COMPUTER SIMULATION OF A MULTICOMPONENT, MULTISTAGE CRUDE DISTILLATION PROCESS

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ABSTRACT

A computer program was created to solve the material and energy balance equations simultaneously for a crude distillation column. The model is designed to handle hypothetical hydrocarbon components, allowing for the simulation of crude oils which are usually made up of unknown components. The accuracy of the model depends to some extent upon the accuracy of the crude characterization and the number and size of fractions used in that characterization.

The model incorporates a second liquid phase, water, if it exists and allows for solution of sidestream strippers and pumparounds which are almost always used in crude distillation. The solution technique is a modified Newton-Raphson technique using the method of Newman for inverting and solving the Jacobian.

The model has been used to simulate crude distillation columns for a variety of crude oils and compared to existing operating data available on crude distillation columns. The model was found to be successful when comparing liquid rate and temperature profiles to those obtained from test run data. Improvements in storage and convergence techniques will allow use of this model for solution of crude columns with microcomputers.

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NOMENCLATURE

B	Bottoms flow rate
B_{ij}	Defined Component Matrix Parameter
C_{ij}	Defined Component Matrix Parameter
D	Distillate flow rate
F_j	Total feed rate to tray j
f_j	Component feed rate to tray j
H	Enthalpy of stream
h_i	Enthalpy of component
i	Reference for a component
j	Reference for a tray
K_{ij}	K-value of component i on tray j
L_j	Liquid leaving tray j
l_j	Component liquid rate on tray j
P	System pressure
P_i	Partial pressure of component i
p_i	Vapor pressure of component i
SL_j	Liquid side draw from tray j
SV_j	Vapor side draw from tray j
T	Temperature
x_i	Mole fraction of liquid component i
V_j	Vapor leaving tray j
v_j	Component vapor rate leaving tray j
y_i	Mole fraction of vapor component i

CHAPTER I

INTRODUTION

Distillation is a separation process in which a mixture is separated into two or more fractions. The separation occurs due to a difference in the boiling points of the fractions. The simplest separation process is a flash separator. In a flash separator a mixture is separated into a liquid and a vapor. A distillation column can be thought of as several flash separators in series in which the liquid from one flash is fed to the flash below and the vapor from below flows to the flash above. Each flash separator can be thought of as an equilibrium stage. Each stage is represented by a tray in the distillation column. Trays are designed so that the liquid and vapor are continuously in contact to increase separation.

Two basic types of distillation columns exist- batch and continuous distillation. Batch distillation involves distilling a fixed amount of material. The modeling of batch distillation must take into account the unsteady state behavior as the composition of the mixture and amount of material change with time. Continuous distillation is called such because feeds and products are introduced and removed continuously. In this manner the column exhibits steady state behavior, the composition and flow rates on a

given tray do not change with time. In modeling a continuous distillation column each tray is modeled as an equilibrium stage. This allows use of thermodynamic equilibrium relationships on each tray.

Crude distillation is a specific application of multicomponent, multistage continuous distillation. Crude oil is often separated into different boiling point fractions in distillation columns with steam. The use of steam for stripping often causes water to condense on some trays giving rise to a second liquid phase. Normal distillation models cannot simulate this behavior as they assume only one liquid and one vapor phase. Other elements that are unique to crude distillation columns include sidestream strippers and pumparounds. These extra elements cause considerable problems in conventional models. In the past, the strippers or interconnected columns were solved as separate columns treating the interconnected streams as recycle problems. In the open literature, no model exists that incorporates these three elements all in one program. Models exist to simulate either three phase distillation or strippers but not both. Most models do not include a hypothetical component correlation either.

The present model successfully incorporates the second liquid phase, water, if it exists. The program also allows solution of sidestream strippers and pumparounds without having to solve independently of the main column. Therefore the simulator becomes a powerful crude distillation simulator successfully incorporating the three main elements

of crude distillation all in the same program. The program successfully models crude columns with accurate results.

The simulation has been developed on a VAX mainframe system using programing language Fortran 77. However, it can easily be incoporated to any computer system with a few minor changes. The program is an interactive package containing a large component database. The progam also has the capacity to generate data for hypothetical fractions using a minimal amount of input data.

CHAPTER II

LITERATURE SEARCH

Previous literature has presented the fundamentals of multicomponent, multistage continuous distillation. Many different algorithms have been employed to model basic distillation columns. However, relatively few can handle the problems associated with crude distillation columns. The intent of this chapter is to present an introduction to basic multistage, multicomponent distillation, to look at available computer simulations of distillation columns and their convergence techniques, and to examine past and current work in modeling three phase and/or crude distillation problems.

Introduction

Crude oil towers represent an important part of any refinery operation. Two basic types of crude towers exist, atmospheric crude towers and vacuum crude towers. Crude fractionation is often the first operations performed on crude when it enters the refinery. Crude towers separate the crude into different boiling fractions so that the fractions may be sent to other units to be further refined. The towers operate at low pressure and use steam to strip the crude. The steam often gives rise to water condensing

on certain trays. If water condenses it can cause considerable problems and it is to the advantage of the designer to know when this can occur and to take measures to prevent or account for it. Other than proprietary methods, no adequate method of simulating crude distillation, combining material and energy balances, sidestream strippers, and pumparounds, currently exists in the open literature.

Distillation Fundamentals

The author assumes that the reader is familar with the basic fundamentals of distillation. Those who are unfamilar with the subject are referred to Seader and Henley's, Equilibrium-Stage Separation Operations in Chemical Engineering. This book presents a comprehensive study of distillation principles. Material and Energy balances are given and explained. Equilibrium relationships are presented and methods of solution used in hand calculations are discussed.

Continuous Distillation

Many techniques are available for solution of continuous distillation columns. The basic equations and solution methods of these techniques are the same as those for three-phase distillation columns with some minor changes.

Amundson and Pontinen (1958) developed one of the first methods for computer solution of a continuous distillation

process. Equations were developed for mass balances on each component and heat balances on each tray. These two sets of equations are expressed in matrix format and solved by matrix inversion. The material balance equations were written in envelopes including the top tray. The material balance for the bottom tray is just around that tray. The Component matrix contains a large number of zero elements and therefore should be easy to invert. Heat balances are written around each tray so that the matrix is tridiagonal and easily solved. The solution method iterates on temperature distribution using Newton's method. First, constant molal overflow calculations are done to give a good starting point for the entire problem. Complete heat balancing is accomplished by using the vapor and temperature profiles from constant molal overflow as initial estimates, solving component mass balances to get new temperature profiles, solving heat balances to get new vapor profiles, and substituting back into the composition matrix and repeating the entire process until the desired accuracy is achieved. The authors presented the method but did not attempt to solve the matrix equations realizing that there were methods available to do this.

Friday and Smith (1964) analyzed the decisions involved in the formulation of a solution method of equilibrium stage equations. The best procedures are dependent upon the type of problem. The authors suggest that the model equations should be grouped by the type of equation, not by stage, and that the concentrations should be solved for first. They

also present a new method for solving the C-matrix which is not susceptible to truncation error, using stripping factors. The authors point out that the bubble point method of convergence should be used on close boiling feeds whereas the sum rates method should be used on wide boiling feeds. The differences in the two methods are based upon the calculations of the vapor and temperature profiles. The bubble point method calculates the vapor profiles from the energy balances and the new temperature profiles are then obtained from bubble point calculations. The sum rates method adds the component rates to get the vapor profiles and calculates the temperatures from the energy balances. The authors felt that the sum rates method represented the most versatile method for a general computer program.

Tomich (1970) described a general method of solution capable of solving either distillation or absorber problems. The component mass balance equations are set up so that the matrix is tridiagonal and a tridiagonal matrix algorithm is used to solve these equations. New temperature and vapor flow rates estimates are obtained by solving component summation equations and heat balances simultaneously. Broyden's method, a modified Newton-Raphson method, is used as it requires only one inverse of the Jacobian matrix per problem by updating the Jacobian at each level using the residuals calculated at the previous level. All equations are solved simultaneously, in a mathematical sense, with the assumption that all data are composition independent. This method is numerically stable for all types of columns since

it avoids unstable tray by tray calculations. However, the method requires that a large number of partial derivatives be calculated numerically and solving the composition profiles independently resticts the range of applications. Tomich was one of the first to solve all equations simultaneously.

Naphtali and Sandholm (1971) presented a new approach to separation calculations which works on a wide variety of problems. The nonlinear equations of mass balances, energy balances, and equilibrium relationships are grouped by stage and solved simultaneously. The equations are linearized and solved by a modified Newton-Raphson procedure. Convergence increases as the solution is approached, as the linearized equations become closer and closer to being exact. When the equations and variables are grouped according to stage, the matrix of partial derivatives is in block tridiagonal form, which is particularly easy to solve. The authors show the proper grouping of equations to create block tridiagonal form and develop the equations needed to calculate the Jacobian matrix. Inversion of the Jacobian is accomplished by a Gaussian elimination scheme. The Newton-Raphson method uses the inverted Jacobian to give the updated values of the variables. Although the authors have tried to decrease the amount of storage space required as much as possible, the method still requires a large amount and is a major disadvantage to this method.

Goldstein and Stanfield (1970) have presented a Newton-Raphson solution technique with improvements in speed and efficiency. They state that one of the keys to the use of

the Newton-Raphson is in finding an economical method of solving the linearized set of equations. They discuss the two methods of ordering the equations. For columns with a large number of trays and few components or composition-dependent data, it is most efficient to group equations by stage. However, for fewer trays and a large number of components, it is more efficient to group the component balances by component. The authors developed the second method of grouping, though the technique is applicable to either method of grouping. The inversion of the Jacobian matrix is carried out by partial triangularization of the matrix. Then Gaussian elimination of matrices is used to invert the reduced matrix instead of elimination of elements. The authors note that the Newton-Raphson technique usually gives good direction toward the solution, though sometimes the magnitude of the correction is too large. They recommend using a fraction of the correction to reduce the error. They also extended the algorithm to allow solution of large towers. The column is sectioned off, then in each section the temperature and loading profiles are linearized. The resulting solution is an approximation, but is very accurate for the product streams. By successive applications of this procedure, the exact solution can be obtained.

Gentry (1970) presents a method for the numerical solution of a stagewise process. All non-linear terms are linearized using Newton's method. The resulting material and energy balances form a block band matrix. A block band

matrix is a matrix in which the band elements are square matrices. The author assumes that each variable, temperature, liquid, composition, is independent of the other variables. This allows for the uncoupling of the equations. This method is very similar to the Amundson-Pontien (1958) method. This method of decoupling the equations is probably not desirable for equations as coupled as these.

A method for solving block band matrices has been developed by Gentry (1970). The matrix is first rewritten by arranging the blocks in a column matrix so that the first non-zero entry appears in the first column. This saves an enormous amount of storage space for large matrices. The method is essentially Gaussian elimination with partial pivoting. After elimination, solution is accomplished by back substitution. Many methods using Gaussian elimination fail where Gentry's method does not because partial pivoting of the blocks is required. The success of this method of solving block band matrices lies in its use of partial pivoting.

Newman (1963) has proposed two methods of solution of a distillation process. The first method updates temperatures making distillation calculations by the Thiele-Geddes method. The Thiele-Geddes method requires an assumption of the temperature profiles in advance. Calculations are done using the assumed temperatures. The differences between one and the sum of the mole fractions on any stage are related to the errors in the assumed

temperatures using linear approximation. A linear relationship is used to approximate the relationship between temperature and sums of mole fractions.

Newman (1967) developed a second method for solution of a distillation process which uses simultaneous solution of all equations. The new method requires a large amount of storage space but is preferrable to the previous method. The new method is more general in its application, especially in the flexibility of the specification of column operating conditions. The first program had some problems in updating liquid flows due to enthalpy balances. The new program, however, includes the linearized enthalpy balances and solves them simultaneously with the material balances. The Jacobian is in the form of a block tridiagonal matrix. The Jacobian is inverted by Newman's (1968) method for solving coupled, ordinary differential equations. This method is applicable to many different problems. Its application to a distillation process is possible since the distillation equations are similar to finite difference approximations to differential equations. The distillation equations involve unknown quantities for the stage above and the stage below the one in question. This gives rise to the similarities to finite difference approximations.

The method saves space by solving the Jacobian in pieces. This allows for storage of the error array (E array) only and negates the storage of the entire Jacobian. The program calculates derivatives associated with each tray, including the tray below and tray above. Calls to

Band, the numerical solution subroutine, are made on a tray to tray basis in order to save storage space. Gaussian elimination with partial pivoting is used to invert the Jacobian.

This method is very general and encompasses a variety of end specifications. Its basic disadvantage lies in the vast amount of storage space needed for solution.

Ishii and Otto (1973) describe a method which employs a multivariate Newton procedure to solve all equations simultaneously for corrections to temperatures, flow rates and compositions. The authors note that previous methods based on linearization and simultaneous solution require that a large number of partial derivatives be evaluated, large storage space requirements, and long computational times. These methods have difficulties incorporating composition dependent enthalpy and equilibrium data. To reduce the computational effort, the authors incorporate the assumptions that K values are pressure, temperature, and composition dependent and that enthalpies are only pressure and temperature dependent. Using these assumptions in the linearization process, the linearized equations only include partial derivatives that have a dominant influence on the solution. The authors used analytical techniques to evaluate the partial derivatives since large amounts of storage and calculational times are required for numerical evaluation. The linearized equations then become sums of matrix equations and are solved by an algorithm of the authors or Gauss-Jordan elimination. The method converges

for a broad range of initial assumptions, readily handles complex column configurations, incorporates easily composition dependent enthalpy and equilibrium data, and is numerically stable for a wide range of feeds.

Venkataraman and Lucia (1988) discuss the convergence characteristics of Newton-like solution methods. These methods are only locally convergent and must have good initial profiles to converge for non-ideal problems. The authors discuss the use of line searching, trust regions and relaxation but prefer to improve the method of generating initial profiles. The authors developed a procedure to create initial profiles based on shortcut techniques such as effective equilibrium ratios, Underwood's correlation and constant molal overflow. The result was a successful tray by tray which converged for ideal and non-ideal problems.

Vickery, Ferrari and Taylor (1988) present a homotopy continuation method to assist in convergence of Newton's method. Homotopy continuation is a method in which a sequence of related problems are solved beginning with an easy problem to solve and ending with the more difficult, desired problem solved. The authors chose to use the tray efficiency as the continuation parameter. The method solved all the difficult problems it was tried on without any difficulty. The algorithm is an effective method for solving separation problems that are difficult otherwise and is easily implemented into any existing Newton method of solution.

Kuno (1984) developed two methods to improve Wang and

Henke's (1966) tridiagonal matrix solution method. The methods focus primarily on improving convergence. The first method is called partial normalization. It improves vapor rates by giving priority to either light or heavy components and replacing the normalization equation with partial normalization. The components given priority have a self-convergence nature and converge very quickly. Likewise, the vapor rates of priority components approach a solution rapidly and convergence is improved.

Kuno improved this method to converge for more types of problems. The later method is called the enthalpy method. The enthalpy method uses the priority components of the partial normalization technique. A parameter is added to the enthapies of each of these components. When the priority components are the light components the parameter is of positive value, when they are the heavy components then the parameter is of negative value. The value of the parameter is determined by trial and error. This method, in effect, varies the numerical values of the enthalpies and increases the speed of convergence.

A new class of solution methods for equilibrium processes that have gained attention belong to the class of 'inside-out' methods generated by Boston (1970). The difference in the new method lies in the choice of variables on which the solution is iterated. Previous methods iterated on temperature profiles and interstage flow rates. Boston's (1970) method iterates on defined volatility and energy parameters. These variables were chosen since they

are very weak functions of the temperatues and interstage flow rates. This reduces the dependence of convergence on initial estimates of temperatures and flow rates and converges with exceptional stability and great speed.

Boston and Sullivan (1974) presented a detailed analysis of the procedure with substantial improvements over the previous procedure of Boston (1970). Defining relations for the successive approximation variables, volatility parameters and energy parameters, are given in detail. The volatility parameters are defined so that they compare easily with K values, and so that the parameters which affect the stage temperatures the most are the least sensitive to variations in the stage temperature. The choice of these parameters tends to avoid the difficulties associated with interactions between variables. Corresponding temperatures are obtained from bubble point calculations with the volatility parameters. The correction of the volatility parameters is the outside loop. The inside loop involves using these parameters to solve mass balance and energy balance equations. The quasi-Newton method of Broyden or direct substitution is used in updating parameters. Using either solution technique, the method has been shown to be both stable and efficient.

Boston and Sullivan (1972) have also developed an improved algorithm for the solution of the component mass conservation equations. There are two approaches which can be used to write the mass balance equations. One of these writes the mass conservation equations in envelopes that

include a terminal stream of the column. This approach has been shown to produce round-off error. The second approach encircles only single stages. The system of equations that result are in tridiagonal matrix form. Two methods currently exist to solve tridiagonal matrices. The first, involves forward elimination and back substitution. The only step that produces round-off error is the subtraction step in the forward elimination. The algorithm presented by Friday and Smith (1964) contains no subtraction steps but is cumbersome. The new algorithm is based on the first method. The algorithm eliminates the subtraction step, reducing error propagation with little increase in computational time.

Boston and Britt (1978) presented an article detailing the 'inside-out' technique as applied to the solution of the single-stage flash problem. The authors summarize the major advantages of the 'inside-out' technique. Simple model parameters are used to replace temperature, pressure, and composition eliminating the need for extensive complex equilibrium and enthalpy models. These simple model parameters are relatively insensitive to the accuracy of initial estimates. Because of the choice of iteration variables the initial Jacobian matrix may be approximated by the identity matrix. Subsequent Jacobians are approximated by Broyden's method, eliminating the need to invert the Jacobian even once. The algorithm takes into account different types of systems, initialization, and convergence.

Boston (1980) outlines the 'inside-out' procedure and

its application to different equilibrium stage equations.

He lists the six most important features of this procedure:

(1) Complex K-value and enthalpy models are used only to generate parameters for simple models. These parameters are unique for each stage of a multi-stage system.

(2) These simple model parameters become the main (or "outer loop") iteration variables, the role played by the primitive variables temperature, pressure, and vapor and liquid composition and phase rates in Class I (Sum rates or Bubble point methods) and Class II methods (simultaneous solution methods).

(3) The new outer loop iteration variables are relatively free of interaction with each other, and are relatively independent of the primitive variables, hence precise initialization is not critical to good algorithm performance.

(4) The describing equations are expressed in terms of the simple models, and are rearranged in a novel way so that a complete solution for the primative variables is possible. In most cases this is achieved by converging an N-dimensional inner iteration loop, where N is the number of stages. In multi-stage applications, this inner loop is particularly amenable to solution by Broyden's quasi-Newton method, using the identity matrix as the initial Jacobian.

(5) The inner loop iteration variable for each stage is a unique combination of temperature and phase ratio which eliminates the need to make a distinction between wide- and narrow-boiling systems. In certain cases there are additional inner loop variables.

(6) The primitive variables resulting from the solution of the inner loop for a given set of values of the outer loop variables are used to calculate a new set of the latter through the actual K-value and enthalpy models. The entire problem is solved when these calculated values match the corresponding assumed values. In single-stage applications, convergence of the outer loop is accelerated using the Broyden quasi-Newton method with the identity matrix as the initial Jacobian. In multi-stage applications, it is converged either by direct substitution or by the bounded Wegstein method.

Russell (1983) of Badger Engineers, Inc. developed an improved method belonging to the 'inside-out' class. Russell had noticed that the set of stripping and withdrawal factors in the Boston-Sullivan (1974) method often cause an initial maldistribution of components when initial temperature and vapor flow estimates are poor. To counter these effects and increase speed of convergence, Sullivan used relative stripping factors instead. The base stripping factor remains the same throughout the calculations. The scaling of the stripping factors moves the process material up or down the column readjusting the compositions and puts the starting point of the convergence procedure in the initial vicinity of the solution.

Crude Distillation

One of the first processes in any petroleum refinery is the separation of crude into various fractions by crude distillation. Two types of crude distillation are used in refining petroleum. First the crude is sent to the atmospheric tower in which the crude is fractionated into various fractions at slighlty above atmospheric pressure. The bottoms product from the atmospheric tower, the unvaporized crude, is sent to a vacuum tower. The vacuum tower operates at subatmospheric pressure to further fractionate the remaining crude. Regardless of its importance, the technical literature is lacking in information on crude distillation with even less information being available on the computer solution of crude

distillation columns.

One of the most comprehensive studies of crude distillation is given by R. N. Watkins in Petroleum Refinery Distillation. The book presents methods used in hand calculations of atmospheric and vacuum distillation columns. Watkins shows how to analyze crude and separate it into hypothetical fractions, how to determine the appropriate amounts of steam, and the number of trays between cuts. Complete heat and material balances are given along with detailed examples following step by step the procedure for both types of columns.

Cecchetti, Johnston, Niedzwiecki and Holland (1963) presented one of the first computer programs for crude distillation. The method uses assumed temperature and vapor rate profiles and iterates until these assumed profiles no longer change. The calculations are carried out from the condenser down to the feed plate and from the bottom plate up. Material balances are used on each plate in this method to obtain the compositions. Holland's Theta-method of convergence is used to correct the calculated compositions, once they are merged at the feed plate. The desired set of theta's are determined by the Newton-Raphson method, numerically evaluating the partial derivatives. The next set of temperatures are calculated from the compositions by the bubble or dew point procedure. Sidestream strippers are solved by the same method as the main column. Calculations are done from the bottom of the sidestream stripper up and the entering vapor compositions merged with those on the

entering tray before continuing down the main column. This method converges slowly because of the need to merge the components at the feed tray. Special consideration must be given to separated components, which can cause convergence problems. Water is assumed to be in the vapor phase on all trays except the condenser where two immiscible liquids are assumed.

Hekim, Orrick, and Erbar (1973) presented a paper recommending a procedure for short cut calculations for distillation calculations in three phase systems. These short cut calculations include predicting the minimum number of trays and product distributions. The Fenske and Winn methods are modified to account for the second liquid phase and/or azeotropic behavior. To avoid complexity the model should be based on total liquid stream compostions. The procedure is based on sound mathematical principles. A short cut procedure enables the design engineer to save time in computing distillation columns that are infeasible.

Block and Hegner (1976) presented one of the first simulation models of three-phase distillation. The model incorporates two partly miscible liquid phases and a vapor phase. The model works well for any combination of two- and three-phase stages, defaulting to a two-phase distillation column in the absence of a second liquid phase. A phase splitting parameter is used to describe the phase separation of the liquid at a stage. The method is a modified bubble point method in which the liquid compositions are treated as the independent variables. The procedure is summarized as

follows:

(1) Assume initial set of average liquid compositions.

(2) Phase splitting is computed by iterative procedure and the compositions of each liquid phase are obtained.

(3) Temperature profiles and vapor compositions are computed by a bubble point procedure.

(4) Liquid and vapor flows are calculated.

(5) Residuals of the component mass balance equations are obtained.

(6) New liquid flows are obtained from the Newton-Raphson procedure. Repeat until desired accuracy is achieved.

A damping factor is used in updating the liquid compositions. Convergence is dependent upon the initial estimates of the independent variables. Stability is good except when absorber type columns are simulated.

Hess, Holland, McDaniel and Tetlow (1977) present a Newton-Raphson method to solve crude distillation columns. Water is regarded as being distributed between the vapor and liquid phases on all stages except for the condenser. On the condenser, two liquid phases are presummed with the liquids being immiscible. The composition matrix has off band elements due to the sidestream strippers. The sidestream strippers are numbered as extensions to the column to keep the Jacobian as sparse as possible. The heat and material balance equations are solved by the Newton-Raphson, similar to its use in conventional distillation

columns. The method requires a large amount of storage space and computational time.

Hofeling and Seader (1978) presented a modified Naphtali-Sandholm (1971) method for interlinked columns. The modification retains the technique of total linearization and simultaneous solution by a Newton-Raphson procedure. The equations are grouped by stage in this method so it is important to number the stages so as to minimize the off band elements. The stages are numbered as if the interlinked column is located within the main column. The solution of the composition matrix is modified to account for the off band elements using forward elimination. This method will also incorporate bypass or pumparound streams.

Hidalgo, Correa, Gomez and Seader (1980) present a method which provides for optimal arrangement of the linearized equations for systems of interlinked, multistaged separators. An algorithm was written which would automatically arrange the linearized equations so that a minimum or nearly minimum number of nonzero blocks appear outside the tridiagonal band. An exhaustive search of all ways of writing the equations is not feasible so the computer search had to be limited. The authors found that optimal arrangements occured along only one vector of the matrix. The computer search space was limited to this area. The algorithm was written for use with the Newton-Raphson procedure of Naphtali and Sandholm (1971).

Wasek and Socha (1980) discuss a solution method using

the Newton-Raphson technique for steam strippers. The strippers have no reboiler or condenser. Steam is assumed to be in a vapor state throughout the column. The stripper is solved separately from the main column for a given liquid flow rate to the top of the stripper. Averaging modules are used to make the choice of initial temperature profile and vapor flow rates less critical, and to induce convergence. All equations are solved similtaneously using the Newton-Raphson technique for convergence. The method converged even for slightly negative initial vapor flows.

Niedzwiecki, Springer, and Wolfe (1980) developed a computer program called TPDIST for three phase distillation. The program specifically accounts for liquid water in water-hydrocarbon systems. The program modifies a previous one liquid phase distillation program with a Newton-Raphson convergence scheme. A technique was developed to modify K-values to account for the second liquid phase. This limits the set of equations to be solved. The program provides accurate results for towers containing two liquid phases.

Stadtherr and Malachowski (1982) present a Newton-Raphson method of solution for interlinked columns. The method of solution is similar to that of Hofeling and Seader (1978). The authors note that storage requirements become an important problem in solving complex systems of columns. With quasi-Newton solution methods, storage requirements are increased significantly. However, the Jacobian is evaluated only once and updated thereafter. This type of method is particularly useful for highly nonideal solutions and when

excessive time is consumed in calculating the partial derivatives. The method of Hofeling and Seader requires storage of all elements above the diagonal for back substitution. The storage space requirements represent a major problem. Therefore, it is desirable to minimize the number of off band elements and their distance from the diagonal. The authors describe two alternative procedures to reduce the storage requirements. Both methods use bordered matrix form. The bordered forms are obtained by permuting the columns and rows of the original matrix. Two bordered forms can be utilized, bordered-block-triangular form and bordered-block tridiagonal form. The bordered forms are solved by a combination of block elimination and implicit back substitution in an algorithm developed by the authors. Although the method offers a significant improvement with respect to storage, there is some trade-off with regard to increase in operations.

Kinoshita, Hashimoto and Takamatsu (1983) presented a method to solve multicomponent distillation columns with three phases of a vapor and two partially immiscible liquids. The authors use phase-splitting parameters for derivation of the basic equations. The method can be applied easily to nonideal solutions or reactive solutions. The liquid compositions of the two phases are in a way averaged and the averaged liquid compositions become the independent variables. If the liquid compostions come out to be the same in both phases, then only one liquid exists. No convergence technique is applied, direct substitution is

used, and still the method converges rapidly. The application is quite successful and represents a powerful procedure for three phase distillation problems.

Schuil and Bool (1984) presented a method to incorporate three phases in flash and distillation calculations. The model was incorporated in Russell's (1983) distillation program. A mixed K-value model, an effective average of the two liquid phase K-values, is used without extensive modifications to the existing program. The mixed K-value model works well for systems with distribution of all components over both liquid phases. The model also applies to systems in which the second liquid phase is pure water.

Russell (1983) describes his method for solving crude distillation columns. The method incorporates sidestrippers and pumparounds in his conventional distillation program discussed earlier. Russell numbers the sidestream stripper stages as if they were located on top of the column. The method of Hofeling and Seader (1978), Gaussian elimination, is used to solve the component mass balance equations which are no longer tridiagonal. Convergence for the system is the same as for the single column, with a few extra variables. The method converges easily and quickly for atmospheric and wet vacuum columns.

Ross (1979) presented a method for three-phase distillation using the 'inside-out' technique of Boston. The model incorporates Murphree tray efficiencies and it was noticed that the appearance of a second liquid phase depended upon the tray efficiency. This suggests that

previous methods for simulating three-phase towers which assume equilibrium trays may give misleading results. Computation time is drastically reduced by removing the liquid-liquid equilibrium calculations to the outer loop. The model successfully simulates azeotropic distillation, both wide- and narrow-boiling systems, and highly nonideal systems which require some damping between iterations of the outer loop.

CHAPTER III

CRUDE DISTILLATION PRINCIPLES

Introduction

Crude distillation represents a special case of multi-component, multistage continuous distillation involving several elements peculiar to crude processing. This chapter will look at the elements of crude distillation, the equations which describe general distillation, particularly those elements of crude distillation, and describe the simulation package.

Crude Distillation

One of the first major processing units in any refinery is crude distillation. Crude units separate the crude oil into fractions according to boiling point. Each of the fractions represents a feed to another processing unit and certain specifications for each fraction must be met. Typical boiling point ranges for the various fractions are given in Table I. Following is a description of crude oil distillation borrowed liberally from Petroleum Refining (Gary and Handwerk, 1984). Crude oil separation can be accomplished in one or two stages. A one stage crude unit involves only one distillation column, the atmospheric tower. The atmospheric tower fractionates the entire crude

TABLE I

CRUDE DISTILLATION PRODUCTS

Fraction	TBP(^{o}F)
Butanes and Lighter	
Light straight-run gasoline	90-190
Naphtha (heavy straight-run gasoline)	190-380
Kerosine	380-520
Light gas oil	520-610
Atmospheric gas oil	610-800
Vacuum gas oil	800-1050
Vacuum reduced crude	1050+

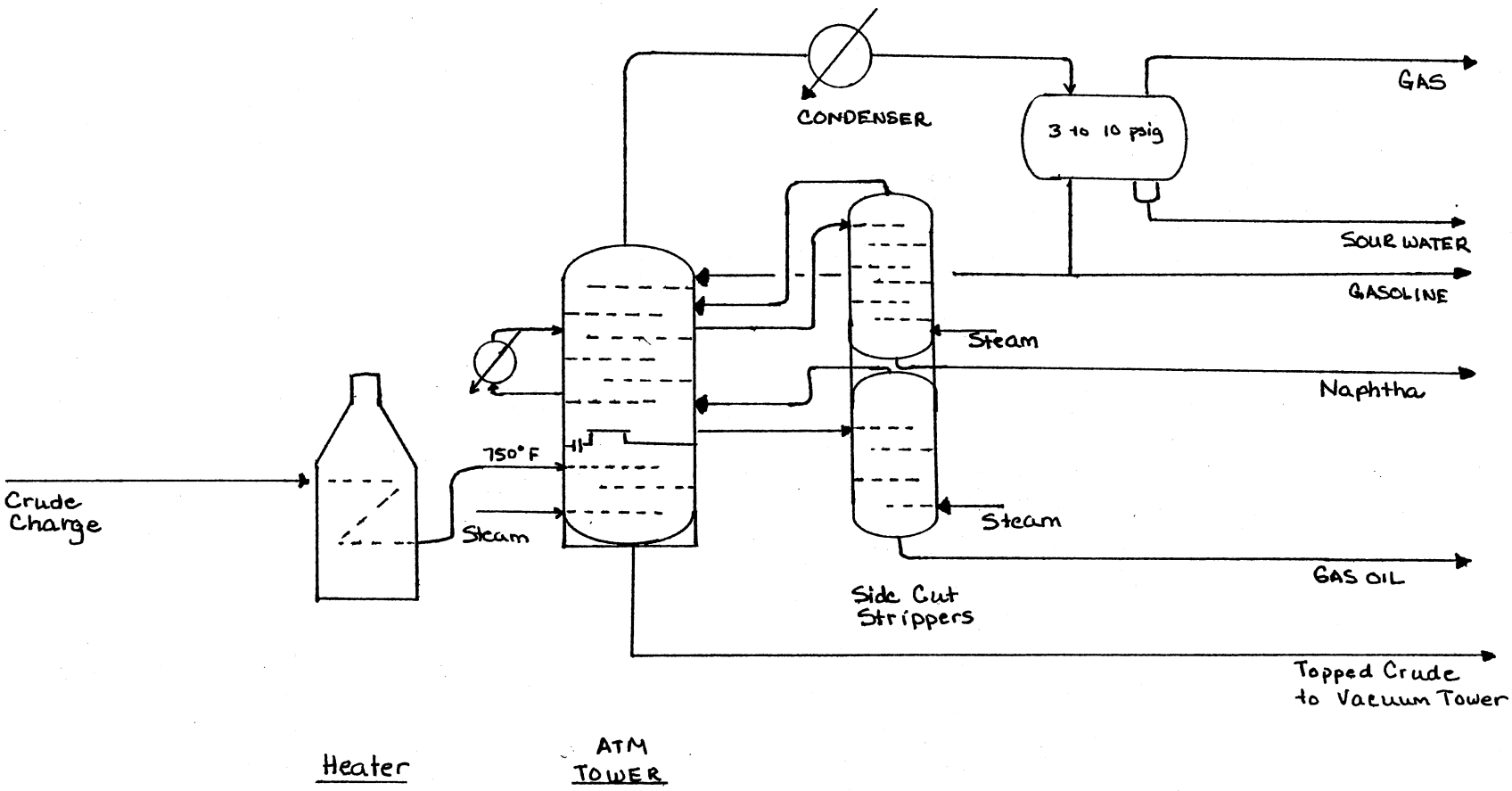
Source: Gary, James H. and Glenn E. Handwerk Petroleum Refining. New York: Marcel Dekker, Inc. (1984) p. 32.

feed at near atmospheric pressure.

