000	68.6552	0.0000
NBP[0]784* (%)	0.0000	0.0000	0.0000	86.1879	0.0000
NBP[0]830* (%)	0.0000	0.0000	0.0000	106.9107	0.0000
NBP[0]888* (%)	0.0000	0.0000	0.0000	112.3561	0.0000
NBP[0]947* (%)	0.0000	0.0000	0.0000	113.0482	0.0000
NBP[0]1009* (%)	0.0000	0.0000	0.0000	113.2397	0.0000
NBP[0]1062* (%)	0.0000	0.0000	0.0000	113.3171	0.0000
NBP[0]1124* (%)	0.0000	0.0000	0.0000	113.3606	0.0000
	Diesel	AGO			
Flow Rate (lbmole/hr)	1.135494e+03	202.4290			
	---	---			
Methane (%)	0.0000	0.0000			
Ethane (%)	0.0000	0.0000			
Propane (%)	0.0001	0.0000			
i-Butane (%)	0.0006	0.0000			
n-Butane (%)	0.0009	0.0000			
H2O (%)	0.9727	0.2167			
NBP[0]49* (%)	0.0025	0.0000			
NBP[0]79* (%)	0.0047	0.0000			
NBP[0]111* (%)	0.0100	0.0000			
NBP[0]144* (%)	0.0222	0.0000			
NBP[0]176* (%)	0.0459	0.0001			
NBP[0]208* (%)	0.1094	0.0002			
NBP[0]240* (%)	0.2578	0.0004			
NBP[0]272* (%)	0.5863	0.0011			
NBP[0]304* (%)	1.2656	0.0026			
NBP[0]336* (%)	2.5820	0.0067			
NBP[0]368* (%)	4.9764	0.0174			
NBP[0]400* (%)	9.2884	0.0450			
NBP[0]433* (%)	18.1153	0.1173			
NBP[0]464* (%)	40.9987	0.2868			
NBP[0]496* (%)	81.0978	0.6589			
NBP[0]528* (%)	90.5303	1.4096			
NBP[0]560* (%)	87.3168	2.8195			
NBP[0]592* (%)	80.9774	5.2833			
NBP[0]624* (%)	70.9141	9.6004			
NBP[0]656* (%)	55.4794	17.4412			
NBP[0]688* (%)	33.5940	30.5745			
NBP[0]720* (%)	11.5458	41.7747			
NBP[0]752* (%)	2.0245	38.6641			
NBP[0]784* (%)	0.2003	23.5770			
NBP[0]830* (%)	0.0016	2.9900			
NBP[0]888* (%)	0.0000	0.0428			
NBP[0]947* (%)	0.0000	0.0003			
NBP[0]1009* (%)	0.0000	0.0000			
NBP[0]1062* (%)	0.0000	0.0000			
NBP[0]1124* (%)	0.0000	0.0000			

COLUMN PROFILES

Reflux Ratio: 0.7194 | Reboil Ratio: 0.3996 | The Flows Option is Selected | Flow Basis: Molar

* Specified by user.

Table 5. 27: column profiles of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

COLUMN PROFILES

Column Profiles Flows

	Temperature	Pressure	Net Liq (lbmole/hr)	Net Vap (lbmole/hr)	Net Feed (lbmole/hr)	Net Draws (lbmole/hr)
*Condenser			1985	---	---	3465
1_Main TS	276.6	28.70	9273	5421	---	---
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	---	---
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	---	---
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	2666	7036	---	---
11_Main TS	437.3	29.51	2534	6839	---	---
12_Main TS	447.2	29.59	2429	6712	---	---
13_Main TS	455.0	29.67	2321	6612	---	---
14_Main TS	462.1	29.75	2180	6509	---	---
15_Main TS	470.5	29.84	1945	6373	---	---
16_Main TS	483.7	29.92	4539	6145	---	---
17_Main TS	511.5	30.00	971.2	6538	---	3172
18_Main TS	552.6	30.08	710.1	6150	---	---
19_Main TS	571.7	30.16	590.5	5894	---	---
20_Main TS	580.6	30.24	508.2	5777	---	---
21_Main TS	586.0	30.32	2467	5697	---	---
22_Main TS	613.0	30.40	717.5	6017	---	1703
23_Main TS	634.3	30.48	601.9	5970	---	---
24_Main TS	643.9	30.57	509.0	5853	---	---
25_Main TS	650.2	30.65	426.4	5759	---	---
26_Main TS	655.5	30.73	321.0	5674	---	---
27_Main TS	662.4	30.81	123.7	5566	---	---
28_Main TS	676.1	30.89	1556	5366	6231	---
29_Main TS	667.1	30.97	---	566.7	416.3	1594
1_Kero_SS	430.2	29.35	1027	---	---	210.1
2_Kero_SS	440.7	29.47	1067	309.5	---	---
3_Kero_SS	449.0	30.00	1089	349.5	---	---
Kero_SS_Reb			---	371.3	---	718.1
1_Diesel_SS	504.1	30.00	1274	---	---	383.1
2_Diesel_SS	498.1	30.08	1223	304.8	---	---
3_Diesel_SS	485.9	30.15	---	254.6	166.5	1135
1_AGO_SS	600.3	30.40	243.6	---	---	211.0
2_AGO_SS	589.6	30.48	225.8	180.0	---	---
3_AGO_SS	568.8	30.56	---	162.2	138.8	202.4

Column Profiles Energy

	Temperature	Liquid Enthalpy	Vapour Enthalpy	Heat Loss (Btu/hr)
*Condenser				---
1_Main TS	276.6	-9.829e+004	-6.850e+004	---
2_Main TS	317.1	-1.004e+005	-7.331e+004	---
3_Main TS	343.5	-1.032e+005	-7.705e+004	---
4_Main TS	357.0	-1.051e+005	-7.866e+004	---
5_Main TS	366.3	-1.068e+005	-7.955e+004	---
6_Main TS	374.5	-1.085e+005	-8.015e+004	---
7_Main TS	383.1	-1.107e+005	-8.064e+004	---

Table 5. 28: column profiles of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