Two stage crude units involve two distillation columns, one operating at atmospheric pressure and the other under vacuum. Gary and Handwerk (1984) note, "Higher efficiencies and lower costs are achieved if the crude oil separation is accomplished in two steps: first, by fractionating the total crude oil at essentially atmospheric pressure; then, by feeding the high-boiling bottoms fraction(topped crude) from the atmospheric still to a second fractionator operated at a high vacuum." A typical two stage crude unit is shown in Figures 1 and 2. The vacuum tower further separates the topped crude into fractions. These fractions cannot be separated in the atmospheric column because the high temperatures involved cause thermal cracking to occur, resulting in coke formation and subsequent equipment fouling. By operating under vacuum, the temperatures at which these cuts vaporize is lowered, allowing for separation. The bottoms product from the vacuum tower is called vacuum reduced crude. The vacuum reduced crude can be sent to a deasphalting unit or a coking unit to make low grade fuel.

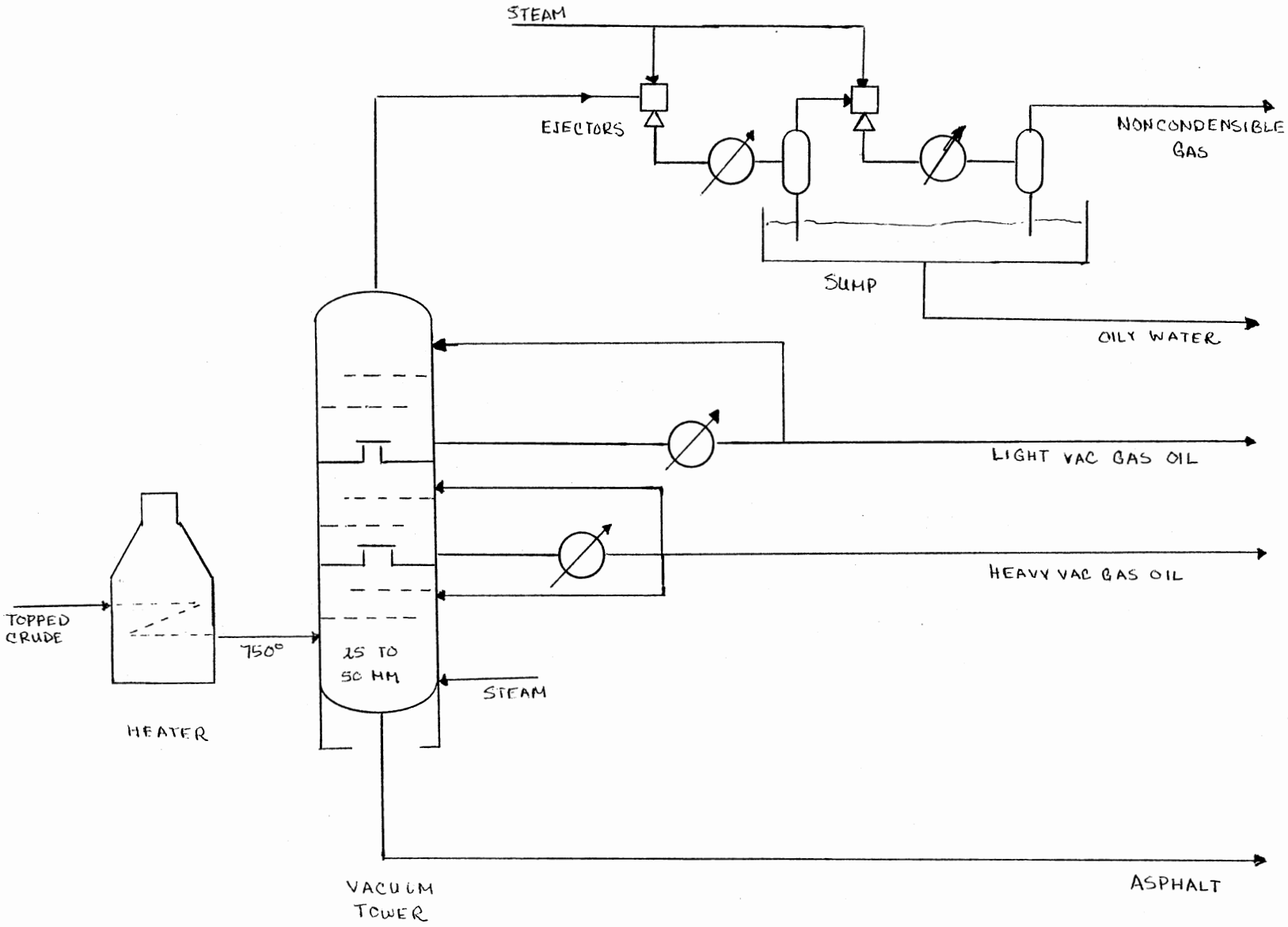
Preprocessing

Before the crude oil is fractionated, it must be desalted in order to minimize fouling and corrosion of equipment. Along with the salt, some metals are partially removed to prevent catalyst poisoning in subsequent catalyst processing units. Desalting can be done in one or two stage



After: Gary, James H. and Glenn E. Handwerk Petroleum Refining. New York: Marcel Dekker, Inc. (1984) p. 41.

Figure 1: Atmospheric Distillation



After: Gary, James H. and Glenn E. Handwerk Petroleum Refining. New York: Marcel Dekker, Inc. (1984) p. 44.

Figure 2: Vacuum Distillation

units depending on the amount of salt in the crude. Desalting is accomplished by mixing the crude oil with water. The salts dissolve in the wash water and the oil and water phases are separated. Phase separation can be accomplished in a settling vessel by adding chemicals to break the emulsion or by inducing an electric field across the vessel. Once desalted, the crude oil is sent through a series of heat exchangers where its temperature is raised to about 550°F. A furnace is used to further heat the crude oil to about 750°F.

Atmospheric Distillation

The crude oil from the furnace is charged to the flash zone of the atmospheric tower. A typical atmospheric tower is shown in Figure 1. The furnace temperature should be hot enough to vaporize all of the side products plus about 20% of the bottoms product. This is called a 20% "over-flash" and helps to increase fractionation by providing an internal reflux and boilup. There are usually a few trays between the flash zone and the introduction of steam below the bottom tray. This is to strip any remaining gas oil from the bottoms product.

Reflux to the tower is provided by condensing part of the overhead vapor and returning it to the top tray, by pumparound streams and by pumpback streams. The liquid sidedraws use a large amount of the liquid flow thereby requiring a large reflux. If all of the reflux is provided at the top tray, a large liquid loading at the top of the

tower results causing a very large diameter at the top of the tower. Also, the liquid would have to be cooled to the bubble point temperature requiring a large amount of low temperature cooling source. By using pumparound or pumpback streams the intermediate cooling can be done at higher temperatures making the column more energy efficient. Therefore, pumparounds and pumpback streams are used to introduce reflux at various points in the column. Pumparound streams involve removing a liquid side stream, cooling it by heat exchange with the feed to the tower, and returning it to the column. Pumpback streams are a portion of a cooled side product stream returned to the column. These cold streams help to condense some of the vapors when returned to the column, increasing the reflux.

The sidestream draws are further stripped in sidestream strippers. These strippers usually contain four to ten trays and are used to strip out any light ends which are then sent back to the tower. The overhead condenser normally condenses any pentane and heavier components and the lighter gases are vented out the top.

Vacuum Distillation

The topped crude from the bottoms of the atmospheric tower is fed to a heater and then to the flash zone of the vacuum tower. A typical vacuum distillation unit is shown in Figure 2. The vacuum tower is designed to minimize the pressure loss from top to bottom thus minimizing the pressure at the flash zone. This pressure determines the

fraction of the feed that is vaporized for a given furnace outlet temperature. The volume of vapor will be greatly increased because of the low pressure, resulting in very large column diameters. The elements of vacuum distillation are very similar to those for an atmospheric unit. The vacuum is maintained by the use of steam ejectors or a vacuum pump. Typical vacuum column pressures are in the range of 10-200 mm Hg.

Background

Distillation is a process based on the principle that different components boil at different temperatures. A mixture can be separated into different fractions by distillation on the basis that the fractions boil at different temperatures. A flash separator is a one-stage distillation process in which a mixture is separated into a vapor and a liquid. A distillation column can be thought of as several flashes in series in which each stage represents a flash process. For the purposes of modeling a distillation column, it is assumed that each stage is an equilibrium stage. In other words, the liquid and vapor leaving each stage are in thermodynamic equilibrium. In order to model distillation columns it is necessary that the following equations are satisfied:

(1) Equilibrium relationships
(2) Total material balance
(3) Component material balances
(4) Energy balances

These equations may be solved by grouping the equations by

type or stage and solving the groups one after another. A preferable method and the method used here is to solve all equations simultaneously to help to account for the interactions between variables.

Equilibrium Relationships

The equilibrium relationships relate the vapor and liquid compositions, decreasing the number of unknowns. The conditions for equilibrium are the same for two- and three-phase mixtures. The conditions for any number of phases in equilibrium are:

(1) The temperatures of each phase must be equal.

(2) The pressures of each phase must be equal.

(3) The possiblity of a molecule of a component to move from one phase to a second phase is exactly counterbalanced by the possiblity of another molecule of the same component moving into the first phase from the second phase.

The last condition is the relationship between the vapor and liquid compostion. This relationship can be described by any of a number of thermodynamic packages. The vapor and liquid compositions are related here by a combination of Henry's and Raoult's laws as:

$$Py_i = p_i x_i \qquad [1]$$

This simplifying assumption can be made since in crude distillation the columns operate at low pressures. It should be realized that this vapor pressure relation only applies in conditions of low pressure, ideal gas, and ideal solutions. The K-value is often used to relate the liquid

and vapor compositions, where K_i is a function of temperature and pressure. For any component i, y can be related to x by:

$$y_i=K_i x_i \qquad [2]$$

For Raoult's law then the K-value is given by:

$$K_i=p_i/P \qquad [3]$$

The mole fractions of each phase will always sum to unity. For two phases the following conditions hold:

$$\Sigma x_i=1.0 \qquad \Sigma y_i=1.0 \qquad [4]$$

In order for two or more phases to occur, the system must be at a temperature between the bubble point and the dew point. The bubble point is that temperature at which the system is all liquid with one drop of vapor just ready to form. The bubble point can be determined by finding the temperature which satisfies:

$$\Sigma K_i x_i=1.0 \qquad [5]$$

The dew point is that temperature at which the system is all vapor and one drop of liquid just ready to form. The dew point can be determined by finding the temperature which satisfies:

$$\Sigma y_i/K_i=1.0 \qquad [6]$$

When between the dew point and the bubble point, a flash calculation can be made at a specific temperature and

pressure to determine the amount of liquid and vapor produced. Equations [2] and [4] and a material balance equation must be satisfied.

Total Material Balance

The total material balance is just a conservation of mass equation written around either the whole column or a single tray. For any column, the sum of streams entering must be equal to the sum of the streams leaving the column. This can be expressed as:

$$F_j = D + B + \Sigma SL_j + \Sigma SV_j \qquad [7]$$

where the sum of the moles of each feed (F_j) equal the sum of the moles of distillate (D), bottoms (B), and each side product (SL_j,SV_j).

A schematic of a general single stage is shown in Figure 3. A material balance around this stage can be written as:

$$F_j=L_j+SL_j+V_j+SV_j-L_{j+1}-V_{j-1} \qquad [8]$$

Component Material Balance

The component material balances are written for each tray. These equations are similar to equation [8] except they are for a single component. For component i:

$$f_j=x_{i,j}(L_j+SL_j)+y_{i,j}(V_j+SV_j)-$$

$$x_{i,j+1}L_{j+1}-y_{i,j-1}V_{j-1} \qquad [9]$$

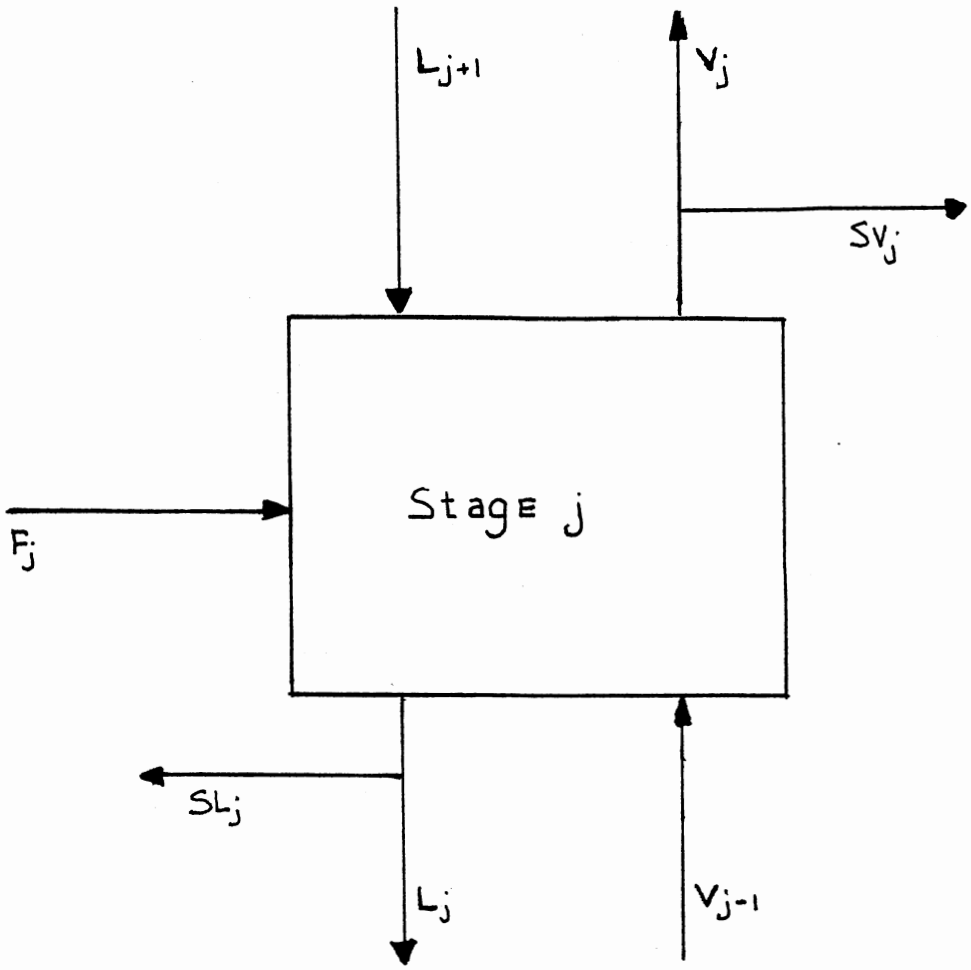


Figure 3: General Equilibrium Tray

If y_{ij} is replaced by $K_{ij}x_{ij}$, then for assumed liquid and vapor flow rates there are m, the number of stages, equations and m unknowns for each component i. Two terms can be defined, similar to Russell's (1983), which simplify equation [9]. They are:

$$B_{ij}=L_j+SL_j+K_{ij}(V_j+SV_j) \qquad [10]$$

$$C_{ij}=-K_{i,j-1}V_{j-1} \qquad [11]$$

Then the component material balance becomes for each tray:

$$f_j=x_{i,j}B_{ij}+x_{i,j-1}C_{ij}-x_{i,j+1}L_{j+1} \qquad [12]$$

This set of equations may be expressed in matrix format as:

$$\begin{bmatrix} B_{im} & C_{im} & & & \\ -L_{im} & B_{im-1} & C_{im-1} & & \\ & \ddots & \ddots & \ddots & \\ & & & -L_3 & B_{i2} & C_{i2} \\ & & & & -L_2 & B_{i1} \end{bmatrix} \begin{bmatrix} x_{im} \\ x_{im-1} \\ \vdots \\ x_{i2} \\ x_{i1} \end{bmatrix} = \begin{bmatrix} f_m \\ f_{m-1} \\ \vdots \\ f_2 \\ f_1 \end{bmatrix} \qquad [13]$$

This tridiagonal matrix format is easy to solve using a method similar to Russell's (1983). There are n matrix equations, one for each component.

In this program the liquid hydrocarbon and the liquid water are assumed immiscible. Therefore, the above equations are not affected by the presence of water or steam. The equations are written to take care of the compositions of the hydrocarbon phase. Since the water and hydrocarbon liquid are assumed immiscible, the liquid compositions in the hydrocarbon phase are not affected by

the presence of water. The water balance can be written around a single tray as:

$$f_{H2O} = v_{H2O,j} + l_{H2O,j} - v_{H2O,j-1} \qquad [14]$$

which implicitly includes the assumption that any liquid water produced is drawn from the column in a water draw tray. If water is produced on any tray the program will notify the user so that the user may change the specifications or make provisions for the water.

Energy Balances

The energy or heat balances are written around each tray. Each tray is assumed to be adiabatic, although it would be easy to incorporate a heat loss or gain on each tray. The total energy balance around the tray shown in Figure 3 is:

$$HF_j=(V_j+SV_j)H^V{}_j+(L_j+SL_j)H^L{}_j-V_{j-1}H^V{}_{j-1} - L_{j+1}H^L{}_{j+1} \qquad [15]$$

The coresponding total vapor and liquid enthalpies are obtained by:

$$H^L{}_j = \Sigma x_{ij} h^L{}_{ij} \qquad [16]$$

$$H^V{}_j = \Sigma y_{ij} h^V{}_{ij} \qquad [17]$$

Substituting these equations into equation [15] and writing y_{ij} in terms of K and x, the enthalpy balance becomes:

$$HF_j=(V_j+SV_j)\sum K_{ij}x_{ij}h^V{}_{ij} +(L_j+SL_j)\sum x_{ij}h^L{}_{ij} -V_{j-1}\sum K_{ij}x_{ij}h^V{}_{i,j-1} -L_{j+1}\sum x_{ij}h^L{}_{i,j+1} \quad [18]$$

When water or steam is present, the enthalpy effect of cooling the steam must be accounted for in the enthalpy balance. This is treated as a heat input to the stage. This heat input is added to the left side of equation [18].

Degrees of Freedom

The degrees of freedom of a system represents the number of process variables that must be set in order to completely describe a system. For an equilibrium system, the Gibb's phase rule (VanWylen and Sonntag, 1978) gives the degrees of freedom as:

$$F = C - P + 2 \quad [19]$$

where: F = the degrees of freedom

C = the number of components

P = the number of phases

This idea can be expanded to an equilibrium process. An analysis of this type was presented in Chapter II of the Phillips Fractionation Workshop (Erbar, 1983). The number of specification variables (degrees of freedom), N_s is given by:

$$N_S = N_V - N_R + N_T \qquad [20]$$

where: N_V = the total number of variables in the process

N_R = the number of variables fixed by restraints on the process

N_T = the number of recurring variables in the process

Using this analysis a process stream is determined to have C+2 independent variables. Therefore, a process stream can be completely described by specifing C independent component flow rates, the temperature, and the pressure. Likewise, the number of independent variables can be obtained for an equilibrium stage. The following is the procedure used to calculate N_S for a single equilibrium stage.

4 streams	4(C+2)
Heat leak	1
	N_V = 4C+9

Component balance equations	C
Phase Distribution equations	C
Energy Balance	1
Equality of temperatures and pressures of stream; leaving	2
	N_R = 2C+3

Therefore, for a single stage, the number of process variables that need to be specified are N_S = 2C+6. The stage could be set by completely specifying the two feeds to the stage, the pressure on the stage, and the heat leak from the stage.

Similar procedures can be used to find the number of specifications required for a general column. The number of specification variables for a condenser or reboiler is C+4.

For a reflux divider, there are C+5 variables. There are 3C+8 variables required for a feed stage. The following is an itemization of the procedure for a single feed column with m stages:

m equilibrium stages	m(2C+6)
Feed stage	3C+8
Reboiler and Condenser	2(C+4)
Reflux Divider	C+5
N_v =	(2m+6)C+6m+21

Implied	
The 2(m-1) interconnecting stages	2(m-1)(C+2)
7 Inteconnecting Streams	7C+14
N_R =	(2m+5)C+4m+10

The number of stages	1
The location of the feed tray	1
N_T =	2

$$N_s = C + 2m + 13$$

The column could be described by specifing the feed stream and tray location, the total number of trays, the pressure and heat leak on each tray and feed tray, pressure in the condenser and reboiler, pressure and heat leak in the divider, temperature out of the condenser, reflux rate and bottoms rate. This represents only one way to specifiy a column. Any of a number of different variables may be specified but the number of specified variables must remain the same.

Simultaneous Solution

The material and energy balance equations given earlier are solved simultaneously by a modified Newton-Raphson technique by Newman. The equations are organized by tray in

the following order: component material balances around the stage, sum of mole fractions equation, energy balance around the stage, and total material balance around the stage. There are n+3 equations for each stage. These equations are then represented by:

$$\bar{A}\,\bar{X} = \bar{f} \qquad [21]$$

where $\bar{X}$ is the vector of process variables to be solved for and $\bar{f}$ is the left side of each equation called the forcing function. The Jacobian matrix is the matrix of partial derivatives.

$$\bar{J} = (d\bar{f}_i/d\bar{X}_i) \qquad [22]$$

Using the Newton-Raphson technique, the updated process variables can be obtained by:

$$\bar{X}_{i+1} = \bar{X}_i - \bar{J}_i^{-1} * \bar{f}_i \qquad [23]$$

where $\bar{J}^{-1}$ is the inverse of the Jacobian.

Using Newman's method the Jacobian matrix is a block band tridiagonal matrix. The inversion of the Jacobian is done in steps, working on the three matrices of the band for each tray. This eliminates the need to store the entire Jacobian at once, storing only the inverted Jacobian. The inversion technique is detailed by Newman (1967).

The Jacobian is made up of the following matrices. For each tray, the three matrices making up the band elements are A, B, and D. The A matrix is the partial derivatives of the change in the forcing function with respect to the tray

below ($d\bar{f}_j/d\bar{X}_{j-1}$). Likewise, B is the matrix of partial derivatives with respect to the tray in question, and D is the matrix of partial derivatives with respect to the tray above. The Jacobian then becomes:

$$\bar{J} = \begin{bmatrix} B_1 & D_1 & & & \\ A_2 & B_2 & D_2 & & \\ & \ddots & \ddots & \ddots & \\ & & A_{m-1} & B_{m-1} & D_{m-1} \\ & & & A_m & B_m \end{bmatrix} \qquad [24]$$

Sidestream Strippers

The side stripper is solved in the same manner as the main column, using the same set of equations. The stripper is different in the respect that it has no condenser or reboiler, and only one liquid feed on the top tray. The stripper is solved using an assumed composition of the liquid feed equal to the current composition of the tray in the main column from which it is drawn. The stripper is completely solved between iterations of the main column, using the vapor from the top tray of the stripper as a feed to the main column.

Pumparounds

Pumparound streams are solved as a combination of a side product stream and a feed stream to the main column. The pumparound stream re-enters the column, as a feed, at a reduced enthalpy. The composition profile of the pumparound

stream is updated between each iteration as the composition of the tray from which it is drawn changes. The solution requires that the user set a duty for heat removal in the pumparound stream.

Simulation

The program is designed to model a crude tower or conventional distillation column in continuous operation. The program will not model startup, shutdown, or transient behavior. The description of the program is split up into the following sections: nomenclature, block data, input section, the main program, and subroutines. The nomenclature section is provided to supply the reader with a reference for all variables used in the code. The listing is a quite extensive list of variables, arrays and subroutines. Included are definitions for each listing.

Block Data

The block data section contains all of the data needed for use in the program for each of the pure components listed in Table II. The component database contains data for Antoine's equation coefficients for calculating the vapor pressure, molecular weights, ideal gas enthalpy coefficients, heats of vaporization, and specific gravities. The coefficients of Antoine's equation were obtained from three sources: Jordan (1954), Lange's Handbook of Chemistry (Dean, 1985), and Felder and Rousseau (1978). The ideal gas enthalpy coefficients are from the API Technical Data Book

TABLE II

PURE COMPONENT LIST

NUMBER	NAME	NUMBER	NAME
1	Hydrogen	32	2-Methyl-1-Butene
2	Methane	33	3-Methyl-1-Butene
3	Ethane	34	2-Methyl-2-Butene
4	Propane	35	1-Hexene
5	i-Butane	36	Cyclopentane
6	n-Butane	37	Methylcyclopentane
7	i-Pentane	38	Cyclohexane
8	n-Pentane	39	Methylcyclohexane
9	neo-Pentane	40	Benzene
10	n-Hexane	41	Toluene
11	n-Heptane	42	o-Xylene
12	n-Octane	43	m-Xylene
13	n-Nonane	44	p-Xylene
14	n-Decane	45	Ethylbenzene
15	n-Undecane	46	Nitrogen
16	n-Dodecane	47	Oxygen
17	n-Tridecane	48	Carbon Monoxide
18	n-Tetradecane	49	Carbon Dioxide
19	n-Pentadecane	50	Hydrogen Sulfide
20	n-Hexadecane	51	Sulfur Dioxide
21	n-Heptadecane	52	2-methyl-Pentane
22	Ethylene	53	3-methyl-Pentane
23	Propylene	54	2,2 dimethyl-Butane
24	1-Butene	55	2,3 dimethyl-Butane
25	cis-2-Butene	56	1-Heptene
26	trans-2-Butene	57	Propadiene
27	i-Butene	58	1,2-Butadiene
28	1,3 Butadiene	59	Ethylcyclohexane
29	1-Pentene	60	Ethylcyclohexane
30	cis-2-Pentene	61	Water
31	trans-2-Pentene		

(1976) and the heats of vaporization from Lange's Handbook of Chemistry (Dean, 1985). The specific gravities are at $60^{o}F$ referenced to water at $60^{o}F$. They come from Lange's Handbook of Chemisty (Dean, 1985) and GPSA Databook (1981).

Although data exists for only 61 components, the data base can be easily extended to accomodate any of a number of components. To extend the utility of the program, the space has been provided to allow for hypothetical components.

Input Section

The input section is part of the main program. Data may be input interactively from a terminal or from a file. After inputting data interactively, the program allows the user to save the file. This allows the user to make simple changes to the column without re-entering all of the data.

The procedure for entering the data by file is reviewed. Data is entered in the same order from a terminal. Interactive input is not discussed here as the program is user friendly and self explanatory. The easiest method is to enter data interactively first and save the file. The format of the file is given so that the user may make changes. To enter data by file, the user must create a data file. This data file name must be entered into the program. The format of the file is given in Figure 4. Where the variables are listed the values of those variables should be substituted. The file is set up to allow the user to substitute his own physical and thermodynamic properties instead of those calculated by the program. This increases

```
FILENAME.DAT

TITLE
NC,NCO,NSTAR,NH2O

ID(I),AK(I),BK(I),CK(I)
AH(I),BH(I),CH(I),DH(I)                    THIS SECTION IS
EH(I),FH(I),WMOL(I),DELH(I)                REPEATED FOR
GSG(I),BP(I)                               EACH COMPONENT

NS,NF,LIM
JCOTYP,ND,NSDS
REBT
STEAM,TSTEAM
AL(NS),AL(1)
T(NS),T(1)
P(1)
P(2)
  .
  .
  .
P(NS-1)
P(NS)

IFTRAY(JF),WLF(J),TFEED(J)
FX(1,J)
FX(2,J)
   .                                       THIS SECTION IS
   .                                       REPEATED FOR
   .                                       EACH FEED
FX(NC-1,J)
FX(NC,J)
HF(J),F(J)


IDRAW(JD)                                  REPEAT FOR
SL(J),SV(J)                                EACH DRAW


ISTRAY(JJ)
SL(JS)
NSS(JS)                                    REPEAT FOR
SR(JJ),TS(JJ)                              EACH STRIPPER
ALS(JJ)
PS(JJ)
```

Figure 4: Input Data File Form

the flexibility of the program and is very useful when working with hypothetical components.

The Main Program

The main program reads in all the input data, initializes liquid, temperature, and vapor profiles, calculates the assumed liquid composition, formulates the Jacobian, calls the solving subroutines, and updates the profiles.

This same routine is followed for both the main column and the strippers. The order for solution is as follows:

(1) Constant molal overflow iterations are done for the main column.

(2) Each stripper is solved using the current liquid compositions on the stage from which the liquid feed is drawn.

(3) The vapor leaving each stripper is fed to the main column.

(4) The feed to the column from a pumparound stream is assumed.

(5) One iteration of the main column is performed.

(6) Check for convergence.

(7) Return to step 2 if convergence criteria are not satisfied.

To further illustrate how the program works, the procedure for solving a single distillation column will be reviewed.

Initially, the program sets all variables to zero to

clear any data from previous runs. The input sections reads data interactively from a terminal or from a file. A database is setup which contains all of the process variables and the data for each component from block data. Input variables include pressures and estimated tray temperatures, number of trays, reflux rate and component identification numbers. For each feed, the inputs are feed rate, liquid fraction, temperature, and component feed rates. Component feed rates should be entered in lb moles.

Hypothetical components can be specified when entering the component identification number. For each hypothetical component the program will prompt the user for the normal boiling point, the molecular weight, and the specific gravity. If the molecular weight is unknown, the program can calculate it if the user desires. The program then calculates the coefficients for Antoines' equation using the formula of Dreisbach (1952). The program will allow the user to change the input variables for the hypothetical components if these coefficients are not acceptable. Once all of the input data has been read, the program allows the user to save the input data to a file. This allows simple changes to be made to the file, without having to re-enter all of the data.

The first step is to initialize temperatures and flow rates. All tray temperatures may be input if desired. Otherwise, the program will calculate intermediate tray temperatures based on a linearization of the top and bottom temperatures over all trays. The liquid flows are

initialized to constant molal overflow flow rates.

The next step occurs within the iteration loop and is the first step of every iteration. Vapor flow rates are generated from a material balance around each tray using the assumed liquid rate profile. The component material balances are set up by component. The resulting tridiagonal matrices are solved by a method similar to Russell's (1983). The result is the liquid composition profile.

The next step is to form the matrix storage arrays which store the Jacobian. The three matrices applying to any one tray are specified for constant molal overflow. If the iteration is a constant molal overflow iteration, the subroutine Band used to solve and invert the Jacobian is called. After calling Band for each tray, the temperature profiles are updated and one iteration is complete. The program returns to the section that calculates vapor flows and starts another iteration.

For non-constant molal overflow iterations, full heat balances are done on each tray. The three matrices must be further specified for the partial derivatives of the enthalpy balance and the total material balance. Once Band has been called for each tray, the liquid and temperature profiles are updated. Liquid changes are limited to 40% fractional change of the current liquid flow for that tray. Temperature changes are limited to $10^{o}F$ on each stage. These limits can be easily changed. Once the profiles are updated, the program returns to the section that calculates the vapor profiles.

Once the liquid compositions are calculated, the program checks for convergence. The program checks the change in liquid compositions from one iteration to the next to see if the difference is less than the tolerance. Next, the program checks the sum of the liquid changes and the sum of the temperature changes. If any of these convergence criteria are not met, the program performs another iteration. If the upper limit of iterations is exceeded, the program exits with a message of an unconverged solution.

The output section calculates reboiler and condenser duties. The program outputs all feeds and products. Also printed out are the amount of sour water produced and the reboiler and condenser duties. The program prints out vapor, liquid, and temperature profiles and the liquid composition profile.

The method described above is for a single column. The solution method with multiple columns is similar with an inner convergence loop of all strippers inside the outer convergence loop of the main column.

Band and Solver Subroutines

The subroutines BAND and SOLVER are modified code from Newman (1967). No attempt was made to derive a matrix solution. Many matrix solution methods are available and can be easily incorporated using the same method as used here. Newman's program was selected for its widely convergent characteristics. The program is versatile and has many uses.

These subroutines invert and solve the Jacobian matrix. The subroutine BAND was developed so that the entire Jacobian does not need to be stored at once. The main program calls BAND on a tray by tray basis. BAND partially inverts the Jacobian at each call, storing the partial Jacobian in the E array. The subroutine BAND calls the subroutine SOLVER to solve the inverted Jacobian. Once BAND has been called for each tray the entire Jacobian has been inverted and solved.

Equilibrium Calculation Subroutine

The subroutine that calculates the K-values and K-value derivatives is called KVALUE. The K-value calculations are based on Raoult's and Henry's law:

$$y_i P = p_i x_i \qquad [1]$$

This law is valid for ideal mixtures at low pressures. Raoult's law is usually valid for values of x_i close to unity and can be valid for any value of x_i where the mixture is of similar substances such as a mixture of straight-chain hydrocarbons (Felder and Rousseau, 1978). According to Green (1984), equation [1] can be used for K-value data if the total system pressure is less than 2 atmospheres and the liquid phase is ideal. From equation [1], the K-values, K_i, which are defined as y_i/x_i may be determined as:

$$K_i = p_i/P \qquad [3]$$

Antoine's equation is used to calculate the vapor

pressure. The form of Antoine's equation is:

$$\log p_i = A_i - B_i/(T+C_i) \qquad [25]$$

For the 61 component database, the values of A_i, B_i and C_i have been provided. For these coefficients, the vapor pressure is in mm Hg and the temperature in ^oC. For hypothetical fractions, equation [25] becomes:

$$\ln p_i = A_i - B_i/(T+C_i) \qquad [26]$$

The values of A_i, B_i and C_i are calculated by the main program using Dreisbach's (1952) relation. The vapor pressure is in psia and the temperature in ^oF.

K-value temperature derivatives are also calculated in this subroutine. For the pure components, the K-value derivatives are:

$$EQDT = p_i(-B_i/(T+C_i)^2 \log 10/P \qquad [27]$$

For the hypothetical fractions, the derivatives are:

$$EQDT = p_i(-B_i/(T+C_i)^2)/P \qquad [28]$$

Enthalpy Calculation Subroutines

The enthalpy calculations include calculations for liquid and vapor enthalpies and enthalpy temperature derivatives.

The subroutine VENTH calculates the vapor enthalpies. For the pure components, the enthalpy is calculated from the correlation in Chapter 7 of the American Petroleum

Institute's Technical Databook (1976). The correlation is for ideal gas enthalpies and is a function of temperature only. The hypothetical enthalpy calculations come from the GPAK*H computer program (Erbar, 1974). The coefficients used are functions of the specific gravity and the mean average boiling point of the fraction. The enthalpy is a function of temperature.

The VLNTH subroutine calculates the liquid enthalpies. The same correlations used above are used here. The liquid enthalpy of the pure components are obtained by subtracting the heat of vaporization from the vapor enthalpy.

The VENHDT calculates the temperature derivatives of the enthalpies. The heat of vaporization assumed not to be a function of temperature. Therefore, the derivatives of the vapor and liquid enthalpies are the same.

Solution Strategy

The details of each section of the program have been presented. The purpose of this section is to show how the sections fit together. The solution strategy is detailed, similar to the solution strategy in Gas Conditioning and Processing, Volume 3 (Maddox and Erbar, 1982).

(1) Input the system conditions
   A) Temperatures of top and bottom stages
   B) Components and relative amounts
   C) Hypothetical component parameters
(2) Calculate the feed conditions
(3) Estimate the temperature and liquid rate profiles

(4) Calculate the vapor rate profile

(5) Calculate the K-values based on the estimated temperatures

(6) Estimate the component liquid and vapor rates for each tray

(7) Set up and solve each stripper

(8) Set return feeds to the main column

(9) Estimate Pumparound streams

(10) Correct temperature and liquid and vapor rate profiles

(11) Check for convergence

(12) Return to step (4) until desired tolerance is reached

(13) Output input variables and solution

CHAPTER IV

RESULTS AND DISCUSSION

Introduction

The intent of this section is to illustrate the uses and validity of the model. Different case studies will be reviewed to show the range of applications of the program. For each case study, the assumptions used and the input variables will be given. The results of each study are given and comparisons to existing data, where available, will be presented. The final tests of the program will be an atmospheric column design and a model of an existing crude unit compared with industrial operating data.

Convergence Techniques

It is important to have good initial estimates of the column conditions for a Newton-Raphson technique of solution. When using the program to model an existing column, these initial estimates are readily available. For these conditions, convergence is virtually assured. When designing a new column, it is necessary to properly estimate temperatures and flows.

To model a crude column it is important to properly characterize the crude. The characterization is dependent

upon the number and size of the fractions. If fractions are split up on a TBP curve, it is important to carefully extrapolate the curve to split up the light and heavy ends. These fractions will not be very accurate so it is important to take as much care in doing this as possible. By increasing the number of fractions, the accuracy of the average boiling point and specific gravity of each fraction is increased. This increases the overall accuracy of the characterization.

When designing a crude tower, it is important to carefully analyze the amounts of each cut. It is recommended that the user utilize the shortcut techniques of Watkins (1979). From Watkins method it is easy to obtain the approximate steam rates to the column and the strippers, the number of trays between cuts, and reflux to the column. When a column is estimated in this manner, speed of convergence will be greatly increased. The most crucial elements to convergence are the specifications of reflux rate and feed boilup. When these specifactions are reasonable, the model will converge for wide ranges of starting temperature profiles.

Model Verification

The different elements of the program were each analyzed by a series of case studies. The purpose of these case studies was to analyze independently each aspect of the program and verify its validity. The inputs, outputs, and results are presented in Appendix B.

The entire program is tested by simulating three different crude towers. The first crude tower is an atmospheric crude tower design. The results are compared with the amounts of each product published for the specific crude. The other two simulations are models of an existing crude unit. The results are compared to operating conditions of these columns. The crude characterizaion, simulation inputs and outputs, and the results are presented in the Appendices.

Case Study I

The first test of the program was to test the default column, a simple conventional 2-phase distillation column. A hypothetical column of 13 trays and one feed was solved. The solution was compared to the solution of the same column from MAXISIM (Erbar, 1987), a proven successful simulator.

Table III shows the comparison of the two simulators for product rates. The temperature and liquid rate profiles are very similar and well within a 1-2% error. These differences can be accounted for by the use of vapor pressures to calculate K-values and the use of ideal gas enthalpies instead of the use of the SRK equation of state.

Case Study II

The second test was to compare the program to published tray by tray output for a stripper. The H_2S stripper data was found in Gas Conditioning and Processing (Erbar and Maddox, 1982). The intent of this test was to show that the

TABLE III

PRODUCT RATE COMPARISON OF 2 PHASE DISTILLATION

	DISTILLATE			
	MAXISIM	PROGRAM	DEV	%DEV
Component				
C3H8	1.000	1.000	0.000	0.00
NC4H10	2.000	2.000	0.000	0.00
NC5H12	2.998	2.998	0.000	0.01
NC7H16	2.002	2.002	-0.000	-0.02
NC10H22	0.000	0.000	0.000	0.00
NC13H28	0.000	0.000	0.000	0.00
NC15H32	0.000	0.000	0.000	0.00
RATE, LBMOLS/HR	8.000	8.000	0.000	0.00
TEMPERATURE, F	126.870	127.117	-0.247	

	BOTTOMS			
	MAXISIM	PROGRAM	DEV	%DEV
Component				
C3H8	0.000	0.000	0.000	0.00
NC4H10	0.000	0.000	0.000	0.00
NC5H12	0.002	0.002	0.000	0.00
NC7H16	2.998	2.998	0.000	0.00
NC10H22	22.000	22.000	0.000	0.00
NC13H28	30.000	30.000	0.000	0.00
NC15H32	40.000	40.000	0.000	0.00
RATE, LBMOLS/HR	95.000	95.000	0.000	0.00
TEMPERATURE, F	373.280	367.579	5.701	
DUTIES				
REBOILER, BTU/HR	947604	708608	238997	
CONDENSER, BTU/HR	-163109	-152686	-10423	

program worked for columns without a reboiler or a condenser.

The products obtained from the program and those published are shown in Table IV. The products produced by the program are very close to those printed by Erbar and Maddox (1982). The errors in distribution of the components can be attributed to the differences in K-values. The heavy component was only classified as 'heavy' with the molecular weight given. The molecular weight was closest to heptadecane so this component was used to approximate the heavy fraction. The success of this simulation shows that the program can be applied to the solution of a steam stripper, which is very similar to the H_2S stripper solved here.

Case Study III

This case study involves the design of a crude tower column for an Iranian crude. The characterization of this crude is given in Table V. The crude tower was designed with sidedraws in place of sidestream strippers and no pumparounds. The column is steam stripped with no reboiler. This test show the applicability of the program for columns using steam stripping and the accuracy of the hypothetical component correlation.

Yields for each cut in an atmospheric column using this crude were given with the assay data along with the required number of trays between cuts. No mention was made of steam rates to the column, operating conditions, or pumparounds.

TABLE IV

H2S STRIPPER PRODUCT COMPARISON

GAS LEAVING

Component	PUBLISHED	PROGRAM	DEV	%DEV
H2S	4.99	5.34	0.35	6.95%
N2	6.82	6.85	0.03	0.41%
CO2	25.82	31.46	5.64	21.83%
CH4	248.49	258.97	10.48	4.22%
C2H6	67.40	77.81	10.41	15.45%
C3H8	75.64	88.81	13.17	17.41%
iC4H10	17.67	19.86	2.19	12.42%
nC4H10	47.45	53.96	6.51	13.72%
iC5H12	15.91	17.42	1.51	9.48%
nC5H12	28.58	31.18	2.60	9.08%
HEAVY	0.00	0.00	0.00	0.00%
RATE, LBMOLS/HR	538.77	591.65	52.88	9.81%
TEMPERATURE, F	130.09	129.64	-0.45	

STRIPPED CRUDE

Component	PUBLISHED	PROGRAM	DEV	%DEV
H2S	2.00	1.65	-0.35	-17.35%
N2	0.25	0.22	-0.03	-12.00%
CO2	5.64	0.01	-5.63	-99.82%
CH4	27.80	17.32	-10.48	-37.70%
C2H6	38.66	28.26	-10.40	-26.90%
C3H8	98.36	85.19	-13.17	-13.39%
iC4H10	48.43	46.24	-2.19	-4.53%
nC4H10	181.97	175.46	-6.51	-3.58%
iC5H12	145.92	144.41	-1.51	-1.03%
nC5H12	338.50	335.90	-2.60	-0.77%
HEAVY	5011.48	5011.48	0.00	0.00%
RATE, LBMOLS/HR	5899.01	5846.14	-52.87	-0.90%
TEMPERATURE, F	129.05	128.36	-0.69	

TABLE V

IRANIAN CRUDE CHARACTERIZATION

COMPONENT	VOL%	TBP	MWT	S.G.
C2	0.04			
C3	0.62			
IC4	0.32			
NC4	1.34			
IC5	1.23			
NC5	1.65			
C6	3.39			
FRACTION 1	1.41	173	90	0.7201
FRACTION 2	6.00	215	104	0.7366
FRACTION 3	6.00	280	116	0.7620
FRACTION 4	8.00	354	140	0.7826
FRACTION 5	8.00	430	170	0.8040
FRACTION 6	8.00	510	195	0.8373
FRACTION 7	8.00	590	230	0.8524
FRACTION 8	8.00	663	265	0.8794
FRACTION 9	8.00	748	340	0.9013
FRACTION 10	6.00	832	420	0.9309
FRACTION 11	6.00	920	480	0.9459
FRACTION 12	6.00	1090	550	0.9712
FRACTION 13	6.00	1130	600	1.0283
FRACTION 14	6.00	1190	690	1.0528

OVERALL CRUDE	
MOLECULAR WEIGHT	233.6
SPECIFIC GRAVITY	0.8756

The program was run using the data available making approximations for reflux rate, steam rate, and pressure.

Comparisons of the yields are given in Table VI. The volume percent yields and gravities obtained by the program are well within the errors expected when designing a column. The column design including inputs and outputs is given in Appendix B.

Case Study IV

The following case study uses the same data and column conditions as in case study III. The column has all the same features as before except now it includes a pumparound stream. The pumparound stream does not upset the balance of products and requires only a few extra iterations. The computer output for this column is given in Appendix B.

Case Study V

This case study uses the same crude and column conditions as in case study III. The only difference is that one of the sidedraws is steam stripped. This case study verifies the sidestream stripper section of the program. The program also allows for the pumpback of some of the liquid product from the stripper to the main column. The pumpback was also applied here. The computer outputs for this case study are located in Appendix B.

Iranian Crude Tower Design

An atmospheric crude column was designed for an Iranian

TABLE VI

COMPARISON OF YIELDS IN CASE STUDY III

	PRODUCT YIELDS		
PRODUCT	PUBLISHED	PROGRAM	DEV
OVERHEAD GAS	2.80%	1.30%	-1.50
GASOLINE	9.30%	9.90%	0.60
No. 1 CUT	9.70%	10.60%	0.90
No. 2 CUT	4.00%	4.00%	0.00
No. 3 CUT	3.90%	4.50%	0.60
No. 4 CUT	13.60%	12.20%	-1.40
RESIDUUM	56.70%	57.50%	0.80

crude whose characterization is given in Table V. The column was designed with four side cuts with the bottom three cuts steam stripped. No pumparounds or pumpbacks were included. The number of trays between cuts was given with the assay data along with volume percent yields. These numbers were used along with estimates of the reflux rate, steam rate and temperature, and pressure of the column. The inputs and outputs of the simulation for this column are given in Appendix C. The program took 25 iterations to converge. The temperature profile and liquid rate profile are shown in Figures 5 and 6, respectively. The profiles look reasonable and the yields are very similar to those published with the assay data.

Louisianna Crude Unit Comparison

The final test of the program was to compare it to industrial operating data available for an existing crude unit. The model consists of two columns, a preflash tower and an atmospheric tower. The products from the preflash tower are the feeds to the atmospheric tower.

The crude charge to the preflash tower is a mixed Louisianna crude. A Louisianna mixed crude was chosen that seemed comparable to the one at which the unit was operating, which was unavailable. The major difference between the characterized crude and the operating crude is in the volume percent of light ends. The characterized crude has only about 1% light ends where the operating crude has about 3% light ends. The difference in light end

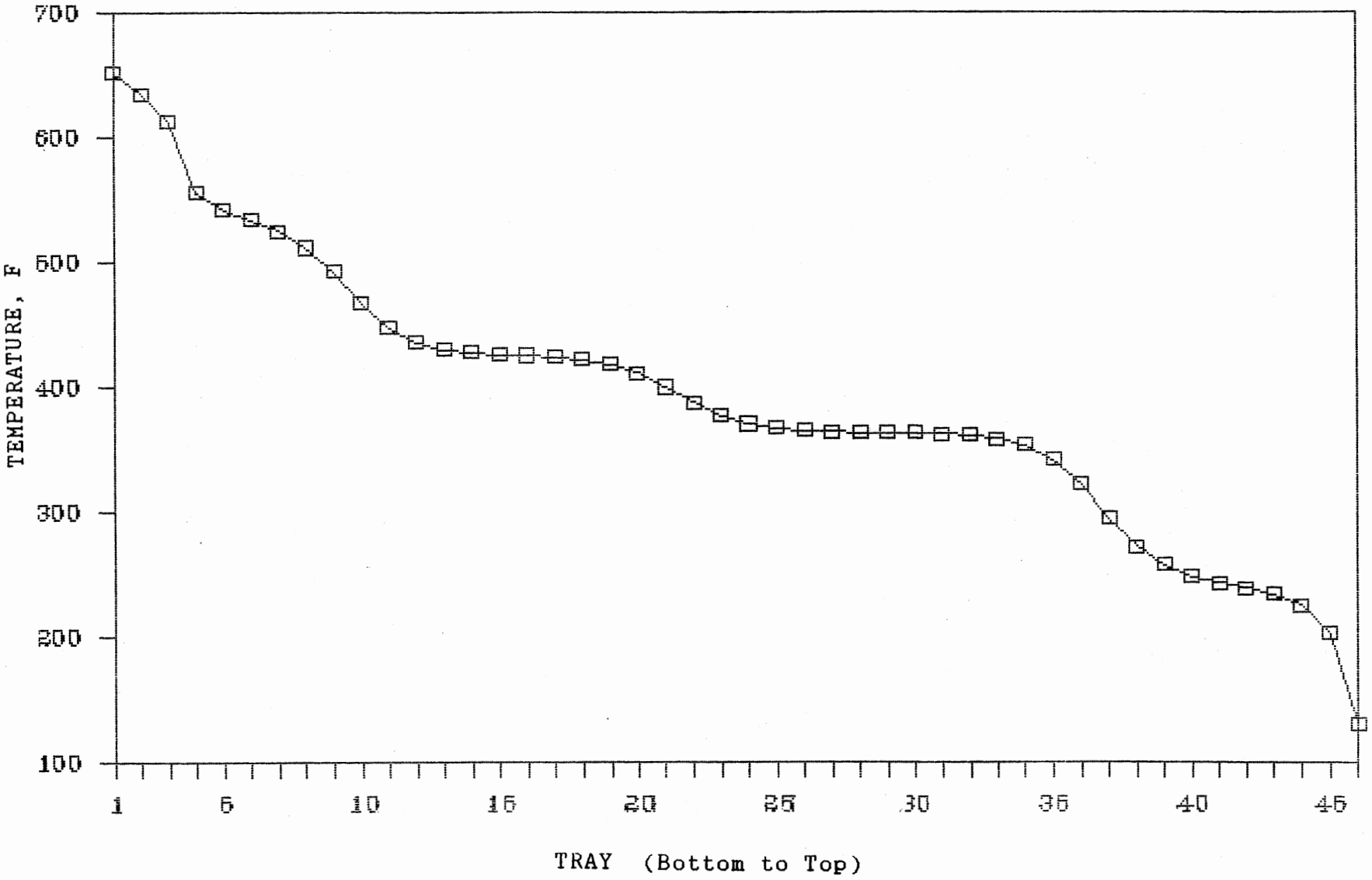


Figure 5: Iranian Crude Temperature Profile

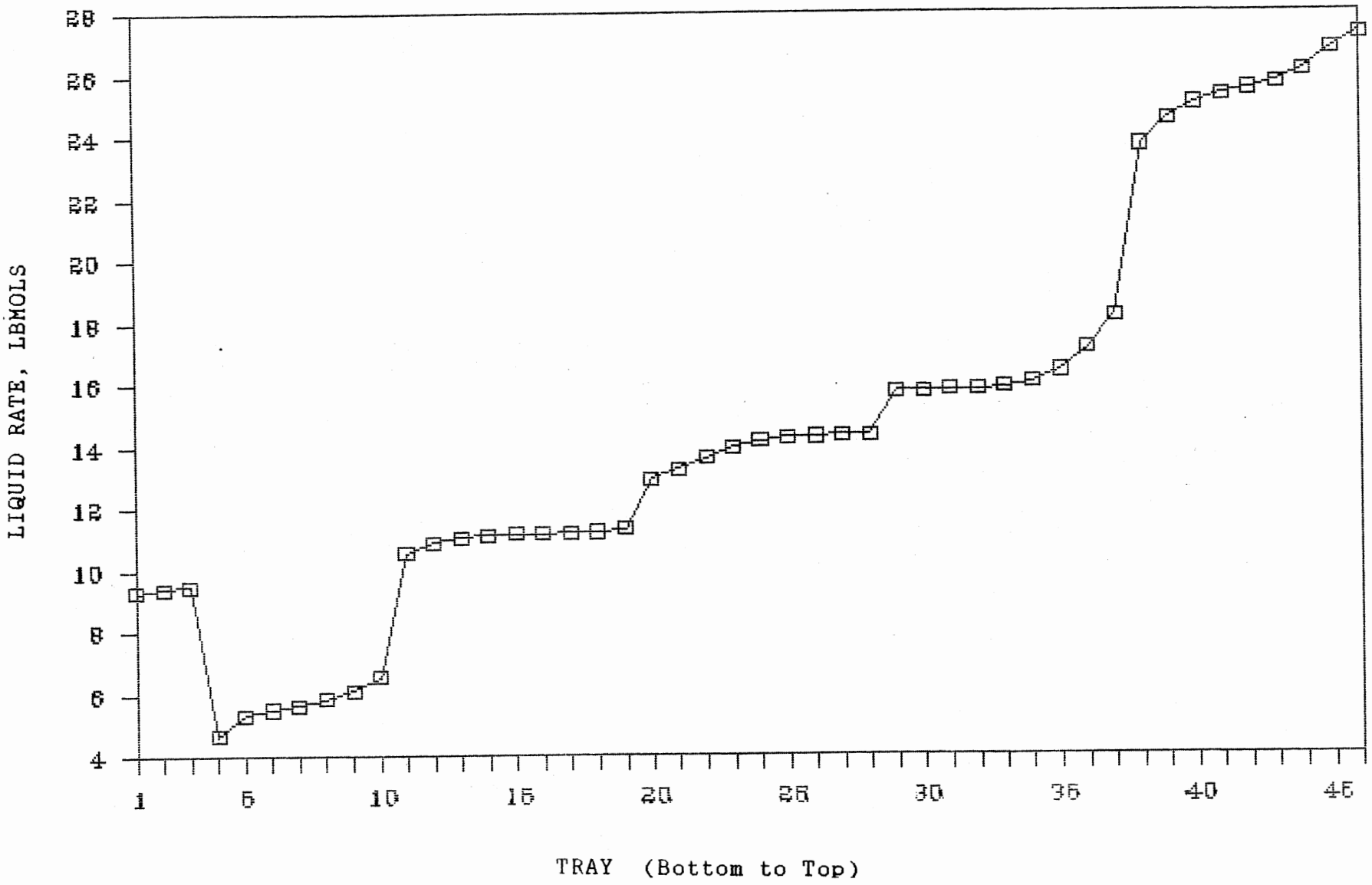


Figure 6: Iranian Crude Liquid Rate Profile

analysis should be considered when looking at the results obtained. The characterization of the crude used in the model is given in Table VII. The results and outputs from the program are given for both columns in Appendix D.

Operating temperature comparisons for both towers are given in Table VIII. Overall, the temperatures are good, but tend to deviate from experimental data by approximately twenty degrees. Part of this error can be attributed to the accuracy of the characterization of the crude. The difference in the condenser temperature can be attributed to the light ends difference discussed above.

Figure 7 shows the temperature profile for the atmospheric column. Overlayed on the graph as, +'s, are the operating temperatures for that column. The profile is smooth and proceeds in the same direction as the operating data.

Each of the side cuts and products were compared based on their gravities. Table IX shows the comparison of each of these gravities with operating data. The numbers obtained from the program are extremely close and show that the program models quite accurately the steam strippers and the splits of each cut.

TABLE VII

LOUISIANNA CRUDE CHARACTERIZATION

	VOL%	TBP	MWT	S.G.
COMPONENT				
C2	0.03			
C3	0.20			
IC4	0.20			
NC4	0.27			
IC5	0.48			
NC5	0.44			
CYCC5	0.02			
FRACTION 1	1.38	110	77	0.6410
FRACTION 2	2.00	143	87	0.6600
FRACTION 3	4.00	190	101	0.6830
FRACTION 4	4.00	242	115	0.7300
FRACTION 5	4.00	288	129	0.7610
FRACTION 6	4.00	332	143	0.7800
FRACTION 7	4.00	374	158	0.7950
FRACTION 8	6.00	416	174	0.8080
FRACTION 9	6.00	461	193	0.8200
FRACTION 10	6.00	508	214	0.8300
FRACTION 11	6.00	552	236	0.8410
FRACTION 12	6.00	599	260	0.8530
FRACTION 13	4.00	635	279	0.8600
FRACTION 14	4.00	664	296	0.8680
FRACTION 15	4.00	697	315	0.8740
FRACTION 16	6.00	744	344	0.8820
FRACTION 17	6.00	810	386	0.8950
FRACTION 18	6.00	886	437	0.9100
FRACTION 19	6.00	980	502	0.9300
FRACTION 20	6.00	1095	578	0.9610
FRACTION 21	3.00	1210	657	0.9840
OVERALL CRUDE				
MOLECULAR WEIGHT		220.6		
SPECIFIC GRAVITY		0.8390		

TABLE VIII

LOUISIANNA CRUDE TEMPERATURE COMPARISON

LOCATION	PROGRAM	OPERATING DATA	DEV
PREFLASH FEED	479	479	0
PREFLASH BOTTOMS	488	472	16
PREFLASH OVHD.	429	407	22
TRAY 6	730	701	29
TRAY 11	647	645	2
TRAY 13	597	576	21
TRAY 18	452	435	17
TRAY 23	332	300	32
OVERHEAD	248	200	48

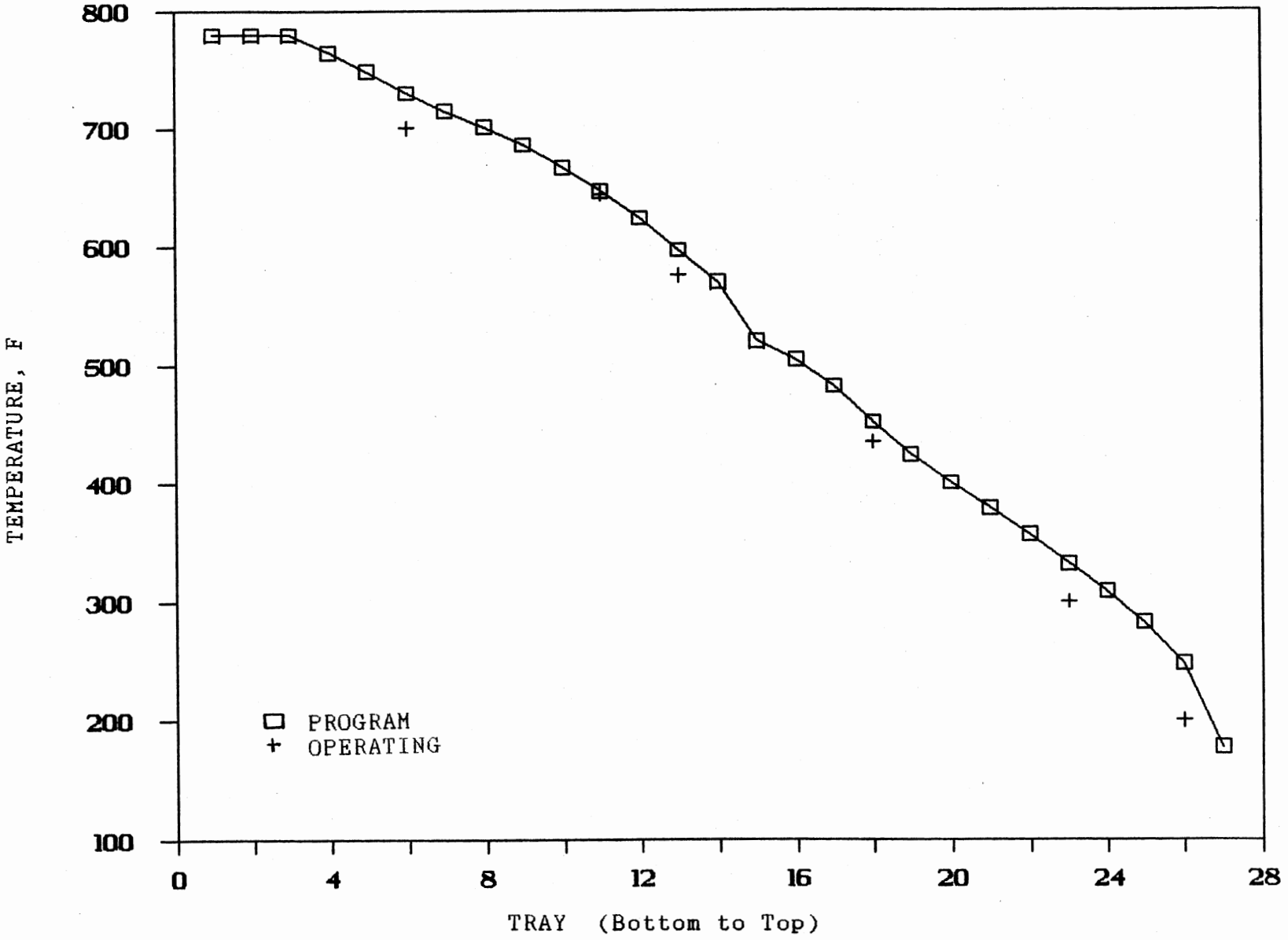


Figure 7: Louisianna Crude Temperature Profile

TABLE IX

LOUISIANNA CRUDE PRODUCT GRAVITY COMPARISON

LOCATION	PROGRAM API	OPERATING DATA, API	DEV
LSR GASOLINE	75.6	72.9	2.7
NAPHTHA	54.1	54.9	-0.8
KEROSENE	43.2	43.7	-0.5
DIESEL	35.5	35.8	-0.3
ATMO. GAS OIL	29.8	29.8	0.0
TOPPED CRUDE	25.3	22.4	2.9

CHAPTER V

CONCLUSION AND RECOMMENDATIONS

Conclusion

A crude tower distillation simulator has been developed for the purpose of modeling steady state crude distillation behavior. The program may be run either interactively or by batch input. A large pure component data base has been provided with provisions for the user to add more pure components. An algorithm was included to allow the user to use hypothetical components. The user may specify initial temperature and liquid rate profiles or have the program calculate them on the basis of top and bottom estimates. An extensive printout of the program results may be requested when solution is complete. The printout includes descriptions of feeds, distillate, bottoms, side draws, and sidestrippers. Also included are the steam rates, pumparound streams, duties and sour water production. The output also includes a tray by tray listing of temperatures, liquid and vapor flows, molecular weights and densities, and a liquid composition profile.

The model was checked for each aspect of the program using a series of case studies. In each of these case studies the model converged easily and fairly rapidly. The

results of each of these case studies independently verified each aspect of the program. The simulation was also checked for its validity in designing a crude column. The outputs from this check were found to be reasonable. The final test was a model of an existing crude unit with comparisons to operating data. The output from the program for these columns were very accurate. This test points to the accuracy of the simulation and its validity for use in modeling crude columns.

The model was found to be widely convergent for a large variety of crudes. Convergence is highly dependent upon the liquid profile or that reflux rate and the feed specifications provide enough liquid reflux to the column. Under these conditions, convergence is fairly independent of the starting temperature profile and was almost assured for temperatures off by as much as 100^{o}F.

Recommendations

The author realizes that no simulation is without fault and there is always room for improvement. The following recommendations represent some of the areas in which the author feels this simulation can be improved.

Newman's solution technique was used here to solve the main column and each stripper. This technique was chosen for its generality and its ability to converge for a wide variety of situations. However, there are several solution techniques available which may be more efficient. One such technique is Boston's 'inside-out' technique. This

technique would probably be more stable for the strippers and increase the speed of convergence.

There are several areas in which storage space could be decreased to make the program more applicable to microcomputers. The data base could be located in a separate file to decrease the amount of data that needs to be stored by the main program.

The current solution method initially uses the assumption of linear temperature profiles when temperature profiles are not given. For a crude column, this is typically not a good assumption and it greatly increases the number of iterations. A new method for estimating the temperature profile should increase the speed of convergence. One such method would be to linearize the temperature profiles between sections of the column. These sections would be from the bottom tray to the feed tray and from the feed tray to the overhead temperature. The overhead temperature should be set approximately $50\text{-}100^{0}$ higher than the condenser temperature.

The addition of steam to the distillation column invokes two major changes to a 2-phase distillation column. These are the enthalpy effect of cooling the steam on each tray and the presence of liquid water. If and when this program is updated for partially immiscible liquids, it is recommended that the program should use for the initial starting point the conditions of this program; assume that the two liquid phases are immiscible and include the enthalpy effect of the steam. Once convergence is reached,

the condition of partial immiscible liquids could be added and the column would be solved with this condition. This would greatly increase convergence and the speed of convergence by initially getting the temperature and liquid profiles in the correct range. Since three phase equilibrium relationships would require considerable comuputation time, having the temperature and liquid profiles as close to the answer as possible would greatly decrease the amount of computation time of the program. This procedure would eliminate some of the instabilities that could be encountered with three phase equilibrium relationships.

REFERENCES

Amundson, Neal R. and Arlene J. Pontinen (1958). Multicomponent Distillation Calculations on a Large Digital Computer. Industrial and Engineering Chemistry, 50(5), 730-736.

API (1976). Technical Data Book-Petroleum Refining (Third Edition). Washington, D.C.: American Petroleum Institute.