COLUMN PROFILES

	Temperature	Liquid Enthalpy	Vapour Enthalpy	Heat Loss (Btu/hr)
8_Main TS	393.9	-1.136e+005	-8.112e+004	---
9_Main TS	407.0	-1.174e+005	-8.156e+004	---
10_Main TS	423.8	-1.216e+005	-8.226e+004	---
11_Main TS	437.3	-1.247e+005	-8.288e+004	---
12_Main TS	447.2	-1.271e+005	-8.331e+004	---
13_Main TS	455.0	-1.291e+005	-8.353e+004	---
14_Main TS	462.1	-1.314e+005	-8.357e+004	---
15_Main TS	470.5	-1.348e+005	-8.338e+004	---
16_Main TS	483.7	-1.417e+005	-8.268e+004	---
17_Main TS	511.5	-1.460e+005	-8.423e+004	---
18_Main TS	552.6	-1.594e+005	-8.352e+004	---
19_Main TS	571.7	-1.659e+005	-8.243e+004	---
20_Main TS	580.6	-1.698e+005	-8.155e+004	---
21_Main TS	586.0	-1.733e+005	-8.073e+004	---
22_Main TS	613.0	-1.736e+005	-8.247e+004	---
23_Main TS	634.3	-1.788e+005	-8.185e+004	---
24_Main TS	643.9	-1.835e+005	-8.058e+004	---
25_Main TS	650.2	-1.879e+005	-7.939e+004	---
26_Main TS	655.5	-1.931e+005	-7.818e+004	---
27_Main TS	662.4	-2.041e+005	-7.636e+004	---
28_Main TS	676.1	-2.403e+005	-7.228e+004	---
29_Main TS	667.1	-2.564e+005	-9.948e+004	---
1_Kero_SS	430.2	-1.153e+005	-8.551e+004	---
2_Kero_SS	440.7	-1.150e+005	-8.879e+004	---
3_Kero_SS	449.0	-1.152e+005	-9.078e+004	---
Kero_SS_Reb				---
1_Diesel_SS	504.1	-1.497e+005	-9.923e+004	---
2_Diesel_SS	498.1	-1.519e+005	-1.027e+005	---
3_Diesel_SS	485.9	-1.553e+005	-1.040e+005	---
1_AGO_SS	600.3	-1.831e+005	-1.057e+005	---
2_AGO_SS	589.6	-1.881e+005	-1.069e+005	---
3_AGO_SS	568.8	-1.951e+005	-1.056e+005	---

FEEDS / PRODUCTS

Flow Basis: Molar

	Stream	Type	Duty (Btu/hr)	State	Flows (lbmole/hr)	Enthalpy (Btu/lbmol)	Temp (F)
*Condenser	Atmos Cond	Energy	1.084e+008 *		---	---	---
	Off Gas	Draw	---	Vapour	7.308 *	-3345 *	114.0 *
	Naphtha	Draw	---	Liquid	2759 *	-5223 *	114.0 *
	Waste Water	Draw	---	Water	698.9 *	-7617 *	114.0 *
1_Main TS	<PA_1>	Energy	-5.501e+007 *		---	---	---
	PA_1_Return	Feed	---	Liquid	---	-4.809e+004 *	139.8 *
2_Main TS	PA_1_Draw	Draw	---	Liquid	4794 *	-6267 *	317.1 *
3_Main TS							
4_Main TS							
5_Main TS							
6_Main TS							
7_Main TS							
8_Main TS	<PA_4>	Energy	---		---	---	---
	Kero_SS_Return	Feed	---	Vapour	---	-3.676e+004 *	430.1 *
9_Main TS	Kero_SS_Draw	Draw	---	Liquid	928.0 *	-7330 *	407.0 *

* Specified by user.

Table 5. 29: feed/products of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

FEEDS / PRODUCTS

10 Main TS								
11 Main TS								
12 Main TS								
13 Main TS								
14 Main TS								
15 Main TS								
16 Main TS	<PA_6>	Energy	-3.500e+007 *		---	---	---	
	Diesel_SS_Return	Feed	---	Vapour	---	-4.266e+004 *	504.1 *	
	<PA_2>	Energy	-3.500e+007 *		---	---	---	
	PA_2_Return	Feed	---	Liquid	---	-7.104e+004 *	364.6 *	
17 Main TS	Diesel_SS_Draw	Draw	---	Liquid	1352 *	-9117 *	511.5 *	
	PA_2_Draw	Draw	---	Liquid	1820 *	-9117 *	511.5 *	
18 Main TS								
19 Main TS								
20 Main TS								
21 Main TS	<PA_5>	Energy	-3.500e+007 *		---	---	---	
	AGO_SS_Return	Feed	---	Vapour	---	-4.546e+004 *	600.3 *	
	<PA_3>	Energy	-3.500e+007 *		---	---	---	
	PA_3_Return	Feed	---	Liquid	---	-8.514e+004 *	481.4 *	
22 Main TS	AGO_SS_Draw	Draw	---	Liquid	274.6 *	-1.083e+004 *	613.0 *	
	PA_3_Draw	Draw	---	Liquid	1429 *	-1.083e+004 *	613.0 *	
23 Main TS								
24 Main TS								
25 Main TS								
26 Main TS								
27 Main TS								
28 Main TS	Q-Trim	Energy	6.642e+007 *		---	---	---	
	Atm Feed	Feed	---	Mixed	6231 *	-5.151e+004 *	612.6 *	
29 Main TS	Main Steam	Feed	---	Vapour	416.3 *	-4.352e+004 *	375.0 *	
	Residue	Draw	---	Liquid	1594 *	-1.600e+004 *	667.1 *	
1 Kero_SS	Kero_SS_Draw	Feed	---	Liquid	---	-5.048e+004 *	407.0 *	
	1	Draw	---	Vapour	210.1 *	-5338 *	430.2 *	
2 Kero_SS								
3 Kero_SS								
Kero_SS_Reb	Kero_SS_Energy	Energy	7.500e+006 *		---	---	---	
	Kerosene	Draw	---	Liquid	718.1 *	-7261 *	459.6 *	
1 Diesel_SS	Diesel_SS_Draw	Feed	---	Liquid	---	-6.278e+004 *	511.5 *	
	2	Draw	---	Vapour	383.1 *	-6195 *	504.1 *	
2 Diesel_SS								
3 Diesel_SS	Diesel Steam	Feed	---	Vapour	166.5 *	-4.379e+004 *	300.0 *	
	Diesel	Draw	---	Liquid	1135 *	-9695 *	485.9 *	
1 AGO_SS	AGO_SS_Draw	Feed	---	Liquid	---	-7.462e+004 *	613.0 *	
	3	Draw	---	Vapour	211.0 *	-6601 *	600.3 *	
2 AGO_SS								
3 AGO_SS	AGO Steam	Feed	---	Vapour	138.8 *	-4.379e+004 *	300.0 *	
	AGO	Draw	---	Liquid	202.4 *	-1.218e+004 *	568.8 *	

SETUP

Column Flowsheet Topology

Total Theor. Stages:	40 *	Total Tray-Sections:	4 *	Condenser + Reboiler:	2 *	Pump Arounds:	6 *
Side Strippers:	3 *	Side Rectifiers:	0 *	Vapour Bypasses:	0 *		

* Specified by user.

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

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name:	C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set:	Field-USGPM
Date/Time:	Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Sub-Flowsheet

Feed Streams

Internal Stream	External Stream	Transfer Basis
Main Steam	Main Steam @Main	T-P Flash
Q-Trim	Q-Trim @Main	None Req'd
Atm Feed	Atm Feed @Main	T-P Flash
Kero_SS_Energy		None Req'd
Diesel Steam	Diesel Steam @Main	T-P Flash
AGO Steam	AGO Steam @Main	T-P Flash
atm residue		<None Set>
vac duty		None Req'd

Product Streams

Internal Stream	External Stream	Transfer Basis
Residue	Residue @Main	T-P Flash
Atmos Cond	Atmos Cond @Main	None Req'd
Off Gas	Off Gas @Main	T-P Flash
Waste Water	Waste Water @Main	T-P Flash
Naphtha	Naphtha @Main	T-P Flash
Kerosene	Kerosene @Main	T-P Flash
Diesel	Diesel @Main	T-P Flash
AGO	AGO @Main	T-P Flash
PA_1_Q		None Req'd
PA_2_Q		None Req'd
PA_3_Q		None Req'd
vac feed		<None Set>