Block, Ulrich and Bernd Hegner (1976). Development and Application of a Simulation Model for Three-Phase Distillation. AICHE Journal, 22(3), 582-589.

Boston, Joseph F. (1970). Ph.D. Dissertation. Tulane University, New Orleans, 28-40.

Boston, J.F. and S.L. Sullivan (1972). An Improved Algorithm for Solving the Mass Balance Equations in Multistage Separation Processes. The Canadian Journal of Chemical Engineering, 50, 663-668.

Boston, J.F. and S.L. Sullivan (1974). A New Class of Solution Methods for Multicomponent, Multistage Separation Processes. The Canadian Journal of Chemical Engineering, 52, 52-63.

Boston, J.F. and H.I. Britt (1978). A Radically Different Formulation and Solution of the Single-Stage Flash Problem. Computers and Chemical Engineering, 2, 109-122.

Boston, Joseph F. (1980). Inside-out Algorithms for Multicomponent Separation Process Calculations. Proceedings of Symposium on Computer Applications to Chemical Engineering Process Design and Simulation, American Chemical Society, 135-151.

Cecchetti, Ralph, R.H. Johnston, J.L. Niedzwiecki and C.D. Holland (1963). Pipestill Products Verify These Computer Estimates. Hydrocarbon Processing and Petroleum Refiner, 42(9), 159-164.

Dean, John A., Editor (1985). Lange's Handbook of Chemistry, (Thirteenth Edition). New York: McGraw Hill Book Company.

Dreisbach, Robert R. (1952). Pressure-Volume-Temperature Relationships of Organic Compounds (Third Edition). Sandusky, Ohio: Handbook Publishers, Inc.

Erbar, John H. (1974). GPA*K: Calculating Phase Equilibria for Hydrocarbon and Selected Non-hydrocarbon Systems. Tulsa, OK; Gas Processors Association.

Erbar, John H. (1983). Phillips Fractionation Workshop. Bartlesville, OK: Phillips Petroleum Company.

Erbar, John H. (1987). MAXISIM: A Dual Purpose Complete Process Design Simulation System/Simple Equilibrium Calculation Program. Stillwater, OK: SCI.

Friday, John R. and Richard D. Smith (1964). An Analysis of the Equilibrium Stage Separations Problem-Formulation and Convergence. AICHE Journal. 10(5), 698-706.

Gary, James H. and Glenn E. Handwerk (1984). Petroleum Refining (Second Edition). New York: Marcel Dekker, Inc.

Gentry, James W. (1970). Solution of Block-Band Matrices. Industrial Engineering and Chemistry Fundamentals, 9(4), 671-673.

Gentry, James W. (August, 1970). An Improved Method for Numerical Solution of Distillation Processes. The Canadian Journal of Chemical Engineering, 48, 451-455.

Goldstein, Robert P. and Robert B. Stanfield (1970). Flexible Method for the Solution of Distillation Design Problems Using the Newton-Raphson Technique. Industrial and Engineering Chemical Process Design and Development, 9(1), 78-84.

GPSA (1981). Engineering Data Book (Ninth Edition, Fifth Revision). Tulsa, OK: Gas Processors Suppliers Association.

Green, Don W. (1984). Perry's Chemical Engineers Handbook (Sixth Edition). New York: McGraw Hill Book Company.

Hekim, Y., C.W. Orrick and J.H. Erbar (1973). Distillation Calculations Involving Three Phase Systems. For Presentation at International Congress Use of Electronic Computers in Chemical Engineering, April 1973.

Hess, R.L., C.D. Holland, R. McDaniel and N.J. Tetlow (1977). Solve More Distillation Problems. Hydrocarbon Processing and Petroleum Refiner, 56(1), 93-102.

Hidalgo, R., A. Correa, A. Gomez and J.D. Seader (1980). An Optimal Arrangement of Simultaneous Linearized

Equations of General Systems of Interlinked, Multistaged Separators. AICHE Journal, 26(4), 585-593.

Holland, Charles D. (1981). Fundamentals of Multicomponent Distillation. New York: McGraw Hill Book Company.

Hsu, Hsien and J.W. Tierney (1981). Exact Calculation Methods for Systems of Interlinked Columns. AICHE Journal, 27(5), 733-739.

Ishii, Yoshikazu and Fred D. Otto (1973). A General Algorithm for Multistage Multicomponent Separation Calculations. The Canadian Journal of Chemical Engineering, 51(10), 601-606.

Jordan, T. Earl (1954). Vapor Pressure of Organic Compounds. New York: Interscience Publishers, Inc.

Kinoshita, Masahiro, Iori Hashimoto and Takeichiro Takamatsu (1983). A Simulation Procedure For Multicomponent Distillation Column Within Which Three Phases of Vapor and Two Partially Immiscible Liquids Are Present. Journal of Chemical Engineering of Japan, 16(6), 513-516.

Kuno, Minoru (1981). Enthalpy Method for converging Distillation Calculations. Industial and Engineering Chemical Process Design and Development, 23(3), 166-168.

Maddox, Robert N. and John H. Erbar (1982). Gas Conditioning and Processing, Volume 3: Advanced Techniques and Applications. Norman, OK: Campbell Petroleum Series.

Naphtali, Leonard M. and Donald P. Sandholm (1971). Multicomponent Separation Calculations by Linearization. AICHE Journal,17(1), 148-153.

Newman, John (1963). Temperature Computed for Distillation. Hydrocarbon Processing and Petroleum Refiner, 42(4), 141-144.

Newman, John (August 1967). Numerical Solution of Coupled, Ordinary Differential Equations. (URCL-17739), Lawrence Radiation Laboratory, University of California, Berkeley.

Newman, John (1968). Numerical Solution of Coupled, Ordinary Differential Equations. Industrial and Engineering Chemistry Fundamentals, 7(13), 514-517.

Niedzwiecki, J.L., R.D. Springer and R.G. Wolfe (1980). Multi-component Distillation In the Presence of Free Water. Chemical Engineering Progress, 76(4), 57-58.

Noel, Henry M. (1959). Petroleum Refinery Manual. New York: Reinhold Publishing Corporation.

Ross, Bradley A. (1980). Ph.D. Dissertation. University of New Hampshire.

Russell, R.A. (1980). Chemical Process Simulation-The GMB System. Computers and Chemical Engineering, 4, 167-190.

Russell, R.A. (1983). A Flexible and Reliable Method Solves Single-Tower and Crude Distillation Column Problems. Chemical Engineering, 90(21), 53-59.

Schuil, J.A. and K.K. Bool (1985). Three-Phase Flash and Distillation. Computers and Chemical Engineering, 9(3), 295-300.

Seader, J.D. and Ernest J. Henley (1981). Equilibrium-Stage Operations in Chemical Engineering. New York: John Wiley and Sons.

Stadtherr, Mark A. and Michael A. Malachowski (1982). On Efficient Solution of Complex Systems of Interlinked Multistage Separators. Computers and Chemical Engineering, 6(2), 121-130.

Tomich, John F. (1970). A New Simulation Method for Equilibrium Stage Processes. AICHE Journal, 16(2), 229-232.

Van Wylen, Gordan J. and Richard E. Sonntag (1978). Fundamentals of Classical Thermodynamics. New York: John Wiley and Sons.

Venkataraman, S. and A. Lucia (1988). Solving Distillation Problems by Newton-Like Methods. Computers and Chemical Engineering, 12(1), 55-69.

Vickery, D.J., J.J. Ferrari and R. Taylor (1988) An "Efficient" Continuation Method for the Solution of Difficult Equilibrium Stage Separation Process Problems. Computers and Chemical Engineering, 12(1), 55-69.

Wang, J.C. and G.E. Henke (1966). Tridiagonal Matrix for Distillation. Hydrocarbon Processing, 45(8), 155-162.

Wasek, K. and J. Socha (1980). Steam Stripper Calculation by Simplified Newton-Raphson Technique. Chemical Engineering Science, 35(10), 623-626.

Watkins, R.N. (1979). Petroleum Refinery Distillation (Second Edition). Houston, TX: Gulf Publishing Company.

APPENDICES

APPENDIX A

SIMULATION CODE

```
C-------------------------------------------------------
C
C    CRUDE DISTILLATION PROGRAM
C
C        CREATED BY:  SUSAN L. PARKER
C
C-------------------------------------------------------
C
C
C               NOMENCLATURE SECTION
C
C-------------------------------------------------------
C
C   VARIABLES
C    ALl            ASSUMED LIQUID BOTTOMS RATE (ALl=0.0)
C    ALNS           ESTIMATED REFLUX RATE
C    API            DENSITY MEASUREMENT OF A COMPONENT
C    COF[A-E]       COEFFICIENTS USED IN DETERMININA A HYPO-
C                     THETICAL MOLECULAR WEIGHT
C    DETERM         MATRIX DETERMINATE
C    DLLIM          MAXIMUM CORRECTION FOR LIQUID FLOW
C    DTLIM          MAXIMUM CORRECTION FOR TRAY TEMPERATURE
C    EQ             TEMPORARY STORAGE FOR K-VALUE AT TRAY JG
C    HL,HLU         TEMPORARY STORAGE FOR COMPONENT LIQUID ENTHALPY
C    HV,HVB         TEMPORARY STORAGE FOR COMPONENT VAPOR ENTHALPY
C    ITERAT         ITERATION COUNTER
C    IPUMP          NO. OF PUMPAROUND STREAMS
C    TITLE          STORAGE FOR TITLE OF RUN
C    JCOTYP         TYPE OF CONDENSER
C    JG             ITERATION COUNTER FOR STAGE NUMBER
C    NC             NUMBER OF COMPONENTS
C    NEGV           CHECK FOR NEGATIVE VAPOR FLOW
C    NDRAW          NO. OF SIDE DRAWS
C    NSDS           NO. OF SIDE STRIPPERS
C    NMXCMP         MAXIMUM NUMBER OF COMPONENTS ALLOWED
C    NMXPRE         MAXIMUM NUMBER OF PURE COMPONENTS IN DATA BASE
C    NP1            NC+1
C    NP2            NC+2
C    NP2T           N-NOVAP IN BAND
C    NP3            NC+3
C    NS             NUMBER OF STAGES
C    NSMAX          MAXIMUM NUMBER OF STAGES ALLOWED
C    NSM1           NS-1
C    N2P1           (2*NP3)+1
C    N2P31          (2*N)+1
C    SUMERR         MAXIMUM ALLOWABLE ERROR IN COMPOSITION
C    REBT           TYPE OF REBOILER, NONE OR PARTIAL
C    TB             TEMPERATURE OF TRAY BELOW
C    TU             TEMPERATURE OF TRAY ABOVE
C    T1             ESTIMATED REBOILER TEMPERATURE
C    TNS            ESTIMATED CONDENSER TEMPERATURE
C
C       ONE-DIMENSIONAL ARRAYS
C   VARIABLE    DIMENSION
C    AK,AA          NC        K-VALUE COMPUTATION VARIABLES
C    AH,AAH         NC        ENTHALPY COMPUTATION VARIABLES
C    AL             NS        ASSUMED LIQUID RATE
C    BK,BB          NC        K-VALUE COMPUTATION VARIABLES
C    BH,BBH         NC        ENTHALPY COMPUTATION VARIABLES
C    BP             NC        NORMAL BOILING POINT OF HYPOTHETICAL
C    CB,QQ          NC        TEMPORARY STORAGE FOR TRAY COMPOSITIONS
C    CK,CC          NC        K-VALUE COMPUTATION VARIABLES
C    CH,CCH         NC        ENTHALPY COMPUTATION VARIABLES
```

```
C    DH,DDH          NC        ENTHALPY COMPUTATION VARIABLES
C    DELTAH,DELH     NC        ENTHALPY COMPUTATION VARIABLES
C    DEQ             NC        TEMPORARY STORAGE FOR K-VALUE DERIVATIVE
C    DHB,DHJG        NC        TEMPORARY STORAGE FOR ENTHALPY
C                                DERIVATIVE
C    EH,EEH          NC        ENTHALPY COMPUTATION VARIABLES
C    EQ              NC        TEMPORARY STORAGE FOR K-VALUE
C    ERR             NSMAX     FUNCTION USED TO CALCULATE COMPOSITIONS
C    F               NS        FEED RATE TO TRAY
C    FH,FFH          NC        ENTHALPY COMPUTATION VARIABLES
C    G               NC+3      TRAY FORCING STORAGE FUNCTION
C    GSG             NC        COMPONENT SPECIFIC GRAVITY
C    HF              NS        ENTHALPY OF FEED TO TRAY
C    HFP             10        COOLING OF PUMPAROUND STREAM
C    HVB             NC        TEMPORARY STORAGE FOR VAPOR ENTHALPY
C    HLU             NC        TEMPORARY STORAGE FOR LIQUID ENTHALPY
C    ID              NC        STORAGE VARIABLE FOR COMPONENT ID NUMBER
C    INN             K=1,3     PRODUCT OPTIONS
C                                INN(1)-FIXED BOTTOMS RATE
C                                INN(2)-FIXED OVERHEAD RATE
C                                INN(3)-NUMBER OF CONSTANT MOLAL
C                                       OVERFLOW ITERATIONS
C    IDRAW           10        LOCATION OF SIDE DRAW
C    IFTRAY          NS        LOCATION OF FEED
C    ISFTRAY         6         LOCATION OF STRIPPER
C    IPTRAY          10        LOCATION OF PUMPAROUND DRAW
C    IPRTRN          10        LOCATION OF PUMPAROUND RETURN
C    NSS             6         NUMBER OF TRAYS IN STRIPPER
C    P               NS        EFFECTIVE PRESSURE ON TRAY
C    PT              NS        TOTAL PRESSURE ON TRAY
C    PAL             6         LIQUID RATE OF PUMPAROUND STREAM
C    PS              6         STRIPPERS' PRESSURE
C    SAL             60        STRIPPERS' ASSUMED LIQUID RATES
C    SG,SPG          NC        SPECIFIC GRAVITY STORAGE
C    SHF             6         ENTHALPY OF FEED TO STRIPPER
C    SL,SV           NS        TRAY LIQUID,VAPOR DRAW RATES
C    SR              6         STEAM FED TO STRIPPERS
C    SVX             60        STRIPPERS' STEAM RATES
C    ST              60        STRIPPERS' TRAY TEMPERATURES
C    T               NS        TRAY TEMPERATURE
C    TFEED           NS        TEMPERATURE OF FEED TO TRAY
C    TS              6         TEMPERATURE OF STEAM FED TO STRIPPERS
C    V               NS        TRAY VAPOR RATE
C    VSX             60        STRIPPERS' TRAY VAPOR RATES
C    VS              NS        TRAY STEAM RATES
C    WMOL,WMW        NC        MOLECULAR WEIGHT COMPUTATION VARIABLES
C    WLF             NS        L/F FRACTION OF FEED TO TRAY
C
C       TWO-DIMENSIONAL ARRAYS
C   VARIABLE    DIMENSION
C    A          NC+3,NC+3      STORAGE FOR COMPOSITION ON TRAY
C                                BELOW
C    B          NC+3,NC+3      STORAGE FOR COMPOSITION ON TRAY
C                                IN QUESTION
C    C          NC+3,NC+3      TRAY MOLAR COMPOSITIONS
C    D          NC+3,(2*NC)+7  TRAY OBJECTIVE FUNCTION;MATRIX SOLN
C    EQDT       NC,NS          FUNCTION FOR K-VALUE DERIVATIVE
C    EQUIL      NC,NS          FUNCTION FOR K-VALUE
C    FX         NC,NS          COMPONENT FEED RATE TO TRAY
C    HHL        NC,NS          FUNCTION FOR LIQUID COMPONENT ENTHPY
C    HHV        NC,NS          FUNCTION FOR VAPOR COMPONENT ENTHLPY
C    Q          NC,NS          TEMP STORAGE FOR TRAY MOLAR COMPS
C    SAVE       NC,NS          TEMP STORAGE FOR TRAY MOLAR COMPS
C    SC         NC,60          STORAGE FOR STRIPPERS' MOLAR
C                                 COMPOSITIONS
C    SFX        NC,60          STRIPPERS' COMPONENT FEED RATE TO TRAY
```

```
C     X           NC+3,NC+3       TRAY COMPOSTION STORAGE
C     Y           NC+3,NC+3       TRAY COMPOSTION STORAGE
C
C        THREE-DIMENSIONAL ARRAYS
C    VARIABLE     DIMENSION
C     E           NC+3,NC+4,NS    TRAY JACOBIAN STORAGE ARRAY
C
C    SUBROUTINES
C     BAND(J)                 SETS UP MATRIX TO BE SOLVED
C     BUBBLE(J)               CALCULATES BUBBLE POINT
C     DEWPT(JG)               CALCULATES DEW POINT
C     KVALUE(I,T)             CALCULATES K-VALUES
C     NORM(FTOT)              NORMALIZES MOLAR COMPOSITIONS
C     SOLVER(K,L,DETERM)      SOLVES THE MATRIX
C     VENHDT(I,T)             CALCULATES VAPOR COMP ENTHALPIES
C     VENTH(I,T)              CALCULATES LIQUID COMP ENTH DERIVATIVES
C     VLNTH(I,T)              CALCULATES LIQUID COMP ENTHALPIES
C     VLNDT(I,T)              CALCULATES VAPOR COMP ENTH DERIVATIVES
C
C-------------------------------------------------------------
C
C       BLOCK DATA SECTION
C
C                   CONTAINS ALL NECESSARY DATA
C                    FOR PURE COMPONENT DATA BASE
C
C-------------------------------------------------------------
      BLOCK DATA DIST
      COMMON/VAR/AA(61),BB(61),CC(61),WMW(61)
      COMMON/VAR2/AAH(61),BBH(61),CCH(61)
      COMMON/VAR3/DDH(61),EEH(61),FFH(61),DELHH(61),GSPG(61)
C
C  JORDAN'S VAPOR PRESSURES OF ORGANIC COMPOUNDS
C    (LANGE'S FOR COMPONENTS 1,13-20,24-25,28-29,32,34-35,
C      46-51,56-58; AND CHEM PROCESSES FOR 61)
C          K=(LOG(A-(B/(T+C)))/PTOT
C
      DATA AA/5.82438,6.611847,6.80266,6.82973,6.74804,6.83029,
     1 6.80380,6.85221,6.73812,6.87773,6.90319,6.92374,6.93893,
     2 6.94365,6.97220,6.99795,7.00756,7.01300,7.02359,7.02867,
     3 7.0143,6.74756,6.81960,6.79290,6.88468,6.86952,6.84134,
     4 6.84999,6.84424,6.87540,6.90575,6.84637,6.82612,
     5 6.96659,6.85770,6.88673,6.86280,6.84498,6.82689,6.89745,
     6 6.95334,7.00289,7.00659,6.9909,6.95366,6.49457,7.28228,
     7 6.69422,9.81066,6.99392,6.69144,6.83907,6.84884,6.83712,
     8 6.8700,6.90187,5.7137,6.99383,6.86472,6.87041,8.10765/
C
      DATA BB/67.5078,389.93,656.40,813.20,882.80,945.90,1027.25,
     1 1064.63,950.84,1171.530,1268.586,1355.126,1431.82,1495.17,
     2 1569.57,1639.27,1690.67,1740.88,1789.95,1830.51,1865.1,
     3 585.00,785.00,908.80,967.32,960.80,923.20,930.546,1044.01,
     4 1069.466,1083.987,1039.69,1013.474,1124.33,1148.62,1124.162,
     5 1186.059,1203.526,1272.864,1206.350,1343.943,1477.519,
     6 1460.498,1453.840,1421.914,255.680,999.900,291.743,
     7 1347.786,768.130,319.013,1135.410,1152.368,1273.594,
     8 1315.503,1258.345,458.06,1014.117,1286.60,1384.036,1750.286/
C
      DATA CC/275.700,266.00,256.00,248.00,2*240.00,234.00,232.00,
     1 237.00,224.366,216.954,209.517,202.01,193.86,187.70,
     2 181.84,174.22,167.72,161.38,154.45,149.20,255.00,247.00,
     3 238.54,237.87,2*240.00,238.854,233.50,230.786,232.965,
     4 236.65,236.816,239.63,225.35,231.361,226.040,222.863,
     5 221.630,220.237,219.377,214.024,214.889,215.367,212.931,
     6 266.550,237.190,267.99,743.00,249.09,266.697,226.572,
     7 227.129,215.072,214.157,219.30,196.07,242.274,219.50,
     8 215.128,235.0/
```

```
C
C  LANGE'S HANDBOOK OF CHEMISTRY
C
      DATA WMW/2.016,16.04,30.07,44.10,58.12,58.12,3*72.15,
     1 86.18,100.21,114.23,128.26,142.29,156.31,170.34,184.37,
     2 198.40,212.42,226.45,240.41,28.05,42.08,4*56.10,54.09,
     3 6*70.14,84.16,70.13,2*84.16,98.19,78.11,92.14,4*106.17,
     4 28.01,32.00,28.01,44.01,34.08,64.07,4*86.18,98.90,
     5 40.06,54.09,98.18,112.22,18.02/
C
C  API:  TECHNICAL DATA BOOK--PETROLEUM REFINING
C          HL=A+(B*T)+(C*T**2)+(D*T**3)+(E*T**4)+(F*T**5)
C               HL[=]BTU/LB  T[=]RANKINE
C
      DATA AAH/12.32674,58.40160,163.05960,165.72380,162.08110,
     1 164.44400,169.01630,173.46090,145.60320,133.19390,
     2 134.12590,130.57280,126.71600,118.42310,156.57930,
     3 152.44400,151.999100,150.25060,148.84100,146.85880,
     4 144.59410,173.77850,193.22930,187.76640,210.45030,
     5 187.24240,179.26680,208.81850,175.27340,5*0.0,167.42860,
     6 229.11130,203.57300,209.80430,191.58840,225.05180,4*0.0,
     7 193.42630,-0.93401,-0.98176,-0.97557,4.77805,-0.61782,
     8 1.39433,152.96710,148.17180,166.76920,0.0,162.87650,
     9 229.26380,197.38100,196.95860,173.54180,-2.46342/
C     A,89*0.0/
C
      DATA BBH/3.199617,0.571700,0.264878,0.172601,0.046682,
     1 0.098571,-0.031504,-0.002795,0.004372,0.229107,0.182090,
     2 0.173084,0.169056,0.203347,-0.023843,-0.018522,-0.022933,
     3 -0.022048,-0.024114,-0.022825,-0.023563,0.144963,
     4 0.030810,-0.018519,-0.042795,0.037032,0.033009,-0.100603,
     5 -0.006874,5*0.0,-0.004262,-0.174553,-0.163500,-0.149848,
     6 -0.168390,-0.122662,4*0.0,-0.093633,0.255204,0.227486,
     7 0.256524,0.114433,0.238575,0.110263,0.041484,0.095013,
     8 -0.119500,0.0,-0.007807,0.033745,0.039560,-0.152454,
     9 -0.084958,0.457392/
C     A,89*0.0/
C
      DATA CCH/3.927862,-2.943122,-0.250140,0.940410,3.348013,
     1 2.691795,4.698836,4.400733,4.064654,-0.815691,0.347292,
     2 0.488101,0.581255,-0.349035,4.607729,4.538933,4.595173,
     3 4.580788,4.607172,4.590237,4.599069,1.710121,3.512242,
     4 4.263451,4.034318,3.551222,3.782637,5.651872,4.210531,
     5 5*0.0,4.196656,4.878999,5.315238,4.572747,5.444843,
     6 4.310824,4*0.0,4.390639,-0.177935,-0.373050,-0.229112,
     7 1.011325,-0.244571,0.330290,3.116936,1.605403,6.005360,
     8 0.0,4.259363,3.715168,3.479813,5.279018,4.278138,
     9 -0.525117/
C     A,89*0.0/
C
      DATA DDH/-2.934520,4.231568,2.923341,2.155433,0.144230,
     1 0.518202,-0.982825,-0.862875,-0.276464,4.527826,3.218786,
     2 3.054008,2.926114,4.070565,-0.998387,-0.964642,-0.997582,
     3 -0.9919639,-1.007675,-1.000209,-1.006645,0.761974,
     4 -0.494661,-0.940582,-0.684280,-0.560440,-0.733312,
     5 -2.123463,-0.908305,5*0.0,-0.882105,-0.790213,-1.239759,
     6 -0.387392,-1.126886,-1.138140,4*0.0,-1.126299,0.158913,
     7 0.483017,0.222803,-0.264936,0.410673,0.089125,0.622252,
     8 2.399189,-1.780500,0.0,-0.904956,-1.062607,-0.756071,
     9 -1.232110,-0.283095,0.645939/
C     A,89*0.0/
C
      DATA EEH/10.900690,-15.267400,-12.860530,-10.709860,
     1 -3.164196,-4.201390,1.029852,0.817644,-2.174529,
     2 -25.231790,-18.366030,-17.365470,-16.558500,-23.064410,
     3 1.084149,1.013931,1.083507,1.071259,1.104474,1.089122,
```

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     4 1.10366,-4.503085,-0.226171,1.072240,0.134493,0.158471,
     5 0.697566,4.830541,1.003804,5*0.0,0.925317,-0.259001,
     6 1.465505,-1.791242,0.751131,1.494985,4*0.0,1.458215,
     7 -0.322032,-1.852433,-0.563256,0.347063,-1.301258,
     8 -0.773135,-6.385257,-16.146920,2.920860,0.0,0.959607,
     9 1.864623,0.840374,1.467753,-2.216911,-2.027592/
C    A,89*0.0/
C
      DATA FFH/-13.878670,19.452610,18.220570,15.927940,
     1 5.428928,6.560421,-0.294847,-0.197154,4.685030,
     2 47.480200,33.769380,31.248310,29.296090,42.968970,
     3 -0.331217,-0.296646,-0.330908,-0.325375,-0.341474,
     4 -0.333900,-0.340757,6.664928,1.125539,-0.349830,
     5 0.878860,0.444673,-0.174830,-4.738449,-0.315910,5*0.0,
     6 -0.270520,1.873384,-0.497681,3.793529,0.606023,
     7 -0.564766,4*0.0,-0.543200,0.158927,2.474881,0.455878,
     8 -0.131400,1.448520,1.292865,12.592750,32.727430,
     9 -1.344890,0.0,-0.284711,-1.435039,-0.265154,
     A -0.499779,4.555816,2.363096/
C    B,89*0.0/
C
C  LANGE'S HANDBOOK OF CHEMISTRY
C     KCAL/GMOLE
C
      DATA DELHH/0.216,1.953,3.517,4.487,2*5.352,3*6.160,
     1 6.896,7.575,8.225,8.820,9.388,9.920,10.430,10.910,
     2 11.380,11.820,12.240,12.640,3.237,4.400,5.238,0.0,5.580,
     3 0.0,5.420,6.020,6.240,6.230,6.094,5.750,6.287,6.760,
     4 6.524,6.950,7.160,7.440,7.352,7.930,8.800,8.690,8.600,
     5 8.500,1.333,1.630,1.444,0.0,4.463,5.955,6.643,6.711,6.287,
     6 6.519,7.430,4.450,5.820,7.715,8.200,9.717/
C    A,89*0.0/
C
C  SPECIFIC GRAVITIES  60F/60F
C     BASE DENSITY   WATER = 62.365 lbm/ft3
C          GPSA DATABOOK FOR 1-14,23-29,36-53,55,58,61
C          LANGE'S HANDBOOK OF CHEM FOR 15-21,35,54,56
C
      DATA GSPG/0.07107,0.3,0.3583,0.5075,0.5630,0.5842,0.6244,
     10.6311,0.5966,0.6640,0.6883,0.7070,0.7219,0.7342,0.741,
     20.751,0.757,0.765,0.770,0.774,0.775,0.835,0.5217,0.6013,
     30.6271,0.6101,0.6005,0.6272,0.6457,5*0.0,0.673,0.7504,
     40.7536,0.7834,0.7740,0.8845,0.8719,0.8847,0.8688,0.8658,
     50.8717,0.8094,1.1421,0.7894,0.8176,0.7871,1.397,0.6579,
     60.6689,0.649,0.6664,0.697,0.0,0.658,2*0.0,1.000/
C
      END
C
C NMAX IS THE MAX NO. OF COMPS THAT CAN BE UTILIZED
C
      PARAMETER (NMAX=50,NSMAX=50)
      CHARACTER*10 NAME(100),SNAME(NMAX)
      REAL P(NSMAX),PT(NSMAX),T(NSMAX),ALW(NSMAX),SLW(NSMAX)
      REAL SL(NSMAX),VS(NSMAX),SV(NSMAX),WLF(NSMAX)
      REAL SHF(6),SAL(60),ST(60),SVX(60),SC(NMAX,60)
      REAL VSX(60),TS(6),SR(6),ALS(6),PS(6)
      REAL SFX(NMAX,60),HFP(5),PSL(5),APL(6)
      INTEGER IPSTRAY(6)
      INTEGER NSS(6),IDRAW(10),IPTRAY(5),IPRTRN(5),IPUMP
      INTEGER ITERAT,ICHECK,NEGV,L1,ISTRAY(6),LNS,REBT
      DIMENSION F(NSMAX),HF(NSMAX),IFTRAY(NSMAX),TFEED(NSMAX),
     $FX(NMAX,NSMAX),CB(100),EQ2(100),DEQ2(100),
     $CB3(100),EQB3(100),DEQ3(100),HVB3(100),HLU3(100),
     $DHB(100),DHJG(100),Q(100,50),QQ(50),VOL(50)
C234567
      COMMON/VAR/AA(61),BB(61),CC(61),WMW(61)
```

```
      COMMON/VAR2/AAH(61),BBH(61),CCH(61)
      COMMON/VAR3/DDH(61),EEH(61),FFH(61),DELHH(61),GSPG(61)
      COMMON/VNAME/NAME
      COMMON/OPTION/INN(3)
      COMMON/BATCH/ID(NMAX),WMOL(NMAX)
      COMMON/KCALC/AK(NMAX),BK(NMAX),CK(NMAX),PK(NMAX),EQK(NMAX)
      COMMON/ENTH1/AH(NMAX),BH(NMAX),CH(NMAX)
      COMMON/ENTH2/DH(NMAX),EH(NMAX),FH(NMAX),DELH(NMAX)
      COMMON/GRAVIT/GSG(NMAX),BP(NMAX)
      COMMON/FEED/C(105,105),D(105,225)
      COMMON/FEED2/E(105,105,50)
      COMMON/MTRX1/X(105,105),Y(105,105)
      COMMON/MTRX2/A(105,105),B(105,105)
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
C
      CHARACTER*50 TITLE
C
C       NAME ABBREVIATIONS FOR EACH
C            COMPONENT
C
      DATA NAME/'H2','CH4','C2H6','C3H8','I-C4H10','N-C4H10',
     A'I-C5H12','N-C5H12','NEO-C5','N-C6H14','N-C7H16',
     B'N-C8H18','N-C9H20','N-C10H22','N-C11H24','N-C12H26',
     C'N-C13H28','N-C14H30','N-C15H32','N-C16H34',
     D'3N-C17H36','C2H4=','C3H6=','1-C4H8','C-2-C4H8',
     E'T-2-C5H8','I-C4H8','13-C4==','1-C5H10','C-2-C5=',
     F'T-2-C5=','2MT-1C4=','3MT-1C4=',
     G'2MT-2C4=','C6H12=','CYC-C5','MTCYC-C5','CYC-C6',
     H'MTCYC-C6','BZ','TOL','O-X','M-X','P-X','EB','N2',
     I'O2','CO','CO2','H2S','SO2','2-MT-C5',
     J'3-MT-C5','2,2DMTC4','2,3DMTC4','1-C7H14','C3H4==',
     K'1,2-C4==','ETCYC-C5','ETCYC-C6','H2O','FR1','FR2',
     L'FR3','FR4','FR5','FR6','FR7','FR8','FR9','FR10',
     M'FR11','FR12','FR13','FR14','FR15','FR16','FR17',
     N'FR18','FR19','FR20','FR21','FR22','FR23','FR24','FR25',
     O'FR26','FR27','FR28','FR29','FR30','FR31','FR32'
     P'FR33','FR34','FR35','FR36','FR37','FR38','FR39','FR40'/
      DATA NI,NO/5,6/
      DATA H2ODEN/62.365/
C      DATA NPROG,NDRAW,ICHECK,NEGV/4*0/
      DATA COFA,COFB/3.4660932,3.66593025E-8/
      DATA COFC,COFD/0.00000319803,0.010703069/
      DATA COFE/0.000510687/
C   INITIALIZE ALL ARRAYS TO ZERO
      DATA AK,BK,CK/50*0.,50*0.,50*0./
      DATA AH,BH,CH,DH/50*0.,50*0.,50*0.,50*0./
      DATA EH,FH,DELH/50*0.,50*0.,50*0./
      NH2O=0
      NMXPRE=61
C    STILL INITIALIZING ARRAYS
C
      DO 2 J=1,NSMAX
      F(J)=0.0
      HF(J)=0.0
      IFTRAY(J)=0
      TFEED(J)=0.0
      DO 2 I=1,NMAX
      FX(I,J)=0.0
    2 CONTINUE
C-------------------------------------------------------------
C  THE NEXT SECTION INPUTS DATA BY FILE
C-------------------------------------------------------------
      WRITE(6,2002)
```