VARIABLES

Column Flowsheet Vars Available as Parameters

Data Source	Variable	Component	Description

COMPONENT MAPS

Feed Streams

Feed Name	In to SubFlowSheet	Out of SubFlowSheet
Main Steam		
Q-Trim		
Atm Feed		
Kero_SS_Energy		
Diesel Steam		
AGO Steam		
atm residue		
vac duty		

Product Stream

Product Name	In to SubFlowSheet	Out of SubFlowSheet
Residue		
Atmos Cond		
Off Gas		
Waste Water		
Naphtha		
Kerosene		
Diesel		
AGO		
PA_1_Q		
PA_2_Q		
PA_3_Q		
vac feed		

DYNAMICS

Vessel Dynamic Specifications

		Condenser	Kero_SS_Reb	
Vessel				
Diameter	(ft)	13.12 *	5.235	
Height.0	(ft)	13.12 *	3.281	

Table 5. 31: holdup details of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Samples\dyncrude3.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

Volume.0	(ft3)	1759	70.63 *	
Liquid Volume Percent	(%)	30.05 *	50.05 *	
Level 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	30.00	
Fixed P Spec Active		Not Active	Not Active	

Other Equipment in Column Flowsheet

PA_1_Cooler @COL1
PA_2_Cooler @COL1
PA_3_Cooler @COL1
VLV-100 @COL1
VLV-101 @COL1
VLV-102 @COL1
vac heater @COL1

Holdup Details

	Pressure (psia)	Volume (ft3)	Bulk Liquid Volume (ft3)
*Condenser	19.70	1759 *	528.5
1_Main TS	28.70	3123 *	512.8
2_Main TS	28.80	3123 *	433.2
3_Main TS	28.90	3123 *	439.3
4_Main TS	28.99	3123 *	441.1
5_Main TS	29.08	3123 *	441.1
6_Main TS	29.17	3123 *	439.9
7_Main TS	29.27	3123 *	437.6
8_Main TS	29.36	3123 *	434.2
9_Main TS	29.45	3123 *	401.3
10_Main TS	29.53	3123 *	399.2
11_Main TS	29.62	3123 *	398.1
12_Main TS	29.70	3123 *	396.8
13_Main TS	29.78	3123 *	394.8
14_Main TS	29.87	3123 *	391.3
15_Main TS	29.95	3123 *	384.4
16_Main TS	30.03	3123 *	487.3
17_Main TS	30.13	3123 *	345.1
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	30.50	3123 *	344.3
23_Main TS	30.58	3123 *	337.9
24_Main TS	30.65	3123 *	331.3
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	3123 *	437.8
29_Main TS	31.00	3123 *	369.2
1_Kero_SS	29.53	19.97 *	3.800
2_Kero_SS	29.75	19.97 *	3.863
3_Kero_SS	30.00	19.97 *	3.905
Kero_SS_Reb	30.00	70.63 *	35.35

Licensed to: TEAM LND

* Specified by user.

Table 5. 32: holdup details of ADU

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name:	C:\Program Files\Hyprotech\HYSYS 3,1.3\Samples\dyncrude3.hsc
Unit Set:	Field-USGPM
Date/Time:	Sun May 03 14:58:57 2009

Column Sub-Flowsheet: T-100 @Main (continued)

	Pressure (psia)	Volume (ft3)	Bulk Liquid Volume (ft3)
1_Diesel_SS	30.20	124.8 *	24.19
2_Diesel_SS	30.43	124.8 *	23.89
3_Diesel_SS	30.62	124.8 *	23.33
1_AGO_SS	30.57	31.20 *	5.158
2_AGO_SS	30.71	31.20 *	5.063
3_AGO_SS	30.84	31.20 *	4.922

NOTES

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:

Figure 5. 3: PFD of vacuum tower

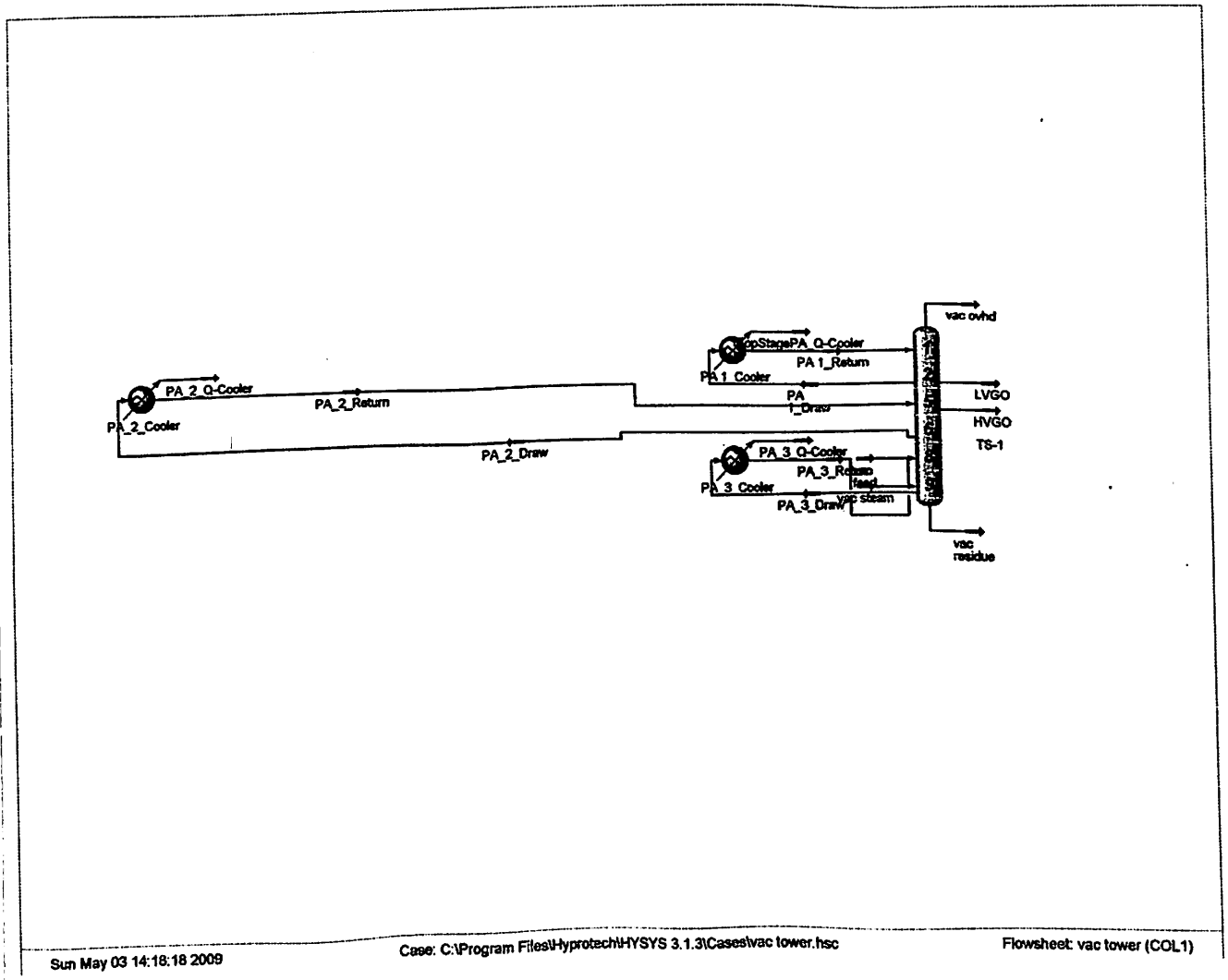

Table 5. 33: connection & monitor of vac tower

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main

CONNECTIONS

Inlet Stream

STREAM NAME	Stage	FROM UNIT OPERATION
vac steam	9__TS-1	
vac feed	8__TS-1	Heater — vac heater