```
 2002 FORMAT(2X,'DO YOU WISH TO INPUT DATA BY FILE? (Y,N)')
      READ(6,1035)IFILE
      IF(IFILE.EQ.'N') GO TO 2309
      WRITE(6,*)'ENTER FILE NAME:'
      READ(6,2004)IFLNAME
 2004 FORMAT(A10)
      OPEN (UNIT=8,FILE='SAMER.DAT',STATUS='OLD')
      READ(8,102)TITLE
      READ(8,2006)NC,NCO,NSTAR,NH2O
      DO 2010 IJ=1,NCO
      READ(8,2008)ID(IJ),AK(IJ),BK(IJ),CK(IJ)
      READ(8,2009)AH(IJ),BH(IJ),CH(IJ),DH(IJ)
      READ(8,2009)EH(IJ),FH(IJ),WMOL(IJ),DELH(IJ)
      READ(8,2009)GSG(IJ),BP(IJ)
      IK=ID(IJ)
      SNAME(IJ)=NAME(IK)
C
C      WRITE(6,2008)ID(IJ),AK(IJ),BK(IJ),CK(IJ)
C      WRITE(6,2009)AH(IJ),BH(IJ),CH(IJ),DH(IJ)
C      WRITE(6,2009)EH(IJ),FH(IJ),WMOL(IJ),DELH(IJ)
C      WRITE(6,2009)GSG(IJ),BP(IJ)
 2010 CONTINUE
 2006 FORMAT(4I5)
 2008 FORMAT(I4,3F14.4)
 2009 FORMAT(4F14.4)
      READ(8,105)NS,NF,LIM
      NSM1=NS-1
      READ(8,105)JCOTYP,NDRAW,NSDS
      READ(8,105)REBT
      READ(8,*)STEAM,TSTEAM
C      READ(8,*)AL(1),AL(NS)
      DO 5012 JJ=1,NS
      READ(8,*)T(JJ),AL(JJ)
 5012 CONTINUE
      DO 2012 IJ=1,NS
      READ(8,*)P(IJ)
 2012 CONTINUE
      DO 2014 JF=1,NF
      READ(8,*)IFTRAY(JF),FRAC,TFE
      J=IFTRAY(JF)
      WLF(J)=FRAC
      TFEED(J)=TFE
      DO 2116 I=1,NCO
      READ(8,*)FX(I,J)
 2116 CONTINUE
      DO 2125 I=1,NC
      IF(WLF(J).GE.1) GO TO 2122
      CALL VENTH(I,TFE)
      HF(J)=HF(J)+FX(I,J)*HHV(I)*(1.0-WLF(J))
 2122 CALL VLNTH(I,TFE)
      HF(J)=HF(J)+FX(I,J)*HHL(I)*WLF(J)
      F(J)=F(J)+FX(I,J)
 2125 CONTINUE
      WRITE(6,*)HF(J),F(J)
 2014 CONTINUE
      DO 2140 J=1,NDRAW
      READ(8,143)IDRAW(J)
      JD=IDRAW(J)
      READ(8,*)SL(JD),SV(JD)
 2140 CONTINUE
      DO 2161 JJ=1,NSDS
      READ(8,*)JS
      ISTRAY(JJ)=JS
      READ(8,*)SL(JS)
      READ(8,*)NSS(JJ)
      READ(8,*)SR(JJ),TS(JJ)
```

```
      FX(NH2O,JS+1)=SR(JJ)
      READ(8,*)ALS(JJ)
      READ(8,*)PS(JJ)
      READ(8,*)APL(JJ)
      READ(8,*)IPSTRAY(JJ)
 2161 CONTINUE
      READ(8,*)IPUMP
      DO 2459 J=1,IPUMP
      READ(8,*)IPTRAY(J),IPRTRN(J)
      READ(8,*)PSL(J)
      READ(8,*)HFP(J)
 2459 CONTINUE
      GO TO 2201
C-------------------------------------------------------
C
C   NORMAL INTERACTIVE INPUT
C
C-------------------------------------------------------
 2309 WRITE(6,101)
  101 FORMAT(2X,'ENTER TITLE :')
      READ(6,102) TITLE
  102 FORMAT(A50)
C
C
C   INPUT COMPONENTS
C
C
      FS=0.0
      IWATER=0
      NSTAR=0
      NB=0
      WRITE (NO,611)
  611 FORMAT (///,5X,'ID NUMBERS 62-150 ARE RESERVED FOR HYPO
     1THETICALS')
 1036 DO 5, J=1,NMAX
      NB=NB+1
      WRITE (NO,6) J
    6 FORMAT (/,5X,'ENTER ID NUMBER FOR COMP ',I2)
      READ (NI,7) ID(J)
      ILK=ID(J)
      SNAME(J)=NAME(ILK)
    7 FORMAT (I3)
      IF(ID(J).EQ.61) NH2O=J
      IF(ID(J).GT.61.AND.ID(J).LE.150) NSTAR=NSTAR+1
      IF(ID(J).LT.1.OR.ID(J).GT.150) GO TO 88
    5 CONTINUE
   88 NCO=NB-1
      WRITE (NO,1031)
 1031 FORMAT (5X,'YOU HAVE ENTERED THE FOLLOWING COMPONENTS')
      DO 1032, I=1,NCO
      WRITE (NO,1033) I,ID(I)
 1033 FORMAT (5X,I3,3X,I3)
 1032 CONTINUE
      WRITE (NO,1034)
 1034 FORMAT (/,5X,'DO YOU WISH TO CHANGE ANY? (Y,N)')
      READ (NI,1035) KID
 1035 FORMAT (A2)
      IF(KID.EQ.'Y') GO TO 1036
      IPURE=NCO-NSTAR
      IF(NH2O.NE.0) IPURE=IPURE-1
      DO 9 J=1,IPURE
      II=ID(J)
      AK(J)=AA(II)
      BK(J)=BB(II)
      CK(J)=CC(II)
      AH(J)=AAH(II)
```

```
      BH(J)=BBH(II)
      CH(J)=CCH(II)
      DH(J)=DDH(II)
      EH(J)=EEH(II)
      FH(J)=FFH(II)
      WMOL(J)=WMW(II)
      GSG(J)=GSPG(II)
      DELH(J)=DELHH(II)
    9 CONTINUE
      IF(NH2O.EQ.0) GO TO 2321
      AK(NH2O)=AA(61)
      BK(NH2O)=BB(61)
      CK(NH2O)=CC(61)
      AH(NH2O)=AAH(61)
      BH(NH2O)=BBH(61)
      CH(NH2O)=CCH(61)
      DH(NH2O)=DDH(61)
      EH(NH2O)=EEH(61)
      FH(NH2O)=FFH(61)
      WMOL(NH2O)=WMW(61)
      GSG(NH2O)=GSPG(61)
      DELH(NH2O)=DELHH(61)
 2321 IF(NSTAR.EQ.0) GO TO 1045
      WRITE (NO,1046)
 1046 FORMAT (/,5X,'FOR THE FOLLOWING HYPOTHETICALS YOU WILL
     1 NEED TO ENTER',/,5X,'THE NORMAL BOILING POINT, THE MOL
     2ECULAR WEIGHT,',/,5X,'AND THE SPECIFIC GRAVITY',/,
     35X,'IF YOU DO NOT KNOW THE VALUE OF THE MOLECULAR
     4 WEIGHT,'/,5X,'ENTER A -9 AND THE PROGRAM WILL
     5 CALCULATE IT'/)
      KBOP=NCO-NSTAR
      IF (NH2O.NE.0) KBOP=KBOP-1
      DO 1045, J=1,NSTAR
      KCAN=KBOP+J
 1166 WRITE (NO,704) J
  704 FORMAT (5X,'ENTER THE NORMAL BOILING POINT, THE MOLECU
     1LAR WEIGHT,',/,5X,'AND THE SPECIFIC GRAVITY FOR FRACTION '
     2,I2)
      READ (NI,*) BP(KCAN),WMOL(KCAN),GSG(KCAN)
C   ANTOINE'S COEFFICIENTS AND CALCULATIONS ARE FROM R. R.
C      DREISBACH, 'PVT RELATIONSHIPS OF ORGANIC COMPOUNDS'
C         (1952)
      IF(BP(KCAN).GT.356.0) GO TO 702
      BK(KCAN)=2.3025851*(1313.01+4.27937*BP(KCAN)+0.002649*
     1BP(KCAN)**2.0)
      GO TO 703
  702 BK(KCAN)=2.3025851*(1653.16+3.1744*BP(KCAN)+0.0042493*
     1BP(KCAN)**2.0)
  703 CK(KCAN)=382.0
      AK(KCAN)=2.687572+BK(KCAN)/(CK(KCAN)+BP(KCAN))
      API=141.5/GSG(KCAN)-131.5
      IF(WMOL(KCAN).LE.0.0) WMOL(KCAN)=EXP(COFA+COFB*BP(KCAN)
     1**2.0+COFC*API*BP(KCAN)+COFD*API+COFE*API**2.0)
      WRITE (NO,777) BP(KCAN),WMOL(KCAN),GSG(KCAN),
     1AK(KCAN),BK(KCAN),CK(KCAN)
  777 FORMAT (/,5X,'YOU HAVE ENTERED THE FOLLOWING PARAMETERS:',
     1/,7X,'NORMAL BOILING POINT = ',F12.4,/,7X,'MOLECULAR
     2 WEIGHT     = ',F12.4,/,7X,'SPECIFIC GRAVITY     = ',F12.4,/
     3,5X,'AND YOU HAVE CALCULATED THESE ANTIONE COEFFICIENTS A,
     4 B, & C:'/,4X,3(2X,F16.4))
      WRITE (NO,1164)
 1164 FORMAT (/,5X,'DO YOU WANT TO RE-ENTER THE DATA'/5X,'FOR
     1 THIS FRACTION ? (Y,N)')
      READ (NI,1165) KRE
 1165 FORMAT (A2)
      IF(KRE.EQ.'Y') GO TO 1166
```

```
 1045 CONTINUE

      WRITE(6,*) 'CHECK INPUT DATA'
      DO 3559 IJ=1,NCO
      WRITE(6,*)'DATA FOR COMPONENT',IJ
      WRITE(6,*)'KVALUE',AK(IJ),BK(IJ),CK(IJ)
      WRITE(6,*)'ENTH',AH(IJ),BH(IJ),CH(IJ),DH(IJ)
      WRITE(6,*)EH(IJ),FH(IJ),DELH(IJ),WMOL(IJ)
 3559 CONTINUE

    INPUT DATA


      IF(NH2O.EQ.0) THEN
      NC=NCO
      ELSE
      NC=NCO-1
      END IF
      WRITE(6,104)
  104 FORMAT(/,5X,'ENTER THE NUMBER OF STAGES, NUMBER OF FEEDS'/5X,
     $'AND THE MAXIMUM NUMBER OF ITERATIONS.')
      READ(6,105)NS,NF,LIM
      NSM1=NS-1
  105 FORMAT(3I5)
      WRITE(6,106)
  106 FORMAT(/,5X,'ENTER COND TYPE (0,1,2)=(NONE,PART,TOTAL),'
     $/,5X,'NUMBER OF SIDEDRAWS, AND SIDESTREAM STRIPPERS.')
      READ(6,105)JCOTYP,NDRAW,NSDS
      WRITE(6,171)
  171 FORMAT(/,5X,'ENTER REBOILER TYPE; 0=NONE,1=PART')
      READ(6,105)REBT
      IF(NH2O.EQ.0) GO TO 173
       WRITE(6,172)
  172 FORMAT(/,5X,'ENTER STEAM RATE TO BOTTOM, TEMPERATURE')
      READ(6,109)STEAM,TSTEAM
  173 WRITE(6,108)
  108 FORMAT(/,5X,'ENTER ESTIMATED BTMS. RATE AND REFLUX RATE')
      READ(6,109)AL(1),AL(NS)
  109 FORMAT(2F10.0)
      WRITE(6,111)
  111 FORMAT(/,5X,'ENTER REBOILER AND CONDENSER TEMP ESTIMATES, F')
      READ(6,109)T(1),T(NS)
      WRITE(6,5131)
 5131 FORMAT(/,5X,'DO YOU WISH TO ENTER OTHER TRAY TEMPS?')
      READ(6,113)IT
      IF(IT.EQ.'N') GO TO 5126
      DO 5127 J=2,NSM1
      WRITE(6,5128)J
 5128 FORMAT(/,5X,'ENTER TEMPERATURE FOR TRAY :',I2)
      READ(6,*)T(J)
 5127 CONTINUE
      GO TO 5132
 5126 TBMBT=(T(NS)-T(1))/FLOAT(NS-1)
      DO 5132 J=2,NSM1
      T(J)=T(1)+TBMBT*FLOAT(J-1)
 5132 CONTINUE
      WRITE(6,114)
  114 FORMAT(/,5X,'ENTER BOTTOM AND TOP PRESSURES IN MM HG')
      READ(6,109) P(1),P(NS)
      WRITE(6,*)P(1),P(NS)
      WRITE(6,115)
  115 FORMAT(/,5X,'DO YOU WISH TO ENTER INTERMEDIATE PRESSURES?')
      READ(6,113)IP
  113 FORMAT(A3)
      IF(IP.EQ.'N') GO TO 126
      DO 127 I=2,NSM1
```

```
      WRITE(6,128)I
  128 FORMAT(/,5X,'ENTER PRESSURE FOR TRAY : ',I2)
      READ(6,*)P(I)
  127 CONTINUE
      GO TO 129
  126 SMP=(P(NS)-P(1))/FLOAT(NS-1)
      DO 131 I=2,NSM1
      P(I)=P(1)+SMP*FLOAT(I-1)
      WRITE(6,*)P(I)
  131 CONTINUE
  129 DO 130 JF=1,NF
      WRITE(6,117)JF
  117 FORMAT(/,5X,'ENTER FEED TRAY, L/F, TEMP OF FEED'
     $,I5)
      READ(6,118)IFTRAY(JF),FRAC,TFE
  118 FORMAT(I3,2F10.0)
      J=IFTRAY(JF)
      WLF(J)=FRAC
      TFEED(J)=TFE
      DO 116 I=1,NCO
      WRITE(6,119)I
  119 FORMAT(/,5X,'ENTER FEED RATE FOR COMPONENT #  ',I3)
      READ(6,*) FX(I,J)
  116 CONTINUE
C
C  AS A BOOK-KEEPING MEASURE THE FEED ENTHALPY AND THE
C  FEED RATE DO NOT INCLUDE WATER
C  NC REPRESENTS THE NUMBER OF COMPONENTS EXCLUDING WATER
C
      DO 125 I=1,NC
      IF (WLF(J).GE.1) GOTO 122
      CALL VENTH(I,TFEED(J))
      HF(J)=HF(J)+FX(I,J)*HHV(I)*(1-WLF(J))
  122 CALL VLNTH(I,TFEED(J))
      HF(J)=HF(J)+FX(I,J)*HHL(I)*WLF(J)
  123 CONTINUE
      F(J)=F(J)+FX(I,J)
  125 CONTINUE
      WRITE(6,*)HF(J),F(J)
C      IF(IFEZ.EQ.0)THEN
C      IF(NH2O.EQ.0) GO TO 1441
C      FLIQ=FLIQ-FX(NH2O,J)
C 1441 FLIQ=FLIQ+F(J)
C      WRITE(6,*)F(J),FLIQ
C      ELSE
C      END IF
  130 CONTINUE
      DO 140 J=1,NDRAW
      WRITE(6,142)
  142 FORMAT(/,5X,'ENTER SIDE DRAW TRAY : ')
      READ(6,143)IDRAW(J)
      JD=IDRAW(J)
  143 FORMAT(I4)
      WRITE(6,144)
  144 FORMAT(/,5X,'ENTER TRAY LIQUID AND TRAY VAPOR DRAW RATES: ')
      READ(6,*)SL(JD),SV(JD)
  140 CONTINUE
      DO 161 JJ=1,NSDS
      WRITE(6,162)
  162 FORMAT(/,5X,'ENTER TRAY NO. FOR SIDESTREAM STRIPPER')
      READ(6,143)JS
      ISTRAY(JJ)=JS
      WRITE(6,167)
  167 FORMAT(/,5X,'ENTER LIQUID RATE TO STRIPPER')
      READ(6,*)SL(JS)
      WRITE(6,169)
```

```
  169 FORMAT(/,5X,'ENTER NO. OF TRAYS IN STRIPPER')
      READ(6,143)NSS(JJ)
      WRITE(6,1773)
 1773 FORMAT(/,5X,'ENTER STEAM RATE AND TEMPERATURE')
      READ(6,109)SR(JJ),TS(JJ)
      FX(NH2O,JS+1)=SR(JJ)
      WRITE(6,174)
  174 FORMAT(/,5X,'ENTER ESTIMATED BOTTOMS RATE')
      READ(6,109)ALS(JJ)
      WRITE(6,178)
  178 FORMAT(/,5X,'ENTER PRESSURE IN STRIPPER')
      READ(6,*)PS(JJ)
  161 CONTINUE
C   PUMPAROUND INPUT
      WRITE(6,2442)
 2442 FORMAT(/,5X,'ENTER THE NUMBER OF PUMPAROUNDS')
      READ(6,2443)IPUMP
 2443 FORMAT(I4)
      DO 2452 JP=1,IPUMP
      WRITE(6,2446)
 2446 FORMAT(/,5X,'ENTER PUMPAROUND DRAW TRAY')
      READ(6,2443)IPTRAY(JP)
      JC=IPTRAY(JP)
      WRITE(6,2444)
 2444 FORMAT(/,5X,'ENTER LIQUID FLOW RATE')
      READ(6,2445)PSL(JP)
 2445 FORMAT(F14.0)
      WRITE(6,2448)
 2448 FORMAT(/,5X,'ENTER COOLING RATE, BTU/HR')
      READ(6,2445)HFP(JP)
      WRITE(6,2450)
 2450 FORMAT(/,5X,'ENTER RETURN TRAY NUMBER')
      READ(6,2443)IPRTRN(JP)
      JR=IPRTRN(JP)
      F(JR)=PSL(JP)
      SL(JC)=PSL(JP)
 2452 CONTINUE
C
C
C
C-----------------------------------------------------
C
C   THIS SECTION WILL OUTPUT THE INPUT VARIABLES TO
C     A FILE SO THAT THEY MAY BE USED AGAIN WITHOUT
C     HAVING TO RE-ENTER
C
C-----------------------------------------------------
      WRITE(6,2202)
 2202 FORMAT(2X,'DO YOU WISH TO SAVE INPUT DATA? (Y,N)')
      READ(6,1035)IFILE
      IF(IFILE.EQ.'N') GO TO 2201
      WRITE(6,*)'ENTER FILE NAME:'
      READ(6,2004)IFLNAME
      OPEN (UNIT=9,FILE='SAMER.DAT',STATUS='NEW')
      WRITE(9,102)TITLE
      WRITE(9,2006)NC,NCO,NSTAR,NH2O
      DO 2210 IJ=1,NCO
      WRITE(9,2008)ID(IJ),AK(IJ),BK(IJ),CK(IJ)
      WRITE(9,2009)AH(IJ),BH(IJ),CH(IJ),DH(IJ)
      WRITE(9,2009)EH(IJ),FH(IJ),WMOL(IJ),DELH(IJ)
      WRITE(9,2009)GSG(IJ),BP(IJ)
 2210 CONTINUE
      WRITE(9,105)NS,NF,LIM
      NSM1=NS-1
      WRITE(9,105)JCOTYP,NDRAW,NSDS
      WRITE(9,105)REBT
```

```
      WRITE(9,*)STEAM,TSTEAM
      WRITE(9,*)AL(1),AL(NS)
      DO 5212 JJ=1,NS
      WRITE(9,*)T(JJ)
 5212 CONTINUE
      DO 2212 IJ=1,NS
      WRITE(9,*)P(IJ)
 2212 CONTINUE
      DO 2214 JF=1,NF
      WRITE(9,*)IFTRAY(JF),WLF(J),TFE
      J=IFTRAY(JF)
      WLF(J)=FRAC
      TFEED(J)=TFE
      DO 2216 I=1,NCO
      WRITE(9,*) FX(I,J)
 2216 CONTINUE
 2214 CONTINUE
      DO 2240 J=1,NDRAW
      JD=IDRAW(J)
      WRITE(9,143)JD
      WRITE(9,*)SL(JD),SV(JD)
 2240 CONTINUE
      DO 2469 JJ=1,NSDS
      READ(9,*)JS
      ISTRAY(JJ)=JS
      READ(9,*)SL(JS)
      READ(9,*)NSS(JJ)
      READ(9,*)SR(JJ),TS(JJ)
      FX(NH2O,JS+1)=SR(JJ)
      READ(9,*)ALS(JJ)
      READ(9,*)PS(JJ)
 2469 CONTINUE
C
C
      WRITE(9,*)IPUMP
      DO 2466 J=1,IPUMP
      WRITE(9,*)IPTRAY(J),IPRTRN(J)
      WRITE(9,*)PSL(J)
      WRITE(9,*)HFP(J)
 2466 CONTINUE
C
C
C
 2201 DO 709 IJ=1,NS
      PT(IJ)=P(IJ)
  709 CONTINUE
C
C
C   COLUMN SPECIFICATIONS?????????
C
C
      SUMERR=0.001
      SUBCOL=0.0
      DLLIM=0.2
      DTLIM=10.0
      CHECK=0.0
      INN(3)=0
      NP1=NC+1
      NP2=NC+2
      NP3=NC+3
C
C
C GENERATE INITIAL PROFILES
C
C
      NSM1=NS-1
```

```
      DO 344 JJ=NSM1,2,-1
      WRITE(6,*)F(JJ),AL(JJ+1)
      AL(JJ)=AL(JJ+1)+F(JJ)*WLF(JJ)-SL(JJ)
  344 CONTINUE
      ENS=FLOAT(NS*(NS+1))/2.0
      TBMTD=(T(NS)-T(1))/FLOAT(NS-1)
      ELBMR=FLIQ/ENS
      DO 444, JA=2,NSM1
      T(JA)=T(1)+TBMTD*FLOAT(JA-1)
      AL(JA)=AL(NS)+ELBMR*FLOAT(NS-JA)
      WRITE(6,*)T(JA),AL(JA)
  444 CONTINUE
C-------------------------------------------------------
C
C   INITIALIZE STRIPPER PROFILES
C
C-------------------------------------------------------
      IF(NSDS.EQ.0) GO TO 446
      DO 183 JJ=1,NSDS
      ND=NSS(JJ)
      NO=(JJ-1)*10
      NDM1=ND-1
      NOP1=NO+1
      NOND=NO+ND
      JS=ISTRAY(JJ)
      SAL(NOND)=SL(JS)
      SAL(NOP1)=ALS(JJ)
      DO 182 JL=NDM1,2,-1
      ICNT=NO+JL
      SAL(ICNT)=SAL(NOND)
  182 CONTINUE
C
      DO 184 JL=2,NDM1
      ICNT=NO+JL
      VSX(ICNT)=SR(JJ)
  184 CONTINUE
      VSX(NOP1)=SR(JJ)
      VSX(NOND)=SR(JJ)
      FX(NH2O,JS+1)=SR(JJ)
  183 CONTINUE
C  INITIAL LIQUID WATER RATES ARE ASSUMED TO BE ZERO
C
C  LIQUID DRAWS ARE ASSUMED TO BE WATER FREE
C  THIS MAY BE CHANGED BY GIVING THE WATER LIQUID
C  DRAW RATE
C
C
C  INITIALLY ASSUME THAT ALL THE STEAM PASSES THROUGH THE
C  COLUMN UNCONDENSED i.e. VS(NS)=VS(1)
C
  446 DO 341 I=1,NS
      ALW(I)=0.0
      SLW(I)=0.0
      VS(J)=0.0
  341 CONTINUE
C
C
C BEGIN ITERATIONS
C
C      IF(ICHECK.GE.1000) GO TO 1111
 9999 CONTINUE
   12 ITERAT=ITERAT+1
C-------------------------------------------------------------
C      PUMPAROUNDS
C
C   FOR THE FIRST ITERATION THE LIQUID SIDE DRAW OF EACH
```

```
C       PUMPAROUND WILL NOT AFFECT THE COMPOSITION PROFILE
C       OR SOLUTION AT ALL OTHER TIMES THAT ARE TAKEN INTO
C       AFFECT
C-------------------------------------------------------------
      IF(ITERAT.LE.1) GO TO 2456
      DO 2457 J=1,IPUMP
      JP=IPTRAY(J)
      JC=IPRTRN(J)
      SL(JP)=PSL(J)
      HLP=0.0
      DO 2458 JL=1,NC
      CALL VLNTH(JL,T(JP))
      HLP=HLP+C(JL,JP)*HHL(JL)*SL(JP)
      FX(JL,JC)=C(JL,JP)*SL(JP)
 2458 CONTINUE
      F(JC)=PSL(J)
      HF(JC)=HLP-HFP(J)
 2457 CONTINUE
 2456 CONTINUE
C
C  IF WATER IS CONDENSED OUT IN SOME INTERMEDIATE
C ITERATION IT WILL NOT AFFECT SUBSEQUENT ITERATIONS
C
      IF(NH2O.EQ.0) GO TO 342
      VS(1)=FX(NH2O,1)+STEAM
      DO 342 J=2,NS
      VS(J)=VS(J-1)+FX(NH2O,J)
  342 CONTINUE
      DO 1280, J=1,NS
      V(J)=0.0
 1280 CONTINUE
C
C
C
C CALCULATE ASSUMED VAPOR RATES
C
      NEGV=0
      V(1)=AL(2)-AL(1)-SL(1)-SV(1)+F(1)
      IF(V(1).GT.0.0) GO TO 11
      V(1)=1.0
      NEGV=1
   11 DO 10, JB=2,NSM1
      V(JB)=AL(JB+1)+V(JB-1)-AL(JB)-SL(JB)-SV(JB)+F(JB)
      IF((V(JB)).LE.CHECK) V(JB)=V(JB-1)
   10 CONTINUE
      V(NS)=V(NSM1)-AL(NS)-SL(NS)-SV(NS)+F(NS)
      IF(V(NS).LE.0)V(NS)=V(NS-1)
      IF(NEGV.EQ.1) V(1)=V(2)
      WRITE(6,*)ITERAT,V(1),V(2),V(3),V(4),V(NS)
      IF(ITERAT.LE.1) GO TO 343
C
C  UPDATE WATER AND STEAM PROFILES
C
      IF(NH2O.EQ.0) GO TO 343
      DO 345 I=1,NS
      CALL KVALUE(NH2O,T(I),PT(1))
      VPW=EQUIL(NH2O)*PT(I)
      PW=VS(I)/(V(I)+VS(I))*PT(I)
      IF(PW.LE.VPW) THEN
C  STILL STEAM NO CHANGE
      IF(I.EQ.NS) GO TO 345
      VS(I+1)=VS(I)+FX(NH2O,I+1)
      ELSE
C  LIQUID WATER IS FORMED ON THIS TRAY
C
C  CHANGE PRESSURE TO ACCOUNT FOR PARTIAL PRESSURE OF H2O
```

```
C
      IF(I.EQ.NS) GO TO 347
C
      WRITE(6,346)I
  346 FORMAT(/,2X,'LIQUID WATER IS FORMED ON TRAY ',I5,/,
     $2X,'PLEASE CHECK SPEIFICATIONS OR USE A WATER DRAW TRAY.')
  347 VSF=VS(I)
      VS(I)=V(I)/(1-VPW/PT(I))-V(I)
      ALW(I)=VSF-VS(I)
      P(I)=V(I)/(VS(I)+V(I))*PT(I)
C
C   CALCULATIONS ASSUME THAT WATER IS DRAWN FROM THE
C      COLUMN IN A WATER DRAW TRAY
C
      IF(I.EQ.NS) GO TO 345
      VS(I+1)=VS(I)+FX(NH2O,I+1)
      END IF
  345 CONTINUE
C
C     IF(ITERAT.GT.1) GO TO 2463
C     IF(IPUMP.LE.0) GO TO 2463
C     DO 2464 J=1,IPUMP
C     JP=IPTRAY(J)
C     JR=IPRTRN(J)
C     SL(JP)=0.
C     F(JR)=0.
C2464 CONTINUE
C2463 CONTINUE
C
C CALCULATE ASSUMED LIQUID COMPOSITIONS
C
  343 DO 170, J=1,NS
      ERR(J)=0.0
      DO 175, I=1,NCO
      EQUIL(I)=0.0
      EQDT(I)=0.0
      Q(I,J)=0.0
  175 CONTINUE
  170 CONTINUE
      DO 17, IA=1,NC
C      IF(IA.EQ.NH2O) GO TO 17
      PCE2=0.0
      PIECE1=0.0
      PIECE3=0.0
      DENOM=0.0
      EQB=0.0
      ERB=0.0
      PCE1B=0.0
      CALL KVALUE(IA,T(1),P(1))
      PCE1B=(V(1)+SV(1))*EQUIL(IA)
      DENOM=AL(1)+SL(1)+PCE1B
      ERR(1)=1.0/DENOM
      Q(IA,1)=FX(IA,1)*ERR(1)
      EQB=EQUIL(IA)
      ERB=ERR(1)
      DO 16, JC=2,NS
      CALL KVALUE(IA,T(JC),P(JC))
      PIECE1=(V(JC)+SV(JC))*EQUIL(IA)
      PCE2=V(JC-1)*EQB*AL(JC)
      DENOM=AL(JC)+SL(JC)+PIECE1-(ERB*PCE2)
      ERR(JC)=1.0/DENOM
      Q(IA,JC)=(FX(IA,JC)+V(JC-1)*EQB*Q(IA,JC-1))*ERR(JC)
      EQB=EQUIL(IA)
      ERB=ERR(JC)
   16 CONTINUE
C
```

```
      DO 15, JD=1,NSM1
      JE=NS-JD
      PIECE3=ERR(JE)*Q(IA,JE+1)*AL(JE+1)
      Q(IA,JE)=Q(IA,JE)+PIECE3
   15 CONTINUE
      DO 1511, JE=1,NS
      IF((Q(IA,JE)).LT.CHECK) Q(IA,JE)=ABS(Q(IA,JE))
 1511 CONTINUE
   17 CONTINUE
C
C NORMALIZE THE ASSUMED COMPOSITIONS
C
      DO 1964, I=1,NC
      DO 1965, J=1,NP3
      C(I,J)=0.0
 1965 CONTINUE
 1964 CONTINUE
      DO 19, JF=1,NS
      SUMX=0.0
      DO 18, IB=1,NC
      SUMX=SUMX+Q(IB,JF)
   18 CONTINUE
      IF(SUMX.EQ.CHECK) GO TO 19
      DO 13, IC=1,NC
      C(IC,JF)=Q(IC,JF)/SUMX
   13 CONTINUE
   19 CONTINUE
C     WRITE(6,*)'COMPOSITION PROFILE LINE 551'
C     DO 3551 J=1,NS
C     WRITE(6,*)J,C(1,J),C(2,J),C(3,J)
C3551 CONTINUE
      IF(ITERAT.GT.LIM) GO TO 220
C
C
C  MAY WANT TO INCORPORATE THIS TO IMPROVE TEMP ESTIMATES
CC
C
C      L1=1
C      CALL BUBBLE(L1,T(1))
C      T1=1.15*T(1)
C      LNS=NS
C      CALL BUBBLE(LNS,T(NS))
C      TNS=T(NS)
C      ITEST=INN(3)+1
C3441 IF(ITERAT.LE.ITEST) GO TO 27
C
C CHECK FOR NON-CONVERGENCE
C
 3440 DO 21, IC=1,NC
      IF(IC.EQ.NH2O) GO TO 21
      DO 21, JCC=1,NS
C     IF(ABS(C(IC,NS)-SAVE(IC,NS)).GT.SUMERR) GO TO 27
      IF(ABS(C(IC,JCC)-SAVE(IC,JCC)).GT.SUMERR) GO TO 27
   21 CONTINUE
C     IF(SUMT.GT.0.001) GO TO 27
C     IF (SUML.GT.0.002) GO TO 27
      NEGV=1
C
C MESSAGES CONCERNING CONVERGENCE
C
  220 IF(NEGV.GT.0) GO TO 221
      WRITE (NO,3500)
 3500 FORMAT (/,2X,'THIS IS NOT A CONVERGED SOLUTION')
      GO TO 1289
  221 WRITE (NO,3600)
 3600 FORMAT (/,2X,'THIS IS A CONVERGED SOLUTION')
```

```
      WRITE(6,*)'NO OF ITERATIONS =',ITERAT
 1289 WRITE(6,1282)
 1282 FORMAT(/,2X,'DO YOU WISH TO PRINT RESULTS?')
      READ(6,1284)NOUT
 1284 FORMAT(A4)
      IF(NOUT.EQ.'N') GO TO 1201
C
C   CALCULATE REBOILER DUTY
C
C234567
      IF(REBT.EQ.0) GO TO 232
      HL=0.0
      HV=0.0
      HLU=0.0
      QR=0.0
      DO 23 IK=1,NC
      IF(IK.EQ.NH2O) GO TO 23
      CALL VLNTH(IK,T(1))
      HL=HL+C(IK,1)*HHL(IK)
      CALL VENTH(IK,T(1))
      CALL KVALUE(IK,T(1),P(1))
      HV=HV+EQUIL(IK)*C(IK,1)*HHV(IK)
      CALL VLNTH(IK,T(2))
      HLU=HLU+C(IK,2)*HHL(IK)
   23 CONTINUE
      IF(NH2O.EQ.0) GO TO 231
      CALL VENTH(NH2O,T(1))
      QR=HHV(NH2O)*VS(1)
  231 QR=QR+(V(1)+SV(1))*HV+
     $(AL(1)+SL(1))*HL-HLU*AL(2)-HF(1)
C
C    CALCULATE CONDENSER DUTY
C
C234567
      IF(JCOTYP.EQ.0) GO TO 2261
  232 HL=0.0
      HV=0.0
      HVB=0.0
      QC=0.0
      DO 24 IK=1,NC
      IF(IK.EQ.NH2O) GO TO 24
      CALL VLNTH(IK,T(NS))
      CALL VENTH(IK,T(NSM1))
      HHVB=HHV(IK)
      CALL VENTH(IK,T(NS))
      CALL KVALUE(IK,T(NSM1),P(NSM1))
      EQB=EQUIL(IK)
      CALL KVALUE(IK,T(NS),P(NS))
      HL=HL+C(IK,NS)*HHL(IK)
      HV=HV+C(IK,NS)*EQUIL(IK)*HHV(IK)
      HVB=HVB+C(IK,NSM1)*EQB*HHVB
   24 CONTINUE
      IF(JCOTYP.LE.1) GO TO 26
      HV=HL
   26 IF(NH2O.EQ.0) GO TO 261
C      WRITE(6,*)'SOUR WATER PRODUCED = ',ALW(NS)
      CALL VLNTH(NH2O,T(NS))
      CALL VENTH(NH2O,T(NSM1))
      HHWB=HHV(NH2O)
      CALL VENTH(NH2O,T(NS))
      QC=-HHWB*VS(NSM1)+HHL(NH2O)*ALW(NS)+
     $HHV(NH2O)*VS(NS)
  261 QC=QC-HVB*V(NSM1)+HL*(AL(NS)+SL(NS))+
     $HV*(V(NS)+SV(NS))-HF(NS)
C
C      OUTPUT  ROUTINE
```

```
C
C234567
      DO 920 JK=1,NF,3
 2261 WRITE(6,900)TITLE
  900 FORMAT('1',12X,A50,//,37X,'FEEDS')
      NFO=NF-JK
      IF(NFO.LT.2) GO TO 918
      WRITE(6,902)JK,JK+1,JK+2
  902 FORMAT(//,12X,'FEED #',16X,I3,13X,I3,14X,I3)
      WRITE(6,908)IFTRAY(JK),IFTRAY(JK+1),IFTRAY(JK+2)
      WRITE(6,2902)
 2902 FORMAT(/,12X,'COMP')
      J=IFTRAY(JK)
      K=IFTRAY(JK+1)
      KK=IFTRAY(JK+2)
      DO 906 I=1,NCO
      WRITE(6,904)SNAME(I),FX(I,J),FX(I,K),FX(I,KK)
  904 FORMAT(12X,A10,7X,F10.4,7X,F10.4,7X,F10.4)
      WMOLJ=WMOLJ+FX(I,J)*WMOL(I)/F(J)
      WMOLK=WOLMK+FX(I,K)*WMOL(I)/F(K)
      WMOLKK=WMOLKK+FX(I,KK)*WMOL(I)/F(KK)
      DJ=DJ+FX(I,J)*GSG(I)*62.4/F(J)
      DK=DK+FX(I,K)*GSG(I)*62.4/F(K)
      DKK=DKK+FX(I,KK)*GSG(I)*62.4/F(KK)
  906 CONTINUE
  908 FORMAT(12X,'TRAY      ',5X,I10,7X,I10,7X,I10)
  910 FORMAT(12X,'TEMP, F   ',7X,F10.4,7X,F10.4,7X,F10.4)
      IF(NH2O.EQ.0) GO TO 916
      CALL VENTH(NH2O,TFEED(J))
      F(J)=F(J)+FX(NH2O,J)
      HF(J)=HF(J)+FX(NH2O,J)*HHV(NH2O)
      F(K)=F(K)+FX(NH2O,K)
      CALL VENTH(NH2O,TFEED(K))
      HF(K)=HF(K)+FX(NH2O,K)*HHV(NH2O)
      F(KK)=F(KK)+FX(NH2O,KK)
      CALL VENTH(NH2O,TFEED(KK))
      HF(KK)=HF(KK)+FX(NH2O,KK)*HHV(NH2O)
  916 WRITE(6,912)F(J),F(K),F(KK)
      WRITE(6,910)TFEED(J),TFEED(K),TFEED(KK)
  912 FORMAT(/,12X,'RATE, LBMOLS',5X,F10.4,7X,F10.4,7X,F10.4)
      WRITE(6,914)HF(J),HF(K),HF(KK)
  914 FORMAT(12X,'ENTHALPY, BTU',1X,F13.2,4X,F13.2,4X,F13.2)
      WRITE(6,932)WLF(J),WLF(K),WLF(KK)
  932 FORMAT(12X,'L/F',14X,F10.4,7X,F10.4,7X,F10.4)
      WRITE(6,934)WMOLJ,WMOLK,WMOLKK
  934 FORMAT(12X,'MOL WT.',10X,F10.4,7X,F10.4,7X,F10.4)
      WRITE(6,936)DJ,DK,DKK
  936 FORMAT(12X,'DENSITY,LB/FT3',3X,F10.4,7X,F10.4,7X,F10.4)
      GO TO 920
  918 IF(NFO.LT.1) GO TO 917
      WRITE(6,901)JK,JK+1
  901 FORMAT(//12X,'FEED #',16X,I3,14X,I3)
      J=IFTRAY(JK)
      K=IFTRAY(JK+1)
      WRITE(6,908)IFTRAY(JK),IFTRAY(JK+1)
      WRITE(6,2902)
      DO 905 I=1,NCO
      WRITE(6,904)SNAME(I),FX(I,J),FX(I,K)
      WMOLJ=WMOLJ+FX(I,J)*WMOL(I)/F(J)
      WMOLK=WMOLK+FX(I,K)*WMOL(I)/F(K)
      DJ=DJ+FX(I,J)*GSG(I)*62.4/F(J)
      DK=DK+FX(I,K)*GSG(I)*62.4/F(K)
  905 CONTINUE
      IF(NH2O.EQ.0) GO TO 919
      CALL VENTH(NH2O,TFEED(J))
      F(J)=F(J)+FX(NH2O,J)
```

```
      HF(J)=HF(J)+FX(NH2O,J)*HHV(NH2O)
      F(K)=F(K)+FX(NH2O,K)
      CALL VENTH(NH2O,TFEED(K))
      HF(K)=HF(K)+FX(NH2O,K)*HHV(NH2O)
  919 WRITE(6,912)F(J),F(K)
      WRITE(6,910)TFEED(J),TFEED(K)
      WRITE(6,914)HF(J),HF(K)
      WRITE(6,932)WLF(J),WLF(K)
      WRITE(6,934)WMOLJ,WMOLK
      WRITE(6,936)DJ,DK
      GO TO 920
  917 WRITE(6,921)JK
  921 FORMAT(//,12X,'FEED #',16X,I3)
      J=IFTRAY(JK)
      WRITE(6,908)IFTRAY(JK)
      WRITE(6,2902)
      DO 925 I=1,NCO
      WRITE(6,904)SNAME(I),FX(I,J)
      WMOLJ=WMOLJ+FX(I,J)*WMOL(I)/F(J)
      DJ=DJ+FX(I,J)*GSG(I)*62.4/F(J)
  925 CONTINUE
      IF(NH2O.EQ.0) GO TO 915
      CALL VENTH(NH2O,TFEED(J))
      F(J)=F(J)+FX(NH2O,J)
      HF(J)=HF(J)+FX(NH2O,J)*HHV(NH2O)
  915 WRITE(6,912)F(J)
      WRITE(6,910)TFEED(J)
      WRITE(6,914)HF(J)
      WRITE(6,932)WLF(J)
      WRITE(6,934)WMOLJ
      WRITE(6,936)DJ
  920 CONTINUE
      WRITE(6,944)STEAM,TSTEAM
  944 FORMAT(//,12X,'STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS ',F10.4
     $,/,12X,'TEMPERATURE, F :',F10.4)
C
C OUTPUT PRODUCTS
C
      WRITE(6,911)TITLE
  911 FORMAT('1',//,12X,A50,//,37X,'PRODUCTS')
      WRITE(6,903)
  903 FORMAT(//,28X,'BOTTOMS',12X,'DISTILLATE')
      WRITE(6,2902)
      DO 913 I=1,NC
      BC=C(I,1)*AL(1)
      WMB=WMB+C(I,1)*WMOL(I)
      DENB=DENB+C(I,1)*GSG(I)*62.4
      IF(JCOTYP.LE.1) GO TO 907
      DC=C(I,NS)*V(NS)
      WMC=WMC+C(I,NS)*WMOL(I)
C      DENC=DENC+C(I,NS)*GSG(I)*62.4
      GO TO 909
  907 CALL KVALUE(I,T(NS),P(NS))
      DC=C(I,NS)*EQUIL(I)*V(NS)
      WMC=WMC+C(I,NS)*EQUIL(I)*WMOL(I)
C      DENC=DENC+C(I,NS)*EQUIL(I)*GSG(I)*62.4
  909 WRITE(6,904)SNAME(I),BC,DC
  913 CONTINUE
      WRITE(6,912)AL(1),V(NS)
      WRITE(6,910)T(1),T(NS)
      IF(JCOTYP.GE.2) WL=1.0
      WB=1.0
      WRITE(6,932)WB,WL
      WRITE(6,934)WMB,WMC
      WRITE(6,936)DENB
C
```

```
C  WRITE OUT DUTIES
C
C234567
      WRITE(6,931)QR,QC
  931 FORMAT(/,12X,'REBOILER DUTY, BTU : ',F14.4,/,12X
     $'CONDENSER DUTY, BTU :',F14.4)
      WRITE(6,956)ALW(NS),VS(NS)
  956 FORMAT(//,12X,'SOUR WATER PRODUCED, LBMOLS: ',F10.4,/,12X,
     $'OVERHEAD STEAM, LBMOLS:      ',F10.4)
C
C   OUTPUT SIDE DRAWS
C
      IF(NDRAW.EQ.0) GO TO 7243
      DO 9105 JD=1,NDRAW,3
      WRITE(6,9111)TITLE
 9111 FORMAT('1',//,12X,A50,//,37X,'SIDE DRAWS')
      WMJ=0.
      WMJJ=0.
      WMJK=0.
      DENJ=0.
      DENJJ=0.
      DENJK=0.
      JO=JD
      JDN=JD+2
      NOP=NDRAW-JD
      IF(JDN.GT.NDRAW) JDN=NDRAW
      WRITE(6,9103)(JO,JO=JD,JDN)
 9103 FORMAT(//,12X,'DRAW #',16X,I3,14X,I3,14X,I3)
      IF(NOP.LT.2) GO TO 9122
      WRITE(6,908)IDRAW(JD),IDRAW(JD+1),IDRAW(JD+2)
      WRITE(6,9104)
 9104 FORMAT(/,12X,'COMP #')
      J=IDRAW(JD)
      JJ=IDRAW(JD+1)
      JK=IDRAW(JD+2)
      DO 9113 I=1,NC
      BJ=C(I,J)*SL(J)
      WMJ=WMJ+C(I,J)*WMOL(I)
      DENJ=DENJ+C(I,J)*GSG(I)*62.4
      BJJ=C(I,JJ)*SL(JJ)
      WMJJ=WMJJ+C(I,JJ)*WMOL(I)
      DENJJ=DENJJ+C(I,JJ)*GSG(I)*62.4
      BJK=C(I,JK)*SL(JK)
      WMJK=WMJK+C(I,JK)*WMOL(I)
      DENJK=DENJK+C(I,JK)*GSG(I)*62.4
      WRITE(6,904)SNAME(I),BJ,BJJ,BJK
 9113 CONTINUE
      WRITE(6,912)SL(J),SL(JJ),SL(JK)
      WRITE(6,910)T(J),T(JJ),T(JK)
      WJ=1.0
      WRITE(6,932)WJ,WJ,WJ
      WRITE(6,934)WMJ,WMJJ,WMJK
      WRITE(6,936)DENJ,DENJJ,DENJK
      GO TO 9105
C
 9122 IF(NOP.LT.1) GO TO 9124
      WRITE(6,908)IDRAW(JD),IDRAW(JD+1)
      WRITE(6,9104)
      J=IDRAW(JD)
      JJ=IDRAW(JD+1)
      DO 9123 I=1,NC
      BJ=C(I,J)*SL(J)
      WMJ=WMJ+C(I,J)*WMOL(I)
      DENJ=DENJ+C(I,J)*GSG(I)*62.4
      BJJ=C(I,JJ)*SL(JJ)
      WMJJ=WMJJ+C(I,JJ)*WMOL(I)
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```
      DENJJ=DENJJ+C(I,JJ)*GSG(I)*62.4
      WRITE(6,904)SNAME(I),BJ,BJJ
 9123 CONTINUE
      WRITE(6,912)SL(J),SL(JJ)
      WRITE(6,910)T(J),T(JJ)
      WJ=1.0
      WRITE(6,932)WJ,WJ
      WRITE(6,934)WMJ,WMJJ
      WRITE(6,936)DENJ,DENJJ
      GO TO 9105
C
C
 9124 J=IDRAW(JD)
      WRITE(6,908)IDRAW(JD)
      WRITE(6,9104)
      DO 9126 I=1,NC
      BJ=C(I,J)*SL(J)
      WMJ=WMJ+C(I,J)*WMOL(I)
      DENJ=DENJ+C(I,J)*GSG(I)*62.4
      WRITE(6,904)SNAME(I),BJ
 9126 CONTINUE
      WRITE(6,912)SL(J)
      WRITE(6,910)T(J)
      WJ=1.0
      WRITE(6,932)WJ
      WRITE(6,934)WMJ
      WRITE(6,936)DENJ
 9105 CONTINUE
C
C  OUTPUT PUMPAROUNDS
C
      DO 2461 JP=1,IPUMP
      WRITE(6,2467)JP,IPTRAY(JP),IPRTRN(JP)
 2467 FORMAT(/,12X,'PUMPAROUND',I3,' FROM TRAY ',I3,' TO',I3)
      WRITE(6,2468)PSL(JP),HFP(JP)
 2468 FORMAT(12X,' RATE, LBMOLS: ',F13.4,' COOLING, BTU:',F15.4)
 2461 CONTINUE
C
C  OUTPUT SIDESTREAM STRIPPERS
C
 7243 IF(NSDS.EQ.0) GO TO 9243
      DO 7105 JD=1,NSDS,3
      WRITE(6,7111)TITLE
 7111 FORMAT('1',//,12X,A50,//,33X,'SIDESTREAM STRIPPERS')
      WMJ=0.
      WMJJ=0.
      WMJK=0.
      DENJ=0.
      DENJJ=0.
      DENJK=0.
      JO=JD
      JDN=JD+2
      NOP=NSDS-JD
      IF(JDN.GT.NSDS) JDN=NSDS
      WRITE(6,7103)(JO,JO=JD,JDN)
 7103 FORMAT(//,12X,'STRIPPER #',12X,I3,14X,I3,14X,I3)
      IF(NOP.LT.2) GO TO 7122
      WRITE(6,908)ISTRAY(JD),ISTRAY(JD+1),ISTRAY(JD+2)
      WRITE(6,7209)NSS(JD),NSS(JD+1),NSS(JD+2)
      WRITE(6,7104)
 7104 FORMAT(/,12X,'COMP #')
      J=ISTRAY(JD)
      JJ=ISTRAY(JD+1)
      JK=ISTRAY(JD+2)
      NO=(JD-1)*10
      NOP1=NO+1
```

```
      NO2=JD*10
      NOP2=NO2+1
      NO3=(JD+1)*10
      NOP3=NO3+1
      DO 7113 I=1,NC
      BJ=SC(I,NOP1)*(SAL(NOP1)-APL(JD))
      WMJ=WMJ+SC(I,NOP1)*WMOL(I)
      DENJ=DENJ+SC(I,NOP1)*GSG(I)*62.4
      BJJ=SC(I,NOP2)*(SAL(NOP2)-APL(JD+1))
      WMJJ=WMJJ+SC(I,NOP2)*WMOL(I)
      DENJJ=DENJJ+SC(I,NOP2)*GSG(I)*62.4
      BJK=SC(I,NOP3)*(SAL(NOP3)-APL(JD+2))
      WMJK=WMJK+SC(I,NOP3)*WMOL(I)
      DENJK=DENJK+SC(I,NOP3)*GSG(I)*62.4
      WRITE(6,904)SNAME(I),BJ,BJJ,BJK
 7113 CONTINUE
 7209 FORMAT(12X,'NO. TRAYS'6X,I10,7X,I10,7X,I10)
      WRITE(6,912)SAL(NOP1),SAL(NOP2),SAL(NOP3)
      WRITE(6,910)ST(NOP1),ST(NOP2),ST(NOP3)
      WJ=1.0
      WRITE(6,932)WJ,WJ,WJ
      WRITE(6,934)WMJ,WMJJ,WMJK
      WRITE(6,936)DENJ,DENJJ,DENJK
      WRITE(6,7219)SR(JD),SR(JD+1),SR(JD+2)
 7219 FORMAT(/,12X,'STEAM, LBMOLS',4X,F10.4,7X,F10.4,7X,F10.4)
      WRITE(6,7220)TS(JD),TS(JD+1),TS(JD+2)
 7220 FORMAT(12X,'TEMP, F',11X,F10.4,7X,F10.4,7X,F10.4)
      WRITE(6,7208)APL(JD),APL(JD+1),APL(JD+2)
 7208 FORMAT(/,12X,'PUMPBACK',11X,F10.4,7X,F10.4,7X,F10.4)
      WRITE(6,7207)IPSTRAY(JD),IPSTRAY(JD+1),IPSTRAY(JD+2)
 7207 FORMAT(12X,'TRAY',12X,I10,7X,I10,7X,I10)
      GO TO 7105
C
 7122 IF(NOP.LT.1) GO TO 7124
      WRITE(6,908)ISTRAY(JD),ISTRAY(JD+1)
      WRITE(6,7209)NSS(JD),NSS(JD+1)
      WRITE(6,7104)
      J=ISTRAY(JD)
      JJ=ISTRAY(JD+1)
      NO=(JD-1)*10
      NOP1=NO+1
      NO2=JD*10
      NOP2=NO2+1
      DO 7123 I=1,NC
      BJ=SC(I,NOP1)*(SAL(NOP1)-APL(JD))
      WMJ=WMJ+SC(I,NOP1)*WMOL(I)
      DENJ=DENJ+SC(I,NOP1)*GSG(I)*62.4
      BJJ=SC(I,NOP2)*(SAL(NOP2)-APL(JD+1))
      WMJJ=WMJJ+SC(I,NOP2)*WMOL(I)
      DENJJ=DENJJ+SC(I,NOP2)*GSG(I)*62.4
      WRITE(6,904)SNAME(I),BJ,BJJ
 7123 CONTINUE
      WRITE(6,912)SAL(NOP1),SAL(NOP2)
      WRITE(6,910)ST(NOP1),ST(NOP2)
      WJ=1.0
      WRITE(6,932)WJ,WJ
      WRITE(6,934)WMJ,WMJJ
      WRITE(6,936)DENJ,DENJJ
      WRITE(6,7219)SR(JD),SR(JD+1)
      WRITE(6,7220)TS(JD),TS(JD+1)
      WRITE(6,7208)APL(JD),APL(JD+1)
      WRITE(6,7207)IPSTRAY(JD),IPSTRAY(JD+1)
      GO TO 7105
 7124 J=ISTRAY(JD)
      WRITE(6,908)ISTRAY(JD)
      WRITE(6,7209)NSS(JD)
```

```
      WRITE(6,7104)
      NO=(JD-1)*10
      NOP1=NO+1
      DO 7126 I=1,NC
      BJ=SC(I,NOP1)*(SAL(NOP1)-APL(JD))
      WMJ=WMJ+SC(I,NOP1)*WMOL(I)
      DENJ=DENJ+SC(I,NOP1)*GSG(I)*62.4
      WRITE(6,904)SNAME(I),BJ
 7126 CONTINUE
      WRITE(6,910)ST(NOP1)
      WRITE(6,912)SAL(NOP1)
      WJ=1.0
      WRITE(6,932)WJ
      WRITE(6,934)WMJ
      WRITE(6,936)DENJ
      WRITE(6,7219)SR(JD)
      WRITE(6,7220)TS(JD)
      WRITE(6,7208)APL(JD)
      WRITE(6,7207)IPSTRAY(JD)
 7105 CONTINUE
C
C  WRITE OUT PROFILES
C
 9243 WRITE(6,972)TITLE
  972 FORMAT('1',//,12X,A50,//,22X,
     $'TEMPERATURE, LIQUID AND VAPOR PROFILES')
      WRITE(6,973)
  973 FORMAT(/,32X,'STREAMS LEAVING TRAY')
      WRITE(6,974)
  974 FORMAT(/,12X,'TRAY',4X,'TEMP',4X,'PRES',4X,'LIQUID',3X,
     $'MOL WT',4X,'DENS',4X,'VAPOR',2X,'MOL WT')
      WRITE(6,975)
  975 FORMAT(13X,'NO',5X,'DEGF',4X,'mmHg',4X,'LB-MOLS',11X,
     $'LB/FT3',2X,'LB-MOLS')
      DO 983 J=NS,1,-1
      SMLJ=0.
      SMVJ=0.
      DLJ=0.
      DO 984 I=1,NC
      SMLJ=SMLJ+C(I,J)*WMOL(I)
      DLJ=DLJ+C(I,J)*GSG(I)*62.4
      CALL KVALUE(I,T(J),P(J))
      SMVJ=SMVJ+C(I,J)*EQUIL(I)*WMOL(I)
c      DVJ=DVJ+C(I,J)*EQUIL(I)*GSG(I)*62.4
  984 CONTINUE
      WRITE(6,976)J,T(J),PT(J),AL(J),SMLJ,DLJ,V(J),SMVJ
  976 FORMAT(12X,I4,3F9.2,2F8.2,F9.2,2F8.2)
  983 CONTINUE
C
C WRITE OUT LIQUID COMPOSITION PROFILE
C
      DO 991 I=1,NC,7
      WRITE(6,987)TITLE
  987 FORMAT('1',//,12X,A50,//,30X,'LIQUID COMPOSITION PROFILE')
      WRITE(6,992)
  992 FORMAT(//,12X,'TRAY',20X,'COMPONENT NUMBER')
      IK=I
      II=I+6
      IF(II.GT.NC) II=NC
      WRITE(6,993)(LL,LL=IK,II)
  993 FORMAT(/,13X,'NO',5X,7(I3,5X))
      DO 994 J=NS,1,-1
      WRITE(6,995)J,(ABS(C(LL,J)),LL=IK,II)
  995 FORMAT(12X,I3,2X,7F8.4)
  994 CONTINUE
  991 CONTINUE
```

```
      WRITE(6,996)
  996 FORMAT('1')

      GO TO 1201
C
C
C
   27 DO 28, IF=1,NC
      DO 28, JIF=1,NS
      SAVE(IF,JIF)=C(IF,JIF)
   28 CONTINUE
C-------------------------------------------------------
C
C   STRIPPER CALCULATIONS
C
C      ALL STRIPPERS ARE SOLVED BY TRAY BY TRAY
C       BEFORE THE NEXT ITERATION
C
C-------------------------------------------------------
C
C   FIRST STEP UPDATE VAPOR FLOWS
C
C-------------------------------------------------------
C
      IF(ITERAT.LE.3) GO TO 699
      IF(NSDS.EQ.0) GO TO 699
      DO 563 JJ=1,NSDS
      NEGVS=0
      ITRS=0
      DO 5163 IJ=1,57
      DO 5163 IK=1,57
      C(IJ,IK)=0.
      X(IJ,IK)=0.
      DO 5163 L=1,50
      E(IJ,IK,L)=0.0
 5163 CONTINUE
      DO 5164 IK=1,57
      DO 5164 IL=1,130
      D(IK,IL)=0.0
 5164 CONTINUE
      SHF(JJ)=0.0
      JS=ISTRAY(JJ)
      ND=NSS(JJ)
      NO=(JJ-1)*10
      NDM1=ND-1
      NOP1=NO+1
      NOND=NO+ND
      DO 8511 J=1,ND
      ICNT=NO+J
      ST(ICNT)=T(JS)
 8511 CONTINUE
      SAL(NOND)=SL(JS)
      SAL(NOP1)=ALS(JJ)
      DO 6588 J=2,NDM1
      ICNT=NO+J
      SAL(ICNT)=SAL(NOND)
 6588 CONTINUE
      DO 588 IK=1,NC
      CALL VLNTH(IK,T(JS))
      SFX(IK,NOND)=SL(JS)*SAVE(IK,JS)
C     WRITE(6,*)SFX(IK,NOND),T(JS)
      SHF(JJ)=SHF(JJ)+SL(JS)*SAVE(IK,JS)*HHL(IK)
  588 CONTINUE
      WRITE(6,*)SHF(JJ),SL(JS)
  589 ITRS=ITRS+1
      SVX(NOP1)=SAL(NO+2)-SAL(NOP1)
```

```
      IF(SVX(NOP1).GT.0.0) GO TO 568
      SVX(NOP1)=0.5
      NEGVS=1
  568 DO 569, JB=2,NDM1
      ICNT=NO+JB
      SVX(ICNT)=SAL(ICNT+1)+SVX(ICNT-1)-SAL(ICNT)
      IF((SVX(ICNT)).LE.CHECK) SVX(ICNT)=SVX(ICNT-1)
  569 CONTINUE
      SVX(NOND)=SVX(NDM1)-SAL(NOND)+SL(JS)
      IF(SVX(NOND).LE.0)SVX(NOND)=SVX(NDM1)
      IF(NEGVS.EQ.1) SVX(NOP1)=SVX(NO+2)
C-----------------------------------------------------
C
C     FOR EACH STRIPPER
C CALCULATE ASSUMED LIQUID COMPOSITIONS
C
C-----------------------------------------------------
  570 DO 571, J=1,ND
      ERR(J)=0.0
      DO 571, I=1,NCO
      EQUIL(I)=0.0
      EQDT(I)=0.0
      Q(I,J)=0.0
  571 CONTINUE
      DO 572, IA=1,NC
      PCE2=0.0
      PIECE1=0.0
      PIECE3=0.0
      DENOM=0.0
      EQB=0.0
      ERB=0.0
      PCE1B=0.0
      CALL KVALUE(IA,ST(NOP1),PS(JJ))
      PCE1B=SVX(NOP1)*EQUIL(IA)
      DENOM=SAL(NOP1)+PCE1B
      ERR(1)=1.0/DENOM
      Q(IA,1)=SFX(IA,NOP1)*ERR(1)
      EQB=EQUIL(IA)
      ERB=ERR(1)
      DO 573, JC=2,ND
      IJC=JC+NO
      CALL KVALUE(IA,ST(IJC),PS(JJ))
      PIECE1=SVX(IJC)*EQUIL(IA)
      PCE2=SVX(IJC-1)*EQB*SAL(IJC)
      DENOM=SAL(IJC)+PIECE1-(ERB*PCE2)
      ERR(JC)=1.0/DENOM
      Q(IA,JC)=(SFX(IA,IJC)+SVX(IJC-1)*EQB*Q(IA,JC-1))*ERR(JC)
      EQB=EQUIL(IA)
      ERB=ERR(JC)
  573 CONTINUE
C
      DO 574, JD=1,NDM1
      JE=ND-JD
      PIECE3=ERR(JE)*Q(IA,JE+1)*SAL(JE+1+NO)
      Q(IA,JE)=Q(IA,JE)+PIECE3
  574 CONTINUE
      DO 576, JE=1,ND
      IF((Q(IA,JE)).LT.CHECK) Q(IA,JE)=ABS(Q(IA,JE))
  576 CONTINUE
  572 CONTINUE
C
C NORMALIZE THE ASSUMED COMPOSITIONS
C
      DO 577, I=1,NC
      DO 577, J=1,NP3
      C(I,J)=0.0
```

```
  577 CONTINUE
      DO 579, JF=1,ND
      SUMX=0.0
      DO 578, IB=1,NC
      SUMX=SUMX+Q(IB,JF)
  578 CONTINUE
      IF(SUMX.EQ.CHECK) GO TO 579
      DO 579, IC=1,NC
      C(IC,JF)=Q(IC,JF)/SUMX
  579 CONTINUE
      IF(ITRS.GT.25) GO TO 580
       ITEST=INN(3)+1
      IF(ITRS.LE.5) GO TO 587
C     DO 6589 IJ=1,NC
C     WRITE(6,*) C(IJ,1),C(IJ,2),C(IJ,3)
C6589 CONTINUE
C
C CHECK FOR NON-CONVERGENCE OF STRIPPER
C
      DO 581, IC=1,NC
      DO 581, JCC=1,ND
      IF(ABS(C(IC,JCC)-SC(IC,JCC+NO)).GT.0.0001) GO TO 587
  581 CONTINUE
      NEGVSS=1
C
C MESSAGES CONCERNING CONVERGENCE
C
  580 F(JS+1)=SVX(NOND)
      HF(JS+1)=0.
      IF(APL(JJ).LE.0) GO TO 8456
      JP=IPSTRAY(JJ)
      F(JP)=APL(JJ)
      HF(JP)=0.
      DO 8457 IJ=1,NC
      FX(IJ,JP)=C(IJ,1)*APL(JJ)
      CALL VLNTH(IJ,ST(NOP1))
      HF(JP)=HF(JP)+C(IJ,1)*HHL(IJ)*APL(JJ)
 8457 CONTINUE
 8456 DO 584 IA=1,NC
      CALL KVALUE(IA,ST(NOND),PS(JJ))
      CALL VENTH(IA,ST(NOND))
      FX(IA,JS+1)=C(IA,ND)*EQUIL(IA)*SVX(NOND)
C     WRITE(6,*)FX(IA,JS+1)
      HF(JS+1)=HF(JS+1)+C(IA,ND)*EQUIL(IA)*SVX(NOND)*HHV(IA)
  584 CONTINUE
      WRITE(6,*)'FEED,ENTH TO COLUMN',F(JS+1),HF(JS+1)
      IF(NEGVSS.GT.0) GO TO 586
      WRITE(6,*) 'THIS IS NOT CONVERGED STRIPPED SOLUTION'
      WRITE(6,585)
  585 FORMAT(/,2X,'THIS IS NOT A CONVERGED STRIPPER SOLUTION')
      GO TO 563
  586 WRITE(6,591)
  591 FORMAT(/,2X,'STRIPPER IS CONVERGED')
      WRITE(6,*)'NO OF ITERATIONS =',ITRS
      GO TO 563
C-------------------------------------------------
C  BEGIN STRIPPER ITERATIONS
C-------------------------------------------------
  587 DO 593 IF=1,NC
      DO 593 IJ=1,ND
      IJN=IJ+NO
      SC(IF,IJN)=C(IF,IJ)
  593 CONTINUE
C     DO 6593 IJ=1,NC
C     WRITE(6,*)SC(IJ,NOP1),SC(IJ,NO+2),SC(IJ,NOND)
C6593 CONTINUE
```