Outlet Stream

STREAM NAME	Stage	TO UNIT OPERATION
vac ovhd	1__TS-1	
vac residue	9__TS-1	
TopStagePA_Q-Cooler	PA 1	
LVGO	2__TS-1	
HVGO	5__TS-1	
PA_2_Q-Cooler	PA_2	
PA_3_Q-Cooler	PA_3	

MONITOR

Specifications Summary

	Specified Value	Current Value	Wt. Error	Wt. Tol.	Abs. Tol.	Active	Estimate	Used
TopStagePA_Rate(Pa)	1211 USGPM *	1211 USGPM	8.943e-005	1.000e-002 *	4.403 USGPM *	On	On	On
LVGO Rate	140.9 USGPM *	140.9 USGPM	8.943e-005	1.000e-002 *	4.403 USGPM *	On	On	On
HVGO Rate	396.3 USGPM *	396.3 USGPM	2.306e-005	1.000e-002 *	4.403 USGPM *	On	On	On
PA 1_Duty(Pa)	---	-9.152e+006 Btu/hr	---	1.000e-002 *	0.9478 Btu/hr *	Off	On	Off
Draw Rate	5732 lb/hr *	5732 lb/hr	-6.637e-005	1.000e-002 *	2.205 lb/hr *	On	On	On
PA_2_Rate(Pa)	1167 USGPM *	1167 USGPM	-4.260e-006	1.000e-002 *	4.403 USGPM *	On	On	On
PA_2_Duty(Pa)	-3.000e+007 Btu/hr *	-3.000e+007 Btu/hr	0.0000	1.000e-002 *	0.9478 Btu/hr *	On	On	On
PA_3_Rate(Pa)	1453 USGPM *	1453 USGPM	-1.485e-005	1.000e-002 *	4.403 USGPM *	On	On	On
PA_3_Duty(Pa)	-1.137e+007 Btu/hr *	-1.137e+007 Btu/hr	0.0000	1.000e-002 *	0.9478 Btu/hr *	On	On	On

SPECS

Column Specification Parameters:

TopStagePA_Rate(Pa)

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Flow Rate	Pumparound:	PA 1	Flow Basis:	Volume		

LVGO Rate

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stream:	LVGO	Flow Basis:	Volume				

HVGO Rate

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stream:	HVGO	Flow Basis:	Volume				

PA 1_Duty(Pa)

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Duty	Pumparound:	PA 1				

Draw Rate

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Stream:	vac ovhd	Flow Basis:	Mass				

PA_2_Rate(Pa)

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---

Licensed to: TEAM LND

* Specified by user.

Table 5. 34: profiles of vac tower

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

Column Specification Parameters

PA_2_Rate(Pa)

Spec Type:	Flow Rate	Pumparound:	PA_2	Flow Basis:	Volume		

PA_2_Duty(Pa)

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Duty	Pumparound:	PA_2				

PA_3_Rate(Pa)

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Flow Rate	Pumparound:	PA_3	Flow Basis:	Volume		

PA_3_Duty(Pa)

Fixed / Ranged:	Fixed	Primary / Alternate:	Primary	Lower Bound:	---	Upper Bound:	---
Spec Type:	Duty	Pumparound:	PA_3				

SUBCOOLING

Degrees of Subcooling		
Subcool to		

User Variables

PROFILES

General Parameters

Sub-Flow Sheet:	vac tower (COL1)	Number of Stages:	9 *

Profile Estimates

	Temperature (F)	Net Liquid (lbmole/hr)	Net Vapour (lbmole/hr)
1_TS-1	250.0 *	2104	241.6
2_TS-1	256.8	0.0000	250.3
3_TS-1	466.0	277.0	381.2
4_TS-1	542.4	261.4	762.4
5_TS-1	585.5	1447	746.8
6_TS-1	623.7	41.57	1064
7_TS-1	708.2	22.32	1003
8_TS-1	717.2	2386	983.6
9_TS-1	700.0 *	853.4	380.8

EFFICIENCIES

Stage Efficiencies

Stages	Overall Efficiency	Methane	Ethane	Propane	I-Butane	n-Butane
1_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
2_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
3_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
4_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
5_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
6_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
7_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
8_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
9_TS-1	1.000	1.000	1.000	1.000	1.000	1.000

Stages	Overall Efficiency	H2O	NBP[0]49*	NBP[0]79*	NBP[0]111*	NBP[0]144*
1_TS-1	1.000	1.000	1.000	1.000	1.000	1.000

Table 5. 35: operation of vac tower

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

Stage Efficiencies

Stages	Overall Efficiency	NBP[0]830*	NBP[0]888*	NBP[0]947*	NBP[0]1009*	NBP[0]1062*
4_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
5_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
6_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
7_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
8_TS-1	1.000	1.000	1.000	1.000	1.000	1.000
9_TS-1	1.000	1.000	1.000	1.000	1.000	1.000

Stages	Overall Efficiency	NBP[0]1124*
1_TS-1	1.000	1.000
2_TS-1	1.000	1.000
3_TS-1	1.000	1.000
4_TS-1	1.000	1.000
5_TS-1	1.000	1.000
6_TS-1	1.000	1.000
7_TS-1	1.000	1.000
8_TS-1	1.000	1.000
9_TS-1	1.000	1.000

SOLVER

Column Solving Algorithm: Modified HYSIM Inside-Out

Solving Options		Acceleration Parameters	
Maximum Iterations:	1000 *	Accelerate K Value & H Model Parameters:	Off
Equilibrium Error Tolerance:	1.000e-05		
Heat/Spec Error Tolerance:	5.000e-004		
Save Solutions as Initial Estimate:	On		
Super Critical Handling Model:	Simple K		
Trace Level:	Low		
Init from Ideal K's:	Off	Damping Parameters	
Initial Estimate Generator Parameters		Azeotrope Check:	Off
Iterative IEG (Good for Chemicals):	Off	Fixed Damping Factor:	1

SIDE STRIPPERS

SIDE RECTIFIERS

PUMP AROUNDS

Pump Around Summary

	Draw Stage	Return Stage	Product Flow (lbmole/hr)	Condenser Duty (Btu/hr)
PA 1	2_TS-1	1_TS-1	2095	-9.152e+006
PA 2	6_TS-1	5_TS-1	1344	-3.000e+007
PA 3	9_TS-1	8_TS-1	1372	-1.137e+007

VAP BYPASSES

RATING

Tray Sections

Tray Section		1_TS-1			
Tray Diameter	(ft)	4.921	*		
Weir Height	(ft)	0.1640	*		

* Specified by user.