```
C
      JG=0
C
C START TRAY CALCULATIONS
C
      VB2=0.0
  601 JG=JG+1
      JGN=JG+NO
C      WRITE(6,*)'C PROFILE LINE 726'
C      DO 3552 J=1,NS
C      WRITE(6,*)J,C(1,J),C(2,J),C(3,J)
C 3552 CONTINUE
      DO 612, IH=1,NP3
      G(IH)=0.0
      DO 612, KA=1,NP3
      A(IH,KA)=0.0
      B(IH,KA)=0.0
      D(IH,KA)=0.0
      EQUIL(IH)=0.0
      EQDT(IH)=0.0
  612 CONTINUE
C
C START WITH CONSTANT MOLAL OVERFLOW
C
      G(NP1)=1.0
      DO 621, II=1,NC
      B(NP1,II)=1.0
      CALL KVALUE(II,ST(JGN),PS(JJ))
      B(II,NP1)=EQDT(II)*C(II,JG)*SVX(JGN)
      IF(JG.GE.ND) GO TO 620
      D(II,II)=-SAL(JGN+1)
      D(II,NP2)=-C(II,JG+1)
  620 IF(JG.LE.1) GO TO 622
      A(II,II)=-EQ2(II)*VB2
      A(II,NP1)=-DEQ2(II)*CB(II)*VB2
      A(II,NP3)=-EQ2(II)*CB(II)
  622 B(II,II)=SAL(JGN)+EQUIL(II)*SVX(JGN)
      B(II,NP2)=C(II,JG)
      B(II,NP3)=EQUIL(II)*C(II,JG)
      G(II)=SFX(II,JGN)
      CB(II)=0.0
      CB(II)=C(II,JG)
      EQ2(II)=0.0
      EQ2(II)=EQUIL(II)
      DEQ2(II)=0.0
      DEQ2(II)=EQDT(II)
  621 CONTINUE
      VB2=SVX(JGN)
C
C     WRITE(6,*)'A MATRIX'
C     DO 3556 JL=1,NP3
C     WRITE(5,3554)A(JL,1),A(JL,2),A(JL,3),A(JL,4),A(JL,5),A(JL,6)
C3556 CONTINUE
C     WRITE(6,*)'D MATRIX'
C     DO 3557 JK=1,NP3
C     WRITE(6,3554)D(JK,1),D(JK,2),D(JK,3),D(JK,4),D(JK,5),D(JK,6)
C3557 CONTINUE
C
C  CONSTANT MOLAL OVERFLOW ITERATIONS
      IF(ITRS.GT.INN(3)) GO TO 641
      NP3=NP3-2
C
C  CALL BAND FOR CONSTANT MOLAL ITERATIONS
C
C      WRITE (NO,3000) ITERAT,JG
C 3000 FORMAT (/,2X,'CONST. MOLAL CALL TO BAND',/,2X,'ITERAT = ',
```

```
C     1 I3,5X,'TRAY = ',I2)
      CALL BAND(JG,NP3,ND)
      NP3=NP3+2
      IF(JG.GE.ND) GO TO 665
      GO TO 601
C
C INCLUDE ITERMEDIATE TRAY EFFECTS AND ENTHALPY EFFECTS
C
C     BOTTOM TRAY HAS NO VAPOR COMING FROM BELOW
  641 WRITE(6,*) 'STRIPPER ITERATION WITH EVERYTHING',ITRS
      DO 631, I=1,NC
      EQUIL(I)=0.0
      EQDT(I)=0.0
      HHV(I)=0.0
      HHL(I)=0.0
      DHVDT(I)=0.0
  631 CONTINUE
      B(NP3,NP2)=1.0
      B(NP3,NP3)=1.0
      G(NP3)=-SAL(JGN)-SVX(JGN)
      D(NP3,NP2)=-1.0
      A(NP3,NP3)=-1.0
      IF(JG.GT.1) GO TO 649
      DO 632, IJ=1,NC
      EQB3(IJ)=0.0
      DEQ3(IJ)=0.0
      HVB3(IJ)=0.0
      HLU3(IJ)=0.0
C  THE VAPOR ENTHALPY DERIVATIVE EQUALS THE LIQUID
C     ENTHALPY DERIVATIVE
      CB3(IJ)=0.0
      CALL KVALUE(IJ,ST(NOP1),PS(JJ))
      EQB3(IJ)=EQUIL(IJ)
      DEQ3(IJ)=EQDT(IJ)
      CALL VENTH(IJ,ST(NOP1))
      HVB3(IJ)=HHV(IJ)
      CALL VENHDT(IJ,ST(NOP1))
      DHB(IJ)=DHVDT(IJ)
      CALL VLNTH(IJ,ST(NO+2))
      HLU3(IJ)=HHL(IJ)
      CB3(IJ)=C(IJ,JG)
      CALL VENHDT(IJ,ST(NO+2))
      DHJG(IJ)=DHVDT(IJ)
  632 CONTINUE
      VB3=SVX(JGN)
C
C HANDLE BOTTOMS SPECIFICATIONS
C  BOTTOM TRAY WITHOUT REBOILER
      DO 638, IJ=1,NC
      CALL VLNTH(IJ,ST(NOP1))
      CALL VENHDT(IJ,ST(NO+2))
      B(NP2,NP2)=B(NP2,NP2)+HHL(IJ)*C(IJ,JG)
      B(NP2,NP3)=B(NP2,NP3)+HVB3(IJ)*EQB3(IJ)
     1*C(IJ,JG)
      D(NP2,NP2)=D(NP2,NP2)-HLU3(IJ)*C(IJ,2)
      B(NP2,IJ)=HHL(IJ)*SAL(JGN)+HVB3(IJ)*SVX(JGN)
     $*EQB3(IJ)
      D(NP2,IJ)=-HLU3(IJ)*SAL(NO+2)
C234567
      B(NP2,NP1)=B(NP2,NP1)+C(IJ,JG)*(SAL(JGN)*
     $DHB(IJ)+SVX(JGN)*(HVB3(IJ)*
     $DEQ3(IJ)+EQB3(IJ)*DHB(IJ)))
      D(NP2,NP1)=D(NP2,NP1)-SAL(NO+2)*C(IJ,2)*DHJG(IJ)
C
C   CHANGES TO MATRIX FOR STEAM
C
```

```
     1 I3,5X,'TRAY = ',I2)
      CALL BAND(JG,NP3,ND)
      NP3=NP3+2
      IF(JG.GE.ND) GO TO 665
      GO TO 601
C
C INCLUDE ITERMEDIATE TRAY EFFECTS AND ENTHALPY EFFECTS
C
C     BOTTOM TRAY HAS NO VAPOR COMING FROM BELOW
  641 WRITE(6,*) 'STRIPPER ITERATION WITH EVERYTHING',ITRS
      DO 631, I=1,NC
      EQUIL(I)=0.0
      EQDT(I)=0.0
      HHV(I)=0.0
      HHL(I)=0.0
      DHVDT(I)=0.0
  631 CONTINUE
      B(NP3,NP2)=1.0
      B(NP3,NP3)=1.0
      G(NP3)=-SAL(JGN)-SVX(JGN)
      D(NP3,NP2)=-1.0
      A(NP3,NP3)=-1.0
      IF(JG.GT.1) GO TO 649
      DO 632, IJ=1,NC
      EQB3(IJ)=0.0
      DEQ3(IJ)=0.0
      HVB3(IJ)=0.0
      HLU3(IJ)=0.0
C  THE VAPOR ENTHALPY DERIVATIVE EQUALS THE LIQUID
C     ENTHALPY DERIVATIVE
      CB3(IJ)=0.0
      CALL KVALUE(IJ,ST(NOP1),PS(JJ))
      EQB3(IJ)=EQUIL(IJ)
      DEQ3(IJ)=EQDT(IJ)
      CALL VENTH(IJ,ST(NOP1))
      HVB3(IJ)=HHV(IJ)
      CALL VENHDT(IJ,ST(NOP1))
      DHB(IJ)=DHVDT(IJ)
      CALL VLNTH(IJ,ST(NO+2))
      HLU3(IJ)=HHL(IJ)
      CB3(IJ)=C(IJ,JG)
      CALL VENHDT(IJ,ST(NO+2))
      DHJG(IJ)=DHVDT(IJ)
  632 CONTINUE
      VB3=SVX(JGN)
C
C HANDLE BOTTOMS SPECIFICATIONS
C  BOTTOM TRAY WITHOUT REBOILER
      DO 638, IJ=1,NC
      CALL VLNTH(IJ,ST(NOP1))
      CALL VENHDT(IJ,ST(NO+2))
      B(NP2,NP2)=B(NP2,NP2)+HHL(IJ)*C(IJ,JG)
      B(NP2,NP3)=B(NP2,NP3)+HVB3(IJ)*EQB3(IJ)
     1*C(IJ,JG)
      D(NP2,NP2)=D(NP2,NP2)-HLU3(IJ)*C(IJ,2)
      B(NP2,IJ)=HHL(IJ)*SAL(JGN)+HVB3(IJ)*SVX(JGN)
     $*EQB3(IJ)
      D(NP2,IJ)=-HLU3(IJ)*SAL(NO+2)
C234567
      B(NP2,NP1)=B(NP2,NP1)+C(IJ,JG)*(SAL(JGN)*
     $DHB(IJ)+SVX(JGN)*(HVB3(IJ)*
     $DEQ3(IJ)+EQB3(IJ)*DHB(IJ)))
      D(NP2,NP1)=D(NP2,NP1)-SAL(NO+2)*C(IJ,2)*DHJG(IJ)
C
C  CHANGES TO MATRIX FOR STEAM
C
```

```
  638 CONTINUE
      CALL VENTH(NH2O,TS(JJ))
      HHBW=HHV(NH2O)
      CALL VENTH(NH2O,ST(JGN))
      G(NP2)=SR(JJ)*HHBW-VSX(JGN)*HHV(NH2O)
      VB3=SVX(JGN)
      G(NP3)=G(NP3)+SAL(JGN+1)
C
C CALL BAND FOR BOTTOM TRAY
C
C     WRITE (NO,3001) ITERAT,JG
C3001 FORMAT (/,2X,'THE SECOND CALL TO BAND',/,2X,'ITERAT = ',
C    1 I3,5X,'STAGE = ',I2)
      CALL BAND(JG,NP3,ND)
      GO TO 601
  649 IF(JG.GE.ND) GO TO 661
C
C TOP TRAY HAS NO LIQUID COMING DOWN FROM ABOVE
C INCLUDE ALL EFFECTS DUE TO COMPOSITION AND ENTHALPY
C
      G(NP3)=G(NP3)+SAL(JGN+1)+SVX(JGN-1)
      NU=JG+1
      NUN=JGN+1
      DO 651, IJ=1,NC
      CALL KVALUE(IJ,ST(JGN),PS(JJ))
      CALL VENTH(IJ,ST(JGN))
      CALL VLNTH(IJ,ST(NUN))
      CALL VENHDT(IJ,ST(NUN))
      B(NP2,NP2)=B(NP2,NP2)+HLU3(IJ)*C(IJ,JG)
      B(NP2,NP3)=B(NP2,NP3)+HHV(IJ)*EQUIL(IJ)
     1*C(IJ,JG)
      D(NP2,NP2)=D(NP2,NP2)-HHL(IJ)*C(IJ,NU)
      A(NP2,NP3)=A(NP2,NP3)-HVB3(IJ)*EQB3(IJ)*CB3(IJ)
      A(NP2,IJ)=-HVB3(IJ)*EQB3(IJ)*VB3
      B(NP2,IJ)=HLU3(IJ)*SAL(JGN)+HHV(IJ)*SVX(JGN)*EQUIL(IJ)
      D(NP2,IJ)=-HHL(IJ)*SAL(NUN)
      A(NP2,NP1)=A(NP2,NP1)-VB3*CB3(IJ)*(HVB3(IJ)*DEQ3(IJ)+
     $EQB3(IJ)*DHB(IJ))
C234567
      B(NP2,NP1)=B(NP2,NP1)+C(IJ,JG)*(SAL(JGN)*
     $DHJG(IJ)+SVX(JGN)*(HHV(IJ)*
     $EQDT(IJ)+EQUIL(IJ)*DHJG(IJ)))
      D(NP2,NP1)=D(NP2,NP1)-SAL(NUN)*C(IJ,NU)*DHVDT(IJ)
      EQB3(IJ)=0.0
      EQB3(IJ)=EQUIL(IJ)
      DEQ3(IJ)=0.0
      DEQ3(IJ)=EQDT(IJ)
      HVB3(IJ)=0.0
      HVB3(IJ)=HHV(IJ)
      HLU3(IJ)=0.0
      HLU3(IJ)=HHL(IJ)
      DHB(IJ)=0.0
      DHB(IJ)=DHJG(IJ)
      DHJG(IJ)=0.0
      DHJG(IJ)=DHVDT(IJ)
      CB3(IJ)=0.0
      CB3(IJ)=C(IJ,JG)
C
C   CHANGES TO MATRIX FOR STEAM
C
  651 CONTINUE
      CALL VENTH(NH2O,ST(JGN-1))
      HHBW=HHV(NH2O)
      CALL VENTH(NH2O,ST(JGN))
      G(NP2)=VSX(JGN-1)*HHBW-VSX(JGN)*HHV(NH2O)
      VB3=SVX(JGN)
```

```
C
C INTERMEDIATE TRAY CALL TO BAND WITH EVERYTHING
C
      CALL BAND(JG,NP3,ND)
      GO TO 601
C
C INCLUDE OVERHEAD SPECIFICATIONS
C
  661 DO 653 IJ=1,NC
      CALL KVALUE(IJ,ST(JGN),PS(JJ))
      CALL VENTH(IJ,ST(JGN))
      B(NP2,NP2)=B(NP2,NP2)+HLU3(IJ)*C(IJ,JG)
      B(NP2,NP3)=B(NP2,NP3)+HHV(IJ)*EQUIL(IJ)
     1*C(IJ,JG)
      A(NP2,NP3)=A(NP2,NP3)-HVB3(IJ)*EQB3(IJ)*CB3(IJ)
      A(NP2,IJ)=-HVB3(IJ)*EQB3(IJ)*VB3
      B(NP2,IJ)=HLU3(IJ)*SAL(JGN)+HHV(IJ)*SVX(JGN)
     $*EQUIL(IJ)
      A(NP2,NP1)=A(NP2,NP1)-VB3*CB3(IJ)*(HVB3(IJ)*DEQ3(IJ)+
     $EQB3(IJ)*DHB(IJ))
C234567
      B(NP2,NP1)=B(NP2,NP1)+C(IJ,JG)*(SAL(JGN)*
     $DHJG(IJ)+SVX(JGN)*(HHV(IJ)*
     $EQDT(IJ)+EQUIL(IJ)*DHJG(IJ)))
  653 CONTINUE
      CALL VENTH(NH2O,ST(JGN-1))
      HHBW=HHV(NH2O)
      CALL VENTH(NH2O,ST(JGN))
      G(NP2)=VSX(JGN-1)*HHBW-VSX(JGN)*HHV(NH2O)
     $+SHF(JJ)
   62 G(NP3)=G(NP3)+SVX(JGN-1)+SL(JS)
      CALL BAND(JG,NP3,ND)
C
C
C LAST CALL TO BAND AND CALCULATE NEW CHANGES IN PROFILES
C  MAX LIQUID RATE CHANGE PER TRAY IS 40%, MAX TEMP CHANGE IS 2.5F
C
      ALPRE=0.0
      SUML=0.
      SUMT=0.
      WRITE(6,*)'STRIPPER:'
      DO 6406, JH=1,ND
      JHN=JH+NO
      WRITE(6,*)'LIQUID AND TEMP CHANGES: ',C(NP2,JH),C(NP1,JH)
      MAXLIQ=ABS(C(NP2,JH))-DLLIM*SAL(JHN)
      IF(MAXLIQ.LE.0.0) GO TO 6466
      C(NP2,JH)=SAL(JHN)*DLLIM*C(NP2,JH)/ABS(C(NP2,JH))
 6466 SAL(JHN)=SAL(JHN)+C(NP2,JH)
      IF(SAL(JHN).GT.SL(JS)) SAL(JHN)=SL(JS)
 6406 CONTINUE
  665 DO 676, JI=1,ND
      JIN=JI+NO
      IF((ABS(C(NP1,JI))-DTLIM).LE.0.0) GO TO 6706
      C(NP1,JI)=DTLIM*C(NP1,JI)/ABS(C(NP1,JI))
 6706 ST(JIN)=ST(JIN)+C(NP1,JI)
  676 CONTINUE
      WRITE(6,*)'NEW PROFILES, ITERATION:',ITRS
      WRITE(6,*)'TRAY        TEMP        LIQUID'
      DO 7003 JH=1,ND
      WRITE(6,*)JH,ST(JH+NO),SAL(JH+NO)
 7003 CONTINUE
      GO TO 589
C
  563 CONTINUE
C RETURN TO CALCULATIONS
C
```

```
C-------------------------------------------------
C
C   MAIN COLUMN ITERATION
C
C-------------------------------------------------
C
      WRITE(6,*)'MAIN COLUMN ITERATION'
C     WRITE(6,*)'TEMP,LIQUID, AND VAPOR PROFILES'
C     DO 8222 J=1,NS
C     WRITE(6,*)T(J),AL(J),V(J)
C8222 CONTINUE
      DO 6997 IJ=1,57
      DO 6997 IK=1,57
      C(IJ,IK)=0.0
      X(IJ,IK)=0.0
      DO 6997 L=1,130
      D(IJ,L)=0.0
 6997 CONTINUE
      WRITE(6,*)'C, X, D CLEARED'
C
C RE-CALCULATE ASSUMED VAPOR RATES
C
      NEGV=0
      V(1)=AL(2)-AL(1)-SL(1)-SV(1)+F(1)
      IF(V(1).GT.0.0) GO TO 8811
      V(1)=1.0
      NEGV=1
 8811 DO 8810, JB=2,NSM1
      V(JB)=AL(JB+1)+V(JB-1)-AL(JB)-SL(JB)-SV(JB)+F(JB)
      IF((V(JB)).LE.CHECK) V(JB)=V(JB-1)
 8810 CONTINUE
      V(NS)=V(NSM1)-AL(NS)-SL(NS)-SV(NS)+F(NS)
      IF(V(NS).LE.0)V(NS)=V(NS-1)
      IF(NEGV.EQ.1) V(1)=V(2)
C RE-CALCULATE ASSUMED LIQUID COMPOSITIONS
C
      DO 8170, J=1,NS
      ERR(J)=0.0
      DO 8175, I=1,NCO
      EQUIL(I)=0.0
      EQDT(I)=0.0
      Q(I,J)=0.0
 8175 CONTINUE
 8170 CONTINUE
      DO 817, IA=1,NC
C      IF(IA.EQ.NH2O) GO TO 17
      PCE2=0.0
      PIECE1=0.0
      PIECE3=0.0
      DENOM=0.0
      EQB=0.0
      ERB=0.0
      PCE1B=0.0
      CALL KVALUE(IA,T(1),P(1))
      PCE1B=(V(1)+SV(1))*EQUIL(IA)
      DENOM=AL(1)+SL(1)+PCE1B
      ERR(1)=1.0/DENOM
      Q(IA,1)=FX(IA,1)*ERR(1)
      EQB=EQUIL(IA)
      ERB=ERR(1)
      DO 816, JC=2,NS
      CALL KVALUE(IA,T(JC),P(JC))
      PIECE1=(V(JC)+SV(JC))*EQUIL(IA)
      PCE2=V(JC-1)*EQB*AL(JC)
      DENOM=AL(JC)+SL(JC)+PIECE1-(ERB*PCE2)
      ERR(JC)=1.0/DENOM
```

```
      Q(IA,JC)=(FX(IA,JC)+V(JC-1)*EQB*Q(IA,JC-1))*ERR(JC)
      EQB=EQUIL(IA)
      ERB=ERR(JC)
  816 CONTINUE
C
      DO 815, JD=1,NSM1
      JE=NS-JD
      PIECE3=ERR(JE)*Q(IA,JE+1)*AL(JE+1)
      Q(IA,JE)=Q(IA,JE)+PIECE3
  815 CONTINUE
      DO 8513, JE=1,NS
      IF((Q(IA,JE)).LT.CHECK) Q(IA,JE)=ABS(Q(IA,JE))
 8513 CONTINUE
  817 CONTINUE
C
C NORMALIZE THE ASSUMED COMPOSITIONS
C
      DO 8964, I=1,NC
      DO 8965, J=1,NP3
      C(I,J)=0.0
 8965 CONTINUE
 8964 CONTINUE
      DO 819, JF=1,NS
      SUMX=0.0
      DO 818, IB=1,NC
      SUMX=SUMX+Q(IB,JF)
  818 CONTINUE
      IF(SUMX.EQ.CHECK) GO TO 819
      DO 813, IC=1,NC
      C(IC,JF)=Q(IC,JF)/SUMX
  813 CONTINUE
  819 CONTINUE
C     DO 6998 IJ=1,NC
C     DO 6998 IK=1,NS
C     C(IJ,IK)=SAVE(IJ,IK)
C6998 CONTINUE
  699 JG=0
      DO 6999 IJ=1,50
      DO 6999 IK=1,51
      DO 6999 L=1,50
      E(IJ,IK,L)=0.0
 6999 CONTINUE
      WRITE(6,*)'E ARRAY CLEARED'
C
C START TRAY CALCULATIONS
C
      VB2=0.0
   30 JG=JG+1
C     WRITE(6,*)'C PROFILE AFTER STRIP CALC'
CC    DO 3552 J=1,NS
C     WRITE(6,*)J,C(1,J),C(2,J),C(3,J)
C3552 CONTINUE
      DO 311, IH=1,NP3
      G(IH)=0.0
      DO 31, KA=1,NP3
      A(IH,KA)=0.0
      B(IH,KA)=0.0
      D(IH,KA)=0.0
      EQUIL(IH)=0.0
      EQDT(IH)=0.0
   31 CONTINUE
  311 CONTINUE
C
C START WITH CONSTANT MOLAL OVERFLOW
C
      G(NP1)=1.0
```

```
      DO 38, II=1,NC
      IF(II.EQ.NH2O) GO TO 38
      B(NP1,II)=1.0
      CALL KVALUE(II,T(JG),P(JG))
C      WRITE(6,*)'KVALUE FOR',II,'AT STAGE',JG,T(JG),'= '
C      WRITE(6,*)EQUIL(II),EQDT(II)
      B(II,NP1)=EQDT(II)*C(II,JG)*(V(JG)+SV(JG))
      IF(JG.GE.NS) GO TO 33
      D(II,II)=-AL(JG+1)
      D(II,NP2)=-C(II,JG+1)
      GO TO 35
   33 IF(JCOTYP.LE.1) GO TO 35
      EQ=1.0
      B(II,NP1)=0.0
      CALL KVALUE(II,T(JG)+SUBCOL,P(JG))
      B(NP1,II)=EQUIL(II)
      B(NP1,NP1)=B(NP1,NP1)+C(II,JG)*EQDT(II)
   35 IF(JG.LE.1) GO TO 37
      A(II,II)=-EQ2(II)*VB2
      A(II,NP1)=-DEQ2(II)*CB(II)*VB2
      A(II,NP3)=-EQ2(II)*CB(II)
   37 B(II,II)=AL(JG)+SL(JG)+EQUIL(II)*(V(JG)+SV(JG))
      B(II,NP2)=C(II,JG)
      B(II,NP3)=EQUIL(II)*C(II,JG)
      G(II)=FX(II,JG)
 6791 CB(II)=0.0
      CB(II)=C(II,JG)
      EQ2(II)=0.0
      EQ2(II)=EQUIL(II)
      DEQ2(II)=0.0
      DEQ2(II)=EQDT(II)
   38 CONTINUE
      VB2=V(JG)
C      WRITE(6,*)T(JG),AL(JG),V(JG)
C      WRITE(6,3554)B(1,1),B(1,2),B(1,3),B(1,4),B(1,5),B(1,6)
C      WRITE(6,3554)B(2,1),B(2,2),B(2,3),B(2,4),B(2,5),B(2,6)
C      WRITE(6,3554)B(3,1),B(3,2),B(3,3),B(3,4),B(3,5),B(3,6)
C      WRITE(6,3554)B(4,1),B(4,2),B(4,3),B(4,4),B(4,5),B(4,6)
C      WRITE(6,3554)B(5,1),B(5,2),B(5,3),B(5,4),B(5,5),B(5,6)
C      WRITE(6,3554)B(6,1),B(6,2),B(6,3),B(6,4),B(6,5),B(6,6)
 3554 FORMAT(1X,6F12.3)
C      WRITE(6,*)'COMPOSITION PROFILE LINE 785'
C      DO 3553 J=1,NS
C      WRITE(6,*)J,C(1,J),C(2,J),C(3,J)
C 3553 CONTINUE
C     WRITE(6,*)'A MATRIX'
C     DO 3556 JL=1,NP3
C     WRITE(5,3554)A(JL,1),A(JL,2),A(JL,3),A(JL,4),A(JL,5),A(JL,6)
C3556 CONTINUE
C     WRITE(6,*)'D MATRIX'
C     DO 3557 JK=1,NP3
C     WRITE(6,3554)D(JK,1),D(JK,2),D(JK,3),D(JK,4),D(JK,5),D(JK,6)
C3557 CONTINUE
C
C  CONSTANT MOLAL OVERFLOW ITERATIONS
      IF(ITERAT.GT.INN(3)) GO TO 40
      NP3=NP3-2
C
C  CALL BAND FOR CONSTANT MOLAL ITERATIONS
C
      WRITE (NO,3000) ITERAT,JG
 3000 FORMAT (/,2X,'CONST. MOLAL CALL TO BAND',/,2X,'ITERAT = ',
     1 I3,5X,'TRAY = ',I2)
      CALL BAND(JG,NP3,NS)
      NP3=NP3+2
      IF(JG.GE.NS) GO TO 65
```

```
      GO TO 30
C
C INCLUDE ITERMEDIATE TRAY EFFECTS AND ENTHALPY EFFECTS
C
C     BOTTOM TRAY HAS NO VAPOR COMING FROM BELOW
   40 WRITE(6,*) 'ITERATION WITH EVERYTHING',ITERAT
      DO 9150, I=1,NC
      DO 9150, J=1,NS
      EQUIL(I)=0.0
      EQDT(I)=0.0
      HHV(I)=0.0
      HHL(I)=0.0
      DHVDT(I)=0.0
 9150 CONTINUE
      B(NP3,NP2)=1.0
      B(NP3,NP3)=1.0
      G(NP3)=F(JG)-SL(JG)-AL(JG)-V(JG)-SV(JG)
  406 D(NP3,NP2)=-1.0
      A(NP3,NP3)=-1.0
      IF(JG.GT.1) GO TO 49
      DO 8951, IJ=1,NC
      EQB3(IJ)=0.0
      DEQ3(IJ)=0.0
      HVB3(IJ)=0.0
      HLU3(IJ)=0.0
C   THE VAPOR ENTHALPY DERIVATIVE EQUALS THE LIQUID
C     ENTHALPY DERIVATIVE
      CB3(IJ)=0.0
      CALL KVALUE(IJ,T(1),P(1))
      EQB3(IJ)=EQUIL(IJ)
      DEQ3(IJ)=EQDT(IJ)
      CALL VENTH(IJ,T(1))
C      WRITE(6,*)'HV ',HHV(IJ)
      HVB3(IJ)=HHV(IJ)
      CALL VENHDT(IJ,T(1))
C      WRITE(6,*)'DH ',DHVDT(IJ)
      DHB(IJ)=DHVDT(IJ)
      CALL VLNTH(IJ,T(2))
C      WRITE(6,*)'HL,2 ',HHL(IJ)
      HLU3(IJ)=HHL(IJ)
      CB3(IJ)=C(IJ,JG)
      CALL VENHDT(IJ,T(2))
C      WRITE(6,*)'DH,2 ',DHVDT(IJ)
      DHJG(IJ)=DHVDT(IJ)
 8951 CONTINUE
      VB3=V(JG)
C
C HANDLE BOTTOMS SPECIFICATIONS (FIXED BOTTOMS RATE)
C
      IF(REBT.EQ.0) GO TO 235
      B(NP2,NP2)=1.0
      G(NP3)=G(NP3)+AL(JG+1)
      GO TO 236
C
C  BOTTOM TRAY WITHOUT REBOILER
  235 DO 238, IJ=1,NC
      CALL VLNTH(IJ,T(1))
      CALL VENHDT(IJ,T(2))
      IF(IJ.EQ.NH2O) GO TO 239
      B(NP2,NP2)=B(NP2,NP2)+HHL(IJ)*C(IJ,JG)
      B(NP2,NP3)=B(NP2,NP3)+HVB3(IJ)*EQB3(IJ)
     1*C(IJ,JG)
      D(NP2,NP2)=D(NP2,NP2)-HLU3(IJ)*C(IJ,2)
      B(NP2,IJ)=HHL(IJ)*(AL(JG)+SL(JG))+HVB3(IJ)*(V(JG)+
     $SV(JG))*EQB3(IJ)
      D(NP2,IJ)=-HLU3(IJ)*AL(2)
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:234567
      B(NP2,NP1)=B(NP2,NP1)+C(IJ,JG)*((AL(JG)+SL(JG))*
     $DHB(IJ)+(V(JG)+SV(JG))*(HVB3(IJ)*
     $DEQ3(IJ)+EQB3(IJ)*DHB(IJ)))
      D(NP2,NP1)=D(NP2,NP1)-AL(2)*C(IJ,2)*DHJG(IJ)
C
C   CHANGES TO MATRIX FOR STEAM
C
  238 CONTINUE
  239 IF(NH2O.EQ.0) GO TO 1152
      CALL VENTH(NH2O,TSTEAM)
      HHBW=HHV(NH2O)
      CALL VENTH(NH2O,T(1))
      WRITE(6,*)STEAM,VS(JG),HHBW,HHV(NH2O)
      CALL VLNTH(NH2O,T(JG))
      G(NP2)=STEAM*HHBW-VS(JG)*HHV(NH2O)-ALW(JG)*HHL(NH2O)
      IF(FX(NH2O,JG).EQ.0) GO TO 1152
      CALL VENTH(NH2O,TFEED(JG))
      G(NP2)=G(NP2)+FX(NH2O,JG)*HHV(NH2O)
 1152 CONTINUE
      VB3=V(JG)
      G(NP2)=G(NP2)+HF(JG)
      WRITE(6,*) 'G(NP2) = ',G(NP2)
      G(NP3)=G(NP3)+AL(JG+1)
C
C CALL BAND FOR BOTTOM TRAY
C
C     WRITE (NO,3001) ITERAT,JG
C3001 FORMAT (/,2X,'THE SECOND CALL TO BAND',/,2X,'ITERAT = ',
C    1 I3,5X,'STAGE = ',I2)
  236 CALL BAND(JG,NP3,NS)
      GO TO 30
   49 IF(JG.GE.NS) GO TO 61
C
C TOP TRAY HAS NO LIQUID COMING DOWN FROM ABOVE
C INCLUDE ALL EFFECTS DUE TO COMPOSITION AND ENTHALPY
C
      G(NP3)=G(NP3)+AL(JG+1)+V(JG-1)
      NU=JG+1
      AHVAVG=0.
      DHLAVG=0.
      BHLAV=0.
      BHVAV=0.
      DDHAV=0.
      ADHAV=0.
      BDHAV=0.
      SUMK=0.
      SUMKB=0.
      HHBW=0.
C      DO 5511 IK=1,NC
C      WRITE(6,*)'HV,HL,DH',HHV(IK),HHL(IK),DHVDT(IK)
C      AHVAVG=AHVAVG+HVB3(IK)*CB3(IK)*EQB3(IK)
C      DHLAVG=DHLAVG+HHL(IK)*C(IK,NU)
C      WRITE(6,*)HLU3(IK),C(IK,JG)
C      BHLAV=BHLAV+HLU3(IK)*C(IK,JG)
C      BHVAV=BHVAV+HHV(IK)*EQUIL(IK)*C(IK,JG)
C      DDHAV=DDHAV+C(IK,NU)*DHVDT(IK)
C      ADHAV=ADHAV+CB3(IK)*EQB3(IK)*DHB(IK)
C      BDHAV=BDHAV+C(IK,JG)*DHJG(IK)
C      SUMK=SUMK+EQUIL(IK)*C(IK,JG)
C      SUMKB=SUMKB+EQB3(IK)*CB3(IK)
C 5511 CONTINUE
C      AHVAVG=AHVAVG/SUMKB
C      BHVAV=BHVAV/SUMK
C      ADHAV=ADHAV/SUMKB
      DO 51, IJ=1,NC
```