Table 5. 36: properties of streams of vac tower

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

Weir Length	(ft)	3.937 *				
Tray Space	(ft)	1.640 *				
Tray Volume	(ft3)	31.20				
Disable Heat Loss Calculations		No				
Heat Model		None				
Rating Calculations		No				
Tray Hold Up	(ft3)	3.120				

Vessels

Vessel					
Diameter					
Length					
Volume					
Orientation					
Vessel has a Boot					
Boot Diameter					
Boot Length					
Hold Up					

Other Equipment in Column Flowsheet

PA 1_Cooler	PA_2_Cooler	PA_3_Cooler	

Pressure Profile

	Pressure (psia)	Pressure Drop (psi)
1_TS-1	1.600 psia *	2.500e-002 psi
2_TS-1	1.625 psia	2.500e-002 psi
3_TS-1	1.650 psia	2.500e-002 psi
4_TS-1	1.675 psia	2.500e-002 psi
5_TS-1	1.700 psia	2.500e-002 psi
6_TS-1	1.725 psia	2.500e-002 psi
7_TS-1	1.750 psia	2.500e-002 psi
8_TS-1	1.775 psia	2.500e-002 psi
9_TS-1	1.800 psia *	—

Pressure Solving Options

Pressure Tolerance	1.000e-004 *	Pressure Drop Tolerance	1.000e-004 *	Damping Factor	1.000 *	Max Press Iterations	100 *

PROPERTIES

Properties: vac feed

		Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction		0.4264	0.4264	0.5736		
Temperature:	(F)	743.0	743.0	743.0		
Pressure:	(psia)	2.178	2.178	2.178		
Molar Flow	(lbmole/hr)	1594	679.9	914.5		
Mass Flow	(lb/hr)	6.981e+005	2.321e+005	4.660e+005		
Std Ideal Liq Vol Flow	(USGPM)	1445	496.2	948.7		
Molar Enthalpy	(Btu/lbmole)	-2.192e+005	-1.515e+005	-2.695e+005		
Mass Enthalpy	(Btu/lb)	-500.6	-443.7	-528.9		
Molar Entropy	(Btu/lbmole-F)	385.7	313.0	439.7		
Mass Entropy	(Btu/lb-F)	0.8807	0.9167	0.8829		
Heat Flow	(Btu/hr)	-3.495e+008	-1.030e+008	-2.465e+008		
Molar Density	(lbmole/ft3)	3.988e-004	1.705e-004	8.735e-002		
Mass Density	(lb/ft3)	0.1746	5.822e-002	44.51		
Std Ideal Liq Mass Density	(lb/ft3)	60.24	58.33	61.24		

* Specified by user.

Table 5. 37: properties of streams of vac tower

HYPROTECH LIFE CYCLE INNOVATIONS

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

Properties : vac feed

		Overall	Vapour Phase	Liquid Phase		
Liq Mass Density @Std Cond	(lb/ft3)	60.17	57.87	61.92		
Molar Heat Capacity	(Btu/lbmole-F)	313.8	228.4	377.3		
Mass Heat Capacity	(Btu/lb-F)	0.7167	0.6690	0.7404		
Thermal Conductivity	(Btu/hr-ft-F)	—	1.564e-002	6.634e-002		
Viscosity	(cP)	—	7.234e-003	0.2704		
Surface Tension	(dyne/cm)	—	—	12.43		
Molecular Weight		437.9	341.4	509.6		
Z Factor		—	0.9886	1.930e-003		

Properties : vac steam

		Overall	Vapour Phase			
Vapour/Phase Fraction		1.0000	1.0000			
Temperature:	(F)	302.0	302.0			
Pressure:	(psia)	21.00	21.00			
Molar Flow	(lbmole/hr)	220.3	220.3			
Mass Flow	(lb/hr)	3968	3968			
Std Ideal Liq Vol Flow	(USGPM)	7.941	7.941			
Molar Enthalpy	(Btu/lbmole)	-1.018e+005	-1.018e+005			
Mass Enthalpy	(Btu/lb)	-5652	-5652			
Molar Entropy	(Btu/lbmole-F)	43.59	43.59			
Mass Entropy	(Btu/lb-F)	2.419	2.419			
Heat Flow	(Btu/hr)	-2.243e+007	-2.243e+007			
Molar Density	(lbmole/ft3)	2.592e-003	2.592e-003			
Mass Density	(lb/ft3)	4.670e-002	4.670e-002			
Std Ideal Liq Mass Density	(lb/ft3)	62.30	62.30			
Liq Mass Density @Std Cond	(lb/ft3)	63.33	63.33			
Molar Heat Capacity	(Btu/lbmole-F)	8.317	8.317			
Mass Heat Capacity	(Btu/lb-F)	0.4617	0.4617			
Thermal Conductivity	(Btu/hr-ft-F)	1.664e-002	1.664e-002			
Viscosity	(cP)	1.407e-002	1.407e-002			
Surface Tension	(dyne/cm)	—	—			
Molecular Weight		18.02	18.02			
Z Factor		0.9912	0.9912			

Properties : vac residue

		Overall	Vapour Phase	Liquid Phase		
Vapour/Phase Fraction		0.0000	0.0000	1.0000		
Temperature:	(F)	709.7	709.7	709.7		
Pressure:	(psia)	1.800	1.800	1.800		
Molar Flow	(lbmole/hr)	853.4	0.0000	853.4		
Mass Flow	(lb/hr)	4.450e+005	0.0000	4.450e+005		
Std Ideal Liq Vol Flow	(USGPM)	903.5	0.0000	903.5		
Molar Enthalpy	(Btu/lbmole)	-2.885e+005	-1.338e+005	-2.885e+005		
Mass Enthalpy	(Btu/lb)	-553.3	-753.2	-553.3		
Molar Entropy	(Btu/lbmole-F)	439.9	182.3	439.9		
Mass Entropy	(Btu/lb-F)	0.8437	1.026	0.8437		
Heat Flow	(Btu/hr)	-2.462e+008	0.0000	-2.462e+008		
Molar Density	(lbmole/ft3)	8.767e-002	1.438e-004	8.767e-002		
Mass Density	(lb/ft3)	46.72	2.555e-002	46.72		
Std Ideal Liq Mass Density	(lb/ft3)	61.41	59.41	61.41		
Liq Mass Density @Std Cond	(lb/ft3)	62.22	58.74	62.22		
Molar Heat Capacity	(Btu/lbmole-F)	379.1	115.1	379.1		
Mass Heat Capacity	(Btu/lb-F)	0.7269	0.8478	0.7269		
Thermal Conductivity	(Btu/hr-ft-F)	6.935e-002	1.904e-002	6.935e-002		

* Specified by user.

Table 5. 38: properties of streams of vac tower

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

Properties : vac residue

		Overall	Vapour Phase	Liquid Phase
Viscosity	(cP)	0.3440	1.296e-002	0.3440
Surface Tension	(dyne/cm)	13.64	—	13.64
Molecular Weight		521.5	177.7	521.5
Z Factor		1.636e-003	0.9972	1.638e-003

Properties : LVGO

		Overall	Vapour Phase	Liquid Phase
Vapour/Phase Fraction		0.0000	0.0000	1.0000
Temperature:	(F)	256.8	256.8	256.8
Pressure:	(psia)	1.625	1.625	1.625
Molar Flow	(lbmole/hr)	243.8	0.0000	243.8
Mass Flow	(lb/hr)	6.325e+004	0.0000	6.325e+004
Std Ideal Liq Vol Flow	(USGPM)	140.9	0.0000	140.9
Molar Enthalpy	(Btu/lbmole)	-2.193e+005	-1.027e+005	-2.193e+005
Mass Enthalpy	(Btu/lb)	-845.3	-3587	-845.3
Molar Entropy	(Btu/lbmole-F)	121.5	52.14	121.5
Mass Entropy	(Btu/lb-F)	0.4682	1.820	0.4682
Heat Flow	(Btu/hr)	-5.347e+007	0.0000	-5.347e+007
Molar Density	(lbmole/ft3)	0.1984	2.115e-004	0.1984
Mass Density	(lb/ft3)	50.97	6.059e-003	50.97
Std Ideal Liq Mass Density	(lb/ft3)	55.97	57.52	55.97
Liq Mass Density @Std Cond	(lb/ft3)	55.94	61.01	55.94
Molar Heat Capacity	(Btu/lbmole-F)	136.5	12.85	136.5
Mass Heat Capacity	(Btu/lb-F)	0.5261	0.4487	0.5261
Thermal Conductivity	(Btu/hr-ft-F)	8.237e-002	1.418e-002	8.237e-002
Viscosity	(cP)	1.099	1.109e-002	1.099
Surface Tension	(dyne/cm)	23.95	—	23.95
Molecular Weight		259.5	28.64	259.5
Z Factor		1.076e-003	0.9991	1.076e-003

Properties : HVGO

		Overall	Vapour Phase	Liquid Phase
Vapour/Phase Fraction		0.0000	0.0000	1.0000
Temperature:	(F)	585.5	585.5	585.5
Pressure:	(psia)	1.700	1.700	1.700
Molar Flow	(lbmole/hr)	475.9	0.0000	475.9
Mass Flow	(lb/hr)	1.881e+005	0.0000	1.881e+005
Std Ideal Liq Vol Flow	(USGPM)	396.3	0.0000	396.3
Molar Enthalpy	(Btu/lbmole)	-2.538e+005	-1.422e+005	-2.538e+005
Mass Enthalpy	(Btu/lb)	-642.1	-671.5	-642.1
Molar Entropy	(Btu/lbmole-F)	292.0	185.3	292.0
Mass Entropy	(Btu/lb-F)	0.7387	0.8748	0.7387
Heat Flow	(Btu/hr)	-1.208e+008	0.0000	-1.208e+008
Molar Density	(lbmole/ft3)	0.1157	1.524e-004	0.1157
Mass Density	(lb/ft3)	45.72	3.227e-002	45.72
Std Ideal Liq Mass Density	(lb/ft3)	59.19	57.20	59.19
Liq Mass Density @Std Cond	(lb/ft3)	58.97	56.72	58.97
Molar Heat Capacity	(Btu/lbmole-F)	268.6	127.2	268.6
Mass Heat Capacity	(Btu/lb-F)	0.6796	0.6007	0.6796
Thermal Conductivity	(Btu/hr-ft-F)	6.960e-002	1.504e-002	6.980e-002
Viscosity	(cP)	0.1932	9.000e-003	0.1932
Surface Tension	(dyne/cm)	15.06	—	15.06
Molecular Weight		395.3	211.8	395.3
Z Factor		1.310e-003	0.9947	1.310e-003

Table 5. 39: properties of streams of vac tower

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

Properties : vac ovhd

		Overall	Vapour Phase
Vapour/Phase Fraction		1.0000	1.0000
Temperature:	(F)	225.3	225.3
Pressure:	(psia)	1.600	1.600
Molar Flow	(lbmole/hr)	241.6	241.6
Mass Flow	(lb/hr)	5732 *	5732
Std Ideal Liq Vol Flow	(USGPM)	12.15	12.15
Molar Enthalpy	(Btu/lbmole)	-1.025e+005	-1.025e+005
Mass Enthalpy	(Btu/lb)	-4318	-4318
Molar Entropy	(Btu/lbmole-F)	49.84	49.84
Mass Entropy	(Btu/lb-F)	2.101	2.101
Heat Flow	(Btu/hr)	-2.475e+007	-2.475e+007
Molar Density	(lbmole/ft3)	2.179e-004	2.179e-004
Mass Density	(lb/ft3)	5.169e-003	5.169e-003
Std Ideal Liq Mass Density	(lb/ft3)	58.83	58.83
Liq Mass Density @Std Cond	(lb/ft3)	61.77	61.77
Molar Heat Capacity	(Btu/lbmole-F)	10.56	10.56
Mass Heat Capacity	(Btu/lb-F)	0.4450	0.4450
Thermal Conductivity	(Btu/hr-ft-F)	1.370e-002	1.370e-002
Viscosity	(cP)	1.017e-002	1.017e-002
Surface Tension	(dyne/cm)	---	---
Molecular Weight		23.73	23.73
Z Factor		0.9991	0.9991

SUMMARY

Flow Basis: Molar — The composition option is selected

Feed Composition

	vac feed	vac steam
Flow Rate (lbmole/hr)	1.594400e+03	220.2752
	---	---
Methane	0.0000	0.0000
Ethane	0.0000	0.0000
Propane	0.0000	0.0000
i-Butane	0.0000	0.0000
n-Butane	0.0000	0.0000
H2O	0.0067	1.0000
NBP[0]49*	0.0001	0.0000
NBP[0]79*	0.0001	0.0000
NBP[0]111*	0.0001	0.0000
NBP[0]144*	0.0002	0.0000
NBP[0]176*	0.0002	0.0000
NBP[0]208*	0.0003	0.0000
NBP[0]240*	0.0004	0.0000
NBP[0]272*	0.0006	0.0000
NBP[0]304*	0.0008	0.0000
NBP[0]336*	0.0011	0.0000
NBP[0]368*	0.0016	0.0000
NBP[0]400*	0.0022	0.0000
NBP[0]433*	0.0036	0.0000
NBP[0]464*	0.0058	0.0000
NBP[0]496*	0.0085	0.0000
NBP[0]528*	0.0113	0.0000
NBP[0]560*	0.0144	0.0000

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Licensed to: TEAM LND

* Specified by user.

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

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

SUMMARY

	vac feed	vac steam
NBP[0]592*	0.0182	0.0000
NBP[0]624*	0.0221	0.0000
NBP[0]656*	0.0265	0.0000
NBP[0]688*	0.0316	0.0000
NBP[0]720*	0.0366	0.0000
NBP[0]752*	0.0410	0.0000
NBP[0]784*	0.0465	0.0000
NBP[0]830*	0.0959	0.0000
NBP[0]888*	0.0929	0.0000
NBP[0]947*	0.0924	0.0000
NBP[0]1009*	0.1181	0.0000
NBP[0]1062*	0.1614	0.0000
NBP[0]1124*	0.1586	0.0000

Flow Basis: Molar — The composition option is selected

Feed Flows

	vac feed	vac steam
Flow Rate (lbmole/hr)	1.594400e+03	220.2752
---	---	---
Methane (lbmole/hr)	0.0001	0.0000
Ethane (lbmole/hr)	0.0003	0.0000
Propane (lbmole/hr)	0.0088	0.0000
i-Butane (lbmole/hr)	0.0093	0.0000
n-Butane (lbmole/hr)	0.0381	0.0000
H2O (lbmole/hr)	10.6450	220.2752
NBP[0]49* (lbmole/hr)	0.1117	0.0000
NBP[0]79* (lbmole/hr)	0.1657	0.0000
NBP[0]111* (lbmole/hr)	0.2086	0.0000
NBP[0]144* (lbmole/hr)	0.2687	0.0000
NBP[0]176* (lbmole/hr)	0.3615	0.0000
NBP[0]208* (lbmole/hr)	0.5205	0.0000
NBP[0]240* (lbmole/hr)	0.7168	0.0000
NBP[0]272* (lbmole/hr)	0.9808	0.0000
NBP[0]304* (lbmole/hr)	1.3246	0.0000
NBP[0]336* (lbmole/hr)	1.7891	0.0000
NBP[0]368* (lbmole/hr)	2.4806	0.0000
NBP[0]400* (lbmole/hr)	3.5796	0.0000
NBP[0]433* (lbmole/hr)	5.6996	0.0000
NBP[0]464* (lbmole/hr)	9.2951	0.0000
NBP[0]496* (lbmole/hr)	13.5894	0.0000
NBP[0]528* (lbmole/hr)	17.9760	0.0000
NBP[0]560* (lbmole/hr)	22.9605	0.0000
NBP[0]592* (lbmole/hr)	29.0422	0.0000
NBP[0]624* (lbmole/hr)	35.2451	0.0000
NBP[0]656* (lbmole/hr)	42.2106	0.0000
NBP[0]688* (lbmole/hr)	50.3988	0.0000
NBP[0]720* (lbmole/hr)	58.2932	0.0000
NBP[0]752* (lbmole/hr)	65.3380	0.0000
NBP[0]784* (lbmole/hr)	74.2190	0.0000
NBP[0]830* (lbmole/hr)	152.8487	0.0000
NBP[0]888* (lbmole/hr)	148.1192	0.0000
NBP[0]947* (lbmole/hr)	147.3452	0.0000

* Specified by user.

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

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

SUMMARY

	vac feed	vac steam			
NBP[0]1009* (lbmole/hr)	188.3519	0.0000			
NBP[0]1062* (lbmole/hr)	257.3472	0.0000			
NBP[0]1124* (lbmole/hr)	252.9105	0.0000			

Products

Flow Basis: Molar — The composition option is selected

Product Compositions

	vac ovhd	LVGO	HVGO	vac residue	
Flow Rate (lbmole/hr)	241.5725	243.7709	475.9398	853.3920	
	---	---	---	---	
Methane	0.0000	0.0000	0.0000	0.0000	
Ethane	0.0000	0.0000	0.0000	0.0000	
Propane	0.0000	0.0000	0.0000	0.0000	
i-Butane	0.0000	0.0000	0.0000	0.0000	
n-Butane	0.0002	0.0000	0.0000	0.0000	
H2O	0.9538	0.0006	0.0002	0.0003	
NBP[0]49*	0.0005	0.0000	0.0000	0.0000	
NBP[0]79*	0.0007	0.0000	0.0000	0.0000	
NBP[0]111*	0.0008	0.0000	0.0000	0.0000	
NBP[0]144*	0.0011	0.0000	0.0000	0.0000	
NBP[0]176*	0.0014	0.0000	0.0000	0.0000	
NBP[0]208*	0.0020	0.0001	0.0000	0.0000	
NBP[0]240*	0.0027	0.0003	0.0000	0.0000	
NBP[0]272*	0.0035	0.0006	0.0000	0.0000	
NBP[0]304*	0.0042	0.0013	0.0000	0.0000	
NBP[0]336*	0.0047	0.0027	0.0000	0.0000	
NBP[0]368*	0.0048	0.0053	0.0000	0.0000	
NBP[0]400*	0.0047	0.0099	0.0001	0.0000	
NBP[0]433*	0.0043	0.0187	0.0002	0.0000	
NBP[0]464*	0.0038	0.0335	0.0004	0.0000	
NBP[0]496*	0.0029	0.0513	0.0008	0.0000	
NBP[0]528*	0.0018	0.0689	0.0015	0.0000	
NBP[0]560*	0.0010	0.0876	0.0027	0.0001	
NBP[0]592*	0.0006	0.1083	0.0049	0.0002	
NBP[0]624*	0.0003	0.1256	0.0088	0.0004	
NBP[0]656*	0.0001	0.1385	0.0162	0.0009	
NBP[0]688*	0.0000	0.1404	0.0308	0.0017	
NBP[0]720*	0.0000	0.1150	0.0575	0.0034	
NBP[0]752*	0.0000	0.0632	0.0937	0.0062	
NBP[0]784*	0.0000	0.0226	0.1242	0.0113	
NBP[0]830*	0.0000	0.0053	0.2440	0.0415	
NBP[0]888*	0.0000	0.0002	0.1795	0.0734	
NBP[0]947*	0.0000	0.0000	0.1112	0.1107	
NBP[0]1009*	0.0000	0.0000	0.0689	0.1823	
NBP[0]1062*	0.0000	0.0000	0.0421	0.2781	
NBP[0]1124*	0.0000	0.0000	0.0124	0.2894	

Flow Basis: Molar — The composition option is selected

Product Flows

	vac ovhd	LVGO	HVGO	vac residue	
Flow Rate (lbmole/hr)	241.5725 *	243.7709 *	475.9398 *	853.3920 *	
	---	---	---	---	
Methane (lbmole/hr)	0.0001 *	0.0000 *	0.0000 *	0.0000 *	

* Specified by user.

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

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

SUMMARY

	vac ovhd	LVGO	HVGO	vac residue
Ethane (lbmole/hr)	0.0003 *	0.0000 *	0.0000 *	0.0000 *
Propane (lbmole/hr)	0.0088 *	0.0000 *	0.0000 *	0.0000 *
i-Butane (lbmole/hr)	0.0092 *	0.0000 *	0.0000 *	0.0000 *
n-Butane (lbmole/hr)	0.0379 *	0.0002 *	0.0000 *	0.0000 *
H2O (lbmole/hr)	230.4175 *	0.1582 *	0.0737 *	0.2707 *
NBP[0]49* (lbmole/hr)	0.1108 *	0.0008 *	0.0001 *	0.0000 *
NBP[0]79* (lbmole/hr)	0.1639 *	0.0016 *	0.0002 *	0.0000 *
NBP[0]111* (lbmole/hr)	0.2053 *	0.0030 *	0.0003 *	0.0000 *
NBP[0]144* (lbmole/hr)	0.2625 *	0.0058 *	0.0005 *	0.0000 *
NBP[0]176* (lbmole/hr)	0.3490 *	0.0118 *	0.0007 *	0.0000 *
NBP[0]208* (lbmole/hr)	0.4917 *	0.0274 *	0.0013 *	0.0000 *
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 *
NBP[0]304* (lbmole/hr)	1.0067 *	0.3111 *	0.0067 *	0.0001 *
NBP[0]336* (lbmole/hr)	1.1257 *	0.6515 *	0.0116 *	0.0003 *
NBP[0]368* (lbmole/hr)	1.1698 *	1.2892 *	0.0210 *	0.0006 *
NBP[0]400* (lbmole/hr)	1.1277 *	2.4105 *	0.0402 *	0.0013 *
NBP[0]433* (lbmole/hr)	1.0483 *	4.5618 *	0.0863 *	0.0032 *
NBP[0]464* (lbmole/hr)	0.9295 *	8.1670 *	0.1905 *	0.0081 *
NBP[0]496* (lbmole/hr)	0.6895 *	12.4997 *	0.3815 *	0.0187 *
NBP[0]528* (lbmole/hr)	0.4352 *	16.7987 *	0.7021 *	0.0400 *
NBP[0]560* (lbmole/hr)	0.2490 *	21.3530 *	1.2743 *	0.0843 *
NBP[0]592* (lbmole/hr)	0.1347 *	26.4004 *	2.3293 *	0.1778 *
NBP[0]624* (lbmole/hr)	0.0660 *	30.6290 *	4.1867 *	0.3634 *
NBP[0]656* (lbmole/hr)	0.0295 *	33.7507 *	7.6919 *	0.7384 *
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 *
NBP[0]752* (lbmole/hr)	0.0007 *	15.4181 *	44.5938 *	5.3254 *
NBP[0]784* (lbmole/hr)	0.0001 *	5.5167 *	59.0902 *	9.6120 *
NBP[0]830* (lbmole/hr)	0.0000 *	1.2886 *	116.1238 *	35.4363 *
NBP[0]888* (lbmole/hr)	0.0000 *	0.0458 *	85.4415 *	62.6320 *
NBP[0]947* (lbmole/hr)	0.0000 *	0.0011 *	52.9088 *	94.4353 *
NBP[0]1009* (lbmole/hr)	0.0000 *	0.0000 *	32.8010 *	155.5508 *
NBP[0]1062* (lbmole/hr)	0.0000 *	0.0000 *	20.0321 *	237.3151 *
NBP[0]1124* (lbmole/hr)	0.0000 *	0.0000 *	5.9110 *	246.9994 *

Flow Basis: Molar — The composition option is selected

Product Recoveries

	vac ovhd	LVGO	HVGO	vac residue
Flow Rate (lbmole/hr)	241.5725	243.7709	475.9398	853.3920
	---	---	---	---
Methane (%)	99.9483	0.0303	0.0212	0.0001
Ethane (%)	99.8721	0.0918	0.0359	0.0002
Propane (%)	99.7342	0.2117	0.0538	0.0003
i-Butane (%)	99.5311	0.3947	0.0737	0.0006
n-Butane (%)	99.4260	0.4922	0.0811	0.0007
H2O (%)	99.7823	0.0685	0.0319	0.1172
NBP[0]49* (%)	99.1980	0.6965	0.1045	0.0010
NBP[0]79* (%)	98.9162	0.9626	0.1200	0.0012
NBP[0]111* (%)	98.4358	1.4214	0.1412	0.0016
NBP[0]144* (%)	97.6686	2.1602	0.1692	0.0020
NBP[0]176* (%)	96.5263	3.2719	0.1992	0.0026

* Specified by user.

Table 5. 43: column profile of vac tower

HYPROTECH LIFECYCLE INNOVATION

TEAM LND
Calgary, Alberta
CANADA

Case Name: C:\Program Files\Hyprotech\HYSYS 3.1.3\Cases\vac tower.hsc
Unit Set: Field-USGPM
Date/Time: Sun May 03 14:03:42 2009

Column Sub-Flowsheet: vac tower @Main (continued)

SUMMARY

	vac dmd	LVGO	HVGO	vac residue	
NBP[0]208* (%)	94.4803	5.2688	0.2473	0.0036	
NBP[0]240* (%)	90.9962	8.6885	0.3103	0.0051	
NBP[0]272* (%)	85.1770	14.4229	0.3928	0.0073	
NBP[0]304* (%)	76.0001	23.4873	0.5021	0.0105	
NBP[0]336* (%)	62.9205	36.4147	0.6493	0.0155	
NBP[0]368* (%)	47.1574	51.9708	0.8485	0.0233	
NBP[0]400* (%)	31.5023	67.3393	1.1230	0.0355	
NBP[0]432* (%)	18.3926	80.0379	1.5142	0.0553	
NBP[0]464* (%)	10.0004	87.8634	2.0497	0.0866	
NBP[0]496* (%)	5.0738	91.9813	2.8073	0.1376	
NBP[0]528* (%)	2.4207	93.4509	3.9060	0.2223	
NBP[0]560* (%)	1.0847	92.9985	5.5498	0.3671	
NBP[0]592* (%)	0.4639	90.9036	8.0204	0.6122	
NBP[0]624* (%)	0.1873	86.9028	11.8787	1.0312	
NBP[0]656* (%)	0.0700	79.9580	18.2226	1.7494	
NBP[0]688* (%)	0.0233	67.9271	29.0902	2.9594	
NBP[0]720* (%)	0.0063	48.0861	46.9546	4.9530	
NBP[0]752* (%)	0.0011	23.5974	68.2510	8.1505	
NBP[0]784* (%)	0.0001	7.4331	79.8160	12.9508	
NBP[0]800* (%)	0.0000	0.8430	75.9730	23.1839	
NBP[0]888* (%)	0.0000	0.0309	57.6842	42.2849	
NBP[0]947* (%)	0.0000	0.0008	35.9080	64.0912	
NBP[0]1009* (%)	0.0000	0.0000	17.4148	82.5852	
NBP[0]1062* (%)	0.0000	0.0000	7.7841	92.2159	
NBP[0]1124* (%)	0.0000	0.0000	2.3372	97.6628	

COLUMN PROFILES

Reflux Ratio: 8.708 Reboil Ratio: 2.916 The Flows Option is Selected Flow Basis: Molar

Column Profiles: Flows

	Temperature (F)	Pressure (psia)	Net Liq (barrels/day)	Net Vap (barrels/day)	Net Feed (barrels/day)	Net Draws (barrels/day)
1 TS-1	225.3	1.600	2104	—	2095	241.6
2 TS-1	256.8	1.625	0.0000	250.3	—	2339
3 TS-1	466.0	1.650	277.0	381.2	—	—
4 TS-1	542.4	1.675	261.4	762.4	—	—
5 TS-1	585.5	1.700	1447	746.8	1344	475.9
6 TS-1	623.7	1.725	41.57	1084	—	1344
7 TS-1	708.2	1.750	22.32	1003	—	—
8 TS-1	717.2	1.775	2386	983.6	2967	—
9 TS-1	709.7	1.800	—	380.8	220.3	2226

Column Profiles: Energy

	Temperature (F)	Liquid Enthalpy (Btu/lbmole)	Vapour Enthalpy (Btu/lbmole)	Heat Loss (Btu/hr)
1 TS-1	225.3	-2.232e+005	-1.025e+005	—
2 TS-1	256.8	-2.193e+005	-1.027e+005	—
3 TS-1	466.0	-2.260e+005	-1.213e+005	—
4 TS-1	542.4	-2.294e+005	-1.428e+005	—
5 TS-1	585.5	-2.538e+005	-1.422e+005	—
6 TS-1	623.7	-2.552e+005	-1.515e+005	—
7 TS-1	708.2	-2.788e+005	-1.472e+005	—
8 TS-1	717.2	-2.811e+005	-1.458e+005	—
9 TS-1	709.7	-2.885e+005	-1.338e+005	—

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* Specified by user.

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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