```
      CALL KVALUE(IJ,T(JG),P(JG))
      CALL VENTH(IJ,T(JG))
      CALL VLNTH(IJ,T(NU))
      CALL VENHDT(IJ,T(NU))
      IF(IJ.EQ.NH2O) GO TO 511
C 9505 B(NP2,NP2)=BHLAV
C      B(NP2,NP3)=BHVAV
C      D(NP2,NP2)=D(NP2,NP2)-DHLAVG*C(IJ,NU)
C      A(NP2,NP3)=A(NP2,NP3)-AHVAVG*EQB3(IJ)*CB3(IJ)
C      A(NP2,IJ)=-AHVAVG*EQB3(IJ)*VB3
C      B(NP2,IJ)=BHLAV*(AL(JG)+SL(JG))+BHVAV*(V(JG)+
C     $SV(JG))*EQUIL(IJ)
C      D(NP2,IJ)=-DHLAVG*AL(NU)
C      A(NP2,NP1)=A(NP2,NP1)-VB3*CB3(IJ)*(AHVAVG*DEQ3(IJ)+
C     $EQB3(IJ)*ADHAV)
C234567
C      B(NP2,NP1)=B(NP2,NP1)+C(IJ,JG)*((AL(JG)+SL(JG))*
C     $BDHAV+(V(JG)+SV(JG))*(BHVAV*
C     $EQDT(IJ)+EQUIL(IJ)*BDHAV))
C      D(NP2,NP1)=D(NP2,NP1)-AL(NU)*C(IJ,NU)*DDHAV
 9505 B(NP2,NP2)=B(NP2,NP2)+HLU3(IJ)*C(IJ,JG)
      B(NP2,NP3)=B(NP2,NP3)+HHV(IJ)*EQUIL(IJ)
     1*C(IJ,JG)
      D(NP2,NP2)=D(NP2,NP2)-HHL(IJ)*C(IJ,NU)
      A(NP2,NP3)=A(NP2,NP3)-HVB3(IJ)*EQB3(IJ)*CB3(IJ)
      A(NP2,IJ)=-HVB3(IJ)*EQB3(IJ)*VB3
      B(NP2,IJ)=HLU3(IJ)*(AL(JG)+SL(JG))+HHV(IJ)*(V(JG)+
     $SV(JG))*EQUIL(IJ)
      D(NP2,IJ)=-HHL(IJ)*AL(NU)
      A(NP2,NP1)=A(NP2,NP1)-VB3*CB3(IJ)*(HVB3(IJ)*DEQ3(IJ)+
     $EQB3(IJ)*DHB(IJ))
C234567
      B(NP2,NP1)=B(NP2,NP1)+C(IJ,JG)*((AL(JG)+SL(JG))*
     $DHJG(IJ)+(V(JG)+SV(JG))*(HHV(IJ)*
     $EQDT(IJ)+EQUIL(IJ)*DHJG(IJ)))
      D(NP2,NP1)=D(NP2,NP1)-AL(NU)*C(IJ,NU)*DHVDT(IJ)
C      IF(NU.EQ.NSM1) GO TO 51
      EQB3(IJ)=0.0
      EQB3(IJ)=EQUIL(IJ)
      DEQ3(IJ)=0.0
      DEQ3(IJ)=EQDT(IJ)
      HVB3(IJ)=0.0
      HVB3(IJ)=HHV(IJ)
      HLU3(IJ)=0.0
      HLU3(IJ)=HHL(IJ)
      DHB(IJ)=0.0
      DHB(IJ)=DHJG(IJ)
      DHJG(IJ)=0.0
      DHJG(IJ)=DHVDT(IJ)
      CB3(IJ)=0.0
      CB3(IJ)=C(IJ,JG)
C
C   CHANGES TO MATRIX FOR STEAM
C
   51 CONTINUE
  511 IF(NH2O.EQ.0) GO TO 1151
      CALL VENTH(NH2O,T(JG-1))
      HHBW=HHV(NH2O)
      CALL VENTH(NH2O,T(JG))
      WRITE(6,*)VS(JG-1),VS(JG),HHBW,HHV(NH2O)
      CALL VLNTH(NH2O,T(JG))
      G(NP2)=VS(JG-1)*HHBW-VS(JG)*HHV(NH2O)-ALW(JG)*HHL(NH2O)
      IF(FX(NH2O,JG).EQ.0) GO TO 1151
      G(NP2)=G(NP2)+FX(NH2O,JG)*HHV(NH2O)
C      CALL VENHDT(NH2O,T(JG-1))
C      DHB(NH2O)=DHVDT(NH2O)
```

```
C       CALL VENHDT(NH2O,T(JG))
C       A(NP2,NP1)=A(NP2,NP1)-VS(JG-1)*DHB(NH2O)
C       B(NP2,NP1)=B(NP2,NP1)+VS(JG)*DHVDT(NH2O)+DHVDT
C      $(NH2O)*ALW(JG)
 1151 CONTINUE
      VB3=V(JG)
      G(NP2)=G(NP2)+HF(JG)
      WRITE(6,*) 'G(NP2) = ',G(NP2)
C     WRITE (NO,3002) ITERAT,JG
C3002 FORMAT (/,2X,'THE THIRD CALL TO BAND',/,2X,'ITERATION = ',
C    1 I3,5X,'STAGE = ',I2)
C
C INTERMEDIATE TRAY CALL TO BAND WITH EVERYTHING
C
      CALL BAND(JG,NP3,NS)
      GO TO 30
C
C INCLUDE OVERHEAD SPECIFICATIONS
C    FIXED OVERHEAD PRODUCT RATE
C
C   52 B(NP2,NP3)=1.0
C   62 G(NP3)=G(NP3)+VB3+VS(JG-1)
C     WRITE (NO,3003) ITERAT,JG
C3003 FORMAT (/,2X,'THE FOURTH CALL TO BAND',/,2X,'ITERATION = ',
C    1 I3,5X,'STAGE = ',I2)
C      CALL BAND(JG,NP3,NS)
C
C
C
C     FIXED REFLUX RATE
C
C
   61 IF(JCOTYP.GE.1) GO TO 52
      DO 53 IJ=1,NC
      CALL KVALUE(IJ,T(JG),P(JG))
      CALL VENTH(IJ,T(JG))
      IF (IJ.EQ.NH2O) GO TO 53
      B(NP2,NP2)=B(NP2,NP2)+HLU3(IJ)*C(IJ,JG)
      B(NP2,NP3)=B(NP2,NP3)+HHV(IJ)*EQUIL(IJ)
     1*C(IJ,JG)
      A(NP2,NP3)=A(NP2,NP3)-HVB3(IJ)*EQB3(IJ)*CB3(IJ)
      A(NP2,IJ)=-HVB3(IJ)*EQB3(IJ)*VB3
      B(NP2,IJ)=HLU3(IJ)*(AL(JG)+SL(JG))+HHV(IJ)*(V(JG)+
     $SV(JG))*EQUIL(IJ)
      A(NP2,NP1)=A(NP2,NP1)-VB3*CB3(IJ)*(HVB3(IJ)*DEQ3(IJ)+
     $EQB3(IJ)*DHB(IJ))
C234567
      B(NP2,NP1)=B(NP2,NP1)+C(IJ,JG)*((AL(JG)+SL(JG))*
     $DHJG(IJ)+(V(JG)+SV(JG))*(HHV(IJ)*
     $EQDT(IJ)+EQUIL(IJ)*DHJG(IJ)))
   53 CONTINUE
      IF(NH2O.EQ.0) GO TO 1153
      CALL VENTH(NH2O,T(JG-1))
      HHBW=HHV(NH2O)
      CALL VENTH(NH2O,T(JG))
      G(NP2)=VS(JG-1)*HHBW-VS(JG)*HHV(NH2O)
      IF(FX(NH2O,JG).EQ.0) GO TO 1153
      G(NP2)=G(NP2)+FX(NH2O,JG)*HHV(NH2O)
 1153 G(NP2)=G(NP2)+HF(JG)
      GO TO 6212
C234567
   52 B(NP2,NP2)=1.0
 6212 G(NP3)=G(NP3)+V(JG-1)
      CALL BAND(JG,NP3,NS)
C
C
```

```
C LAST CALL TO BAND AND CALCULATE NEW CHANGES IN PROFILES
C  MAX LIQUID RATE CHANGE PER TRAY IS 40%, MAX TEMP CHANGE IS 2.5F
C
      ALPRE=0.0
      SUML=0.
      SUMT=0.
      DO 640, JH=1,NS
      WRITE(6,*)'LIQUID AND TEMP CHANGES: ',C(NP2,JH),C(NP1,JH)
      MAXLIQ=ABS(C(NP2,JH))-DLLIM*AL(JH)
C
C
C
      SUML=SUML+ABS(C(NP2,JH))
C
C
      IF(MAXLIQ.LE.0.0) GO TO 64
      C(NP2,JH)=AL(JH)*DLLIM*C(NP2,JH)/ABS(C(NP2,JH))
   64 AL(JH)=AL(JH)+C(NP2,JH)
C     IF(JH-2) 640,6411,6410
C6411 ALPRE=AL(JH)
C     GO TO 640
C6410 IF(AL(JH).GT.ALPRE) AL(JH)=ALPRE
C     ALPRE=AL(JH)
  640 CONTINUE
      TPRE=0.0
   65 DO 67, JI=1,NS
C     SUMT=SUMT+ABS(C(NP1,JI))
      IF((ABS(C(NP1,JI))-DTLIM).LE.0.0) GO TO 670
      C(NP1,JI)=DTLIM*C(NP1,JI)/ABS(C(NP1,JI))
  670 T(JI)=T(JI)+C(NP1,JI)
C      IF(JI.GT.1) GO TO 6710
C      TPRE=T(JI)
C      GO TO 67
C 6710 IF(T(JI).GT.TPRE) T(JI)=TPRE
C      TPRE=T(JI)
   67 CONTINUE
      WRITE(6,*)'NEW PROFILES, ITERATION:',ITERAT
      WRITE(6,*)'TRAY        TEMP        LIQUID'
      DO 7001 JH=1,NS
      WRITE(6,*)JH,T(JH),AL(JH)
 7001 CONTINUE
      GO TO 12
C
C RETURN TO CALCULATIONS
 1201 CONTINUE
C
      STOP
      END
C
C SUBROUTINE BAND
C
      SUBROUTINE BAND(J,N,NJ)
      PARAMETER (NMAX=50,NSMAX=50)
      COMMON/VAR/AA(61),BB(61),CC(61),WMW(61)
      COMMON/VAR2/AAH(61),BBH(61),CCH(61)
      COMMON/VAR3/DDH(61),EEH(61),FFH(61),DELHH(61),GSPG(61)
      COMMON/OPTION/INN(3)
      COMMON/BATCH/ID(NMAX),WMOL(NMAX)
      COMMON/KCALC/AK(NMAX),BK(NMAX),CK(NMAX),PK(NMAX),EQK(NMAX)
      COMMON/ENTH1/AH(NMAX),BH(NMAX),CH(NMAX)
      COMMON/ENTH2/DH(NMAX),EH(NMAX),FH(NMAX),DELH(NMAX)
      COMMON/GRAVIT/GSG(NMAX),BP(NMAX)
      COMMON/FEED/C(105,105),D(105,225)
      COMMON/FEED2/E(105,105,50)
      COMMON/MTRX1/X(105,105),Y(105,105)
      COMMON/MTRX2/A(105,105),B(105,105)
```

```
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
      DATA DETERM,NVAP,NP2T,NBAND,N2P1/0.0,4*0/
  101 FORMAT(/,2X,'THE DETERMINATE IS ZERO AT STAGE ',I3)
      NBAND=N+1
C      WRITE(6,*)'COMPOSITION PROF. ENTER BAND '
C      DO 3551 JH=1,NS
C      WRITE(6,*)JH,C(1,JH),C(2,JH),C(3,JH)
C 3551 CONTINUE
C
C      WRITE(6,*) 'IN BAND :'
C      WRITE(6,3554)B(1,1),B(1,2),B(1,3),B(1,4),B(1,5),B(1,6)
C      WRITE(6,3554)B(2,1),B(2,2),B(2,3),B(2,4),B(2,5),B(2,6)
C      WRITE(6,3554)B(3,1),B(3,2),B(3,3),B(3,4),B(3,5),B(3,6)
C      WRITE(6,3554)B(4,1),B(4,2),B(4,3),B(4,4),B(4,5),B(4,6)
C      WRITE(6,3554)B(5,1),B(5,2),B(5,3),B(5,4),B(5,5),B(5,6)
C      WRITE(6,3554)B(6,1),B(6,2),B(6,3),B(6,4),B(6,5),B(6,6)
C 3554 FORMAT(1X,3F14.3,3F12.3)
C      WRITE(6,*)'D MATRIX'
C234567
C      DO 3657 JI=1,6
C      WRITE(6,3554)D(JI,1),D(JI,2),D(JI,3),D(JI,4),D(JI,5),D(JI,6)
C 3657 CONTINUE
C      WRITE(6,*)'A MATRIX'
C      DO 3556 JL=1,6
C      WRITE(5,3554)A(JL,1),A(JL,2),A(JL,3),A(JL,4),A(JL,5),A(JL,6)
C 3556 CONTINUE
      N2P1=2*N+1
      IF((J-2).GT.0) GO TO 11
      IF((J-2).EQ.0) GO TO 6
C
C THE FIRST CALCULATION IS DIFFERENT BECAUSE IT LACKS AN IMAGE
C    POINT FOR CENTRAL DIFFERENCE
C EQUATION 6 IN ARTICLE:  STORE X AND G IN D
C
      DO 1, I=1,N
      D(I,N2P1)=G(I)
      DO 2, L=1,N
      LPN=L+N
      D(I,LPN)=X(I,L)
    2 CONTINUE
    1 CONTINUE
      CALL SOLVER(N,N2P1,DETERM)
      DETERM=ABS(DETERM)-0.00001
      IF (DETERM.GT.0.0) GO TO 4
      WRITE (NO,101) J
C
C  CREATE THE BEGINNINGS OF THE E ARRAY
C
    4 DO 5, K=1,N
      E(K,NBAND,1)=D(K,N2P1)
C      WRITE(6,*)'E(K,NB,1) ',E(K,NBAND,1)
      DO 51, L=1,N
      E(K,L,1)=-D(K,L)
C      WRITE(6,*)'E(K,L,1) ',E(K,L,1)
      LPN=L+N
      X(K,L)=-D(K,LPN)
C      WRITE(6,*)'X(K,L) ',X(K,L)
   51 CONTINUE
    5 CONTINUE
C      WRITE(6,*)'COMPOSITION PROF. LEAVE BAND '
C      DO 3553 JH=1,NS
C      WRITE(6,*)JH,C(1,JH),C(2,JH),C(3,JH)
C 3553 CONTINUE
```

```
      RETURN
C
C THE SECOND TRAY IS DIFFERENT BECAUSE OF THE EFFECT ON THE
C   FIRST TRAY
C
    6 DO 7, I=1,N
      DO 7, K=1,N
      DO 7, L=1,N
      D(I,K)=D(I,K)+A(I,L)*X(L,K)
C      WRITE(6,*)'D(I,K) ',D(I,K)
    7 CONTINUE
   11 CONTINUE
C       IF((J-NJ).LT.0) GO TO 260
C       DO 111, I=1,N
C       DO 111, L=1,N
C       DO 111, M=1,N
C       G(I)=G(I)-Y(I,L)*E(L,NBAND,J-2)
C       A(I,L)=A(I,L)+Y(I,M)*E(M,L,J-2)
C  111 CONTINUE
C  260 CONTINUE
      DO 12, I=1,N
      D(I,NBAND)=-G(I)
C
C EQUATION 8 IN ARTICLE
C
      DO 12, L=1,N
      D(I,NBAND)=D(I,NBAND)+A(I,L)*E(L,NBAND,J-1)
C      WRITE(6,*)'D(I,NB) ',D(I,NBAND)
      DO 12, K=1,N
C
C EQUATION 10 IN ARTICLE
C
      B(I,K)=B(I,K)+A(I,L)*E(L,K,J-1)
C      WRITE(6,*)'B(I,K) ',B(I,K)
   12 CONTINUE
      CALL SOLVER(N,NBAND,DETERM)
      DETERM=ABS(DETERM)-1.0E-5
      IF(DETERM.GT.0.0) GO TO 14
      WRITE (NO,101) J
   14 DO 15, K=1,N
      DO 15, M=1,NBAND
      E(K,M,J)=-D(K,M)
C      WRITE(6,*)'E(K,M,J) ',E(K,M,J)
   15 CONTINUE
      IF(J.LT.NJ) GO TO 20
C      WRITE(6,*) 'IN BAND J,NS',J,NJ
      DO 17, K=1,N
      C(K,J)=E(K,NBAND,J)
C      WRITE(6,*)'C(K,J) ',C(K,J)
   17 CONTINUE
      DO 18, JJ=2,NJ
      M=NJ-JJ+1
C      NP2T=N-NVAP
      DO 18, K=1,N
      C(K,M)=E(K,NBAND,M)
C      WRITE(6,*)'C(K,M) ',C(K,M)
      DO 18, L=1,N
C
C EQUATION 7 IN ARTICLE
C
      C(K,M)=C(K,M)+E(K,L,M)*C(L,M+1)
C      WRITE(6,*)'C(K,M) ',C(K,M)
   18 CONTINUE
      DO 19, LL=1,N
      DO 19, KK=1,N
C
```

```
C EQUATION 5 IN ARTICLE
C
      C(KK,1)=C(KK,1)+X(KK,LL)*C(LL,3)
C      WRITE(6,*)'C(KK,1) ',C(KK,1)
   19 CONTINUE
   20 CONTINUE
C      WRITE(6,*)'COMPOSITION PROFILE  LEAVE BAND'
C      DO 3552 JH=1,NS
C      WRITE(6,*)JH,C(1,JH),C(2,JH),C(3,JH)
C 3552 CONTINUE
      RETURN
      END
C
C     THIS SUBROUTINE (SOLVER) SOLVES THE MATRICES USING A
C        A VERY SIMPLE FORM OF PIVOTING WHEN APPLIED TO THE
C        GAUSS-JORDAN TECHNIQUE
C
      SUBROUTINE SOLVER(N,M,DETERM)
      PARAMETER (NMAX=50,NSMAX=50)
      COMMON/VAR/AA(61),BB(61),CC(61),WMW(61)
      COMMON/VAR2/AAH(61),BBH(61),CCH(61)
      COMMON/VAR3/DDH(61),EEH(61),FFH(61),DELHH(61),GSPG(61)
      COMMON/OPTION/INN(3)
      COMMON/BATCH/ID(NMAX),WMOL(NMAX)
      COMMON/KCALC/AK(NMAX),BK(NMAX),CK(NMAX),PK(NMAX),EQK(NMAX)
      COMMON/ENTH1/AH(NMAX),BH(NMAX),CH(NMAX)
      COMMON/ENTH2/DH(NMAX),EH(NMAX),FH(NMAX),DELH(NMAX)
      COMMON/GRAVIT/GSG(NMAX),BP(NMAX)
      COMMON/FEED/C(105,105),D(105,225)
      COMMON/FEED2/E(105,105,50)
      COMMON/MTRX1/X(105,105),Y(105,105)
      COMMON/MTRX2/A(105,105),B(105,105)
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
      DATA CHECK1,CHECK2/1.0E-8,0.0/
C
C LOOK FOR ZEROES IN THE MAIN DIAGONAL. IF THEY EXIST, LOOK
C    DOWN THE COLUMN FOR A NON-ZERO ENTRY AND SWAP ROWS.  THIS
C    IS REQUIRED BY THE STORAGE TECHNIQUE.
C
C      WRITE(6,*)'COMPOSITION PROF. ENTER SOLVER'
C      DO 3551 JH=1,NS
C      WRITE(6,*)JH,C(1,JH),C(2,JH),C(3,JH)
C 3551 CONTINUE
      DO 3000, J=1,N
      AMULT=0.0
      ABJJ=ABS(B(J,J))
      QUIRK=ABJJ-CHECK1
      IF (QUIRK.LT.CHECK2) GO TO 2101
      IF(ABJJ) 2101,2101,2000
 2101 IF(J.EQ.N) GO TO 4000
      JP1=J+1
      KK=0
      DO 1995, IK=JP1,N
      KK=KK+1
      QUIRK=ABS(B(IK,J))-1.0E-8
      IF(QUIRK.GT.0.0) GO TO 1990
 1995 CONTINUE
      GO TO 4000
 1990 DO 1999, I=J,N
      QUIRK=B(J,I)
      B(J,I)=B(IK,I)
      B(IK,I)=QUIRK
 1999 CONTINUE
```

```
      DO 1998, I=1,M
      QUIRK=D(IK,I)
      D(J,I)=D(IK,I)
      D(IK,I)=QUIRK
 1998 CONTINUE
 2000 IF((B(J,J)).EQ.CHECK2) GO TO 4000
      AMULT=1.0/B(J,J)
C
C THE SOLUTION OF THE MATRIX NOW FOLLOWS THE TYPICAL GAUSS-
C    JORDAN MANNER WITH THE MODIFICATION OF THIS FORM OF
C    PIVOTING.
C
      DO 2001, I=J,N
      B(J,I)=B(J,I)*AMULT
 2001 CONTINUE
      DO 2010, I=1,M
      D(J,I)=D(J,I)*AMULT
 2010 CONTINUE
      DO 2200, K=1,N
      AMLT=0.0
      AMLT=B(K,J)
      IF(K.EQ.J) GO TO 2200
      DO 2100, LIA=J,N
      B(K,LIA)=B(K,LIA)-AMLT*B(J,LIA)
 2100 CONTINUE
      DO 2200, MI=1,M
      D(K,MI)=D(K,MI)-AMLT*D(J,MI)
 2200 CONTINUE
 3000 CONTINUE
C      WRITE(6,*)'COMPOSITION PROF. LEAVE SOLVER'
C      DO 3552 JH=1,NS
C      WRITE(6,*)JH,C(1,JH),C(2,JH),C(3,JH)
C 3552 CONTINUE
      RETURN
 4000 DETERM=0.0
      RETURN
      END
C
C     SUBROUTINE CALCULATES BUBBLE POINT TEMPERATURE
C
      SUBROUTINE BUBBLE(J,TBUB)
      COMMON/BATCH/ID(100),WMOL(100)
      COMMON/ENH/AH(61),BH(61),CH(61),DH(61),EH(61),FH(61),DLTH(61)
      COMMON/KVAL/AK(100),BK(100),CK(100),PK(100),EQK(100)
      COMMON/GRAVIT/GSG(100),BP(100)
      COMMON/FEED/C(105,105),E(105,105,50),D(105,225)
      COMMON/MTRX1/X(105,105),Y(105,105)
      COMMON/MTRX2/A(105,105),B(105,105)
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
      DIMENSION YK(99),YKDT(99),PKK(99),EKK(99),EKD(99)
      INTEGER NCOND
      EOP=LOG(10.0)
      NCOND=J
      PSI=(PTOT/760.0)*14.696
   32 SUMB=0.0
      SUMBDT=0.0
      DO 320, I=1,NC
      PKK(I)=0.0
      EKK(I)=0.0
      EKD(I)=0.0
      YK(I)=0.0
      YKDT(I)=0.0
  320 CONTINUE
```

```
      TBF=(TBUB-32.0)/1.8
      DO 33, II=1,NC
      IF((C(II,NCOND)).LE.0.0) GO TO 33
      IF(ID(II).GT.NMXPRE) GO TO 3700
      DENOM=TBF+CK(II)
      IF(DENOM.LE.(0.0)) GO TO 33
      PKK(II)=AK(II)-BK(II)/DENOM
      CALC=PKK(II)-LOG10(PTOT)
      IF(CALC.GE.(-7.0)) GO TO 350
      GO TO 33
  350 IF(CALC.GE.(7.0)) GO TO 360
      EKK(II)=10.0**CALC
      GO TO 370
  360 EKK(II)=10000000.0
  370 EKD(II)=EOP*EKK(II)*BK(II)/(DENOM**2)
      GO TO 3750
C  DREISBACH
 3700 DINO=TBUB+CK(II)
      IF(DINO.LE.0.0) GO TO 33
      PK(II)=AK(II)-BK(II)/DINO
      IF((PK(II)).LE.(-20.0)) GO TO 33
      EKK(II)=EXP(PK(II))/PSI
      EKD(II)=EKK(II)*BK(II)/(DINO**2.0)
 3750 YK(II)=C(II,NCOND)*EKK(II)
      YKDT(II)=C(II,NCOND)*EKD(II)
      SUMB=SUMB+YK(II)
      SUMBDT=SUMBDT+YKDT(II)
   33 CONTINUE
      FCALC=SUMB-1.0
      IF(ABS(FCALC).LT.TOL) GO TO 34
      IF(SUMBDT.EQ.0.0) GO TO 34
      TBUB=TBUB-FCALC/SUMBDT
      GO TO 32
   34 RETURN
      END
C
C     SUBROUTINE CALCULATES DEW POINT TEMPERATURE
C
      SUBROUTINE DEWPT(J,TDEW)
      COMMON/BATCH/ID(100),WMOL(100)
      COMMON/ENH/AH(61),BH(61),CH(61),DH(61),EH(61),FH(61),DLTH(61)
      COMMON/KVAL/AK(100),BK(100),CK(100),PK(100),EQK(100)
      COMMON/GRAVIT/GSG(100),BP(100)
      COMMON/FEED/C(105,105),E(105,105,50),D(105,225)
      COMMON/MTRX/X(105,105),Y(105,105),A(105,105),B(105,105)
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
      DIMENSION XK(100),XKDT(100)
      INTEGER NREB
      EOP=LOG(10.0)
      NREB=J
      PSI=(PTOT/760.0)*14.696
   32 SUMD=0.0
      SUMDDT=0.0
      TDF=(TDEW-32.0)/1.8
      DO 33, II=1,NC
      IF((C(II,NREB)).LE.0.0) GO TO 33
      IF(ID(II).GT.NMXPRE) GO TO 2300
      DENOM1=TDF+CK(II)
      IF (DENOM1.LE.0.0) GO TO 33
      PK(II)=AK(II)-BK(II)/DENOM1
      CLC=PK(II)-LOG10(PTOT)
      IF(CLC.GE.10.0) CLC=10.0
      IF(CLC.LE.(-10.0)) CLC=-10.0
```

```
      EQKK=10.0**CLC
      EQKD=EOP*EQKK*BK(II)/(DENOM1**2.0)
      GO TO 5
C  DREISBACH
 2300 DENOM2=TDEW+CK(II)
      IF(DENOM2.LE.0.0) GO TO 33
      PK(II)=AK(II)-BK(II)/DENOM2
      CLC=PK(II)-LOG(PTOT)
      IF(CLC.GT.20.0) CLC=20.0
      IF(CLC.LE.(-20.0)) CLC=-20.0
      EQKK=EXP(CLC)/PSI
      EQKD=EQKK*BK(II)/(DENOM2**2)
    5 XK(II)=C(II,NREB)/EQKK
      XKDT(II)=C(II,NREB)*EQKD/(EQKK**2.0)
      SUMD=SUMD+XK(II)
      SUMDDT=SUMDDT+XKDT(II)
   33 CONTINUE
      FCALC=SUMD-1.0
      IF(ABS(FCALC).LT.TOL) GO TO 34
      IF(SUMDDT.EQ.0.0) GO TO 34
      TDEW=TDEW+FCALC/SUMDDT
      GO TO 32
   34 RETURN
      END
C
C     VAPOR PRESSURE SUBROUTINE - CALCULATES K-VALUES
C                                    AND K-VALUE DERIVATIVES
C       LOG VP IN MM OF HG
C       TEMPERATURE IN DEGREES CELCIUS
C               RAOULT'S LAW
C
      SUBROUTINE KVALUE(I,T,P)
      PARAMETER (NMAX=50,NSMAX=50)
      COMMON/BATCH/ID(NMAX),WMOL(NMAX)
      COMMON/KCALC/AK(NMAX),BK(NMAX),CK(NMAX),PK(NMAX),EQK(NMAX)
      COMMON/ENTH1/AH(NMAX),BH(NMAX),CH(NMAX)
      COMMON/ENTH2/DH(NMAX),EH(NMAX),FH(NMAX),DELH(NMAX)
      COMMON/GRAVIT/GSG(NMAX),BP(NMAX)
      COMMON/FEED/C(105,105),D(105,225)
      COMMON/FEED2/E(105,105,50)
      COMMON/MTRX1/X(105,105),Y(105,105)
      COMMON/MTRX2/A(105,105),B(105,105)
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
C
      REAL PSIA
      IF(ID(I).GT.NMXPRE) GO TO 10
      EPO=LOG(10.0)
      TC=(T-32.0)/1.8
      DENOM=TC+CK(I)
C     WRITE(6,*)TC,DENOM
      IF(DENOM.GE.(0.0)) GO TO 15
      EQUIL(I)=0.0
      EQDT(I)=0.0
      RETURN
   15 PK(I)=AK(I)-BK(I)/DENOM
      CALC=PK(I)-LOG10(P)
      IF(CALC.GE.(-10.0)) GO TO 25
      EQUIL(I)=1.0E-10
      GO TO 27
   25 IF(CALC.LE.(10.0)) GO TO 26
      EQUIL(I)=1.0E+10
      GO TO 27
   26 EQUIL(I)=(10**PK(I))/P
```

```
   27 EQDT(I)=EPO*EQUIL(I)*BK(I)/((TC+CK(I))**2)
C       WRITE(6,*) 'KVALUE ',EQUIL(I),EQDT(I)
      RETURN
C  DREISBACH
   10 DINO=T+CK(I)
      PSIA=P/760.*14.7
      IF(DINO.LE.0.0) GO TO 3750
      PK(I)=AK(I)-BK(I)/(T+CK(I))
      EQUIL(I)=EXP(PK(I))/PSIA
      EQDT(I)=EQUIL(I)*BK(I)/((T+CK(I))**2)
      RETURN
 3750 EQUIL(I)=0.0
      EQDT(I)=0.0
      RETURN
      END
C
C     SUBROUTINE CALCULATES VAPOR ENTHALPIES
C
      SUBROUTINE VENTH(I,T)
      PARAMETER (NMAX=50,NSMAX=50)
      COMMON/BATCH/ID(NMAX),WMOL(NMAX)
      COMMON/KCALC/AK(NMAX),BK(NMAX),CK(NMAX),PK(NMAX),EQK(NMAX)
      COMMON/ENTH1/AH(NMAX),BH(NMAX),CH(NMAX)
      COMMON/ENTH2/DH(NMAX),EH(NMAX),FH(NMAX),DELH(NMAX)
      COMMON/GRAVIT/GSG(NMAX),BP(NMAX)
      COMMON/FEED/C(105,105),D(105,225)
      COMMON/FEED2/E(105,105,50)
      COMMON/MTRX1/X(105,105),Y(105,105)
      COMMON/MTRX2/A(105,105),B(105,105)
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
C     DIMENSION DLBTU(25)
C
C  BASE OF HHV = 0.0, AT T = -200.0 FAHRENHEIT
C
C  ENTHALPY CALCULATIONS FOR THE HYPOTHETICAL FRACTIONS ARE
C     FROM JOHN ERBAR'S AND THE GPSA GPA*K COMPUTER PROGRAM
C       (AUGUST, 1974)
C
      DATA C0,C1,C2,C3,C4,C5,C6/311.30335,3.542502,-.052716
     A944,3.5353777,43.401618,-8.9530784,.35118581/
      DATA C7,C8,C9,C10,C11,C12/-.81862040,-2.8568658,
     A1.7024654,-1.4886832,.08395563,.0042934703/
      DATA C13,C14,C15,C16,C17,C18/-3.0836492,3.1073165
     1,-5.8323815,-5.6334748,9.1745019,-5.9563889/
      TABS=T+459.6
      IF(ID(I).GT.NMXPRE) GO TO 100
      HHV(I)=AH(I)+TABS*(BH(I)+TABS*(CH(I)*1.0E-4+TABS*(DH(I)
     1*1.0E-7+TABS*(EH(I)*1.0E-11+TABS*(FH(I)*1.0E-15)))))
      HHV(I)=HHV(I)*WMOL(I)
      RETURN
C  GPAK*H
  100 UOPK=((BP(I)+459.6)**(1.0/3.0))/GSG(I)
      API=141.5/GSG(I)-131.5
      HA=C0+C1*API+C2*(API**2.0)+C3*1.0E-4*(API**3.0)+C4*UOPK
     1+C5*(UOPK**2.0)+C6*(UOPK**3.0)+C18*1.0E-3*API*UOPK
      HB=C7+C8*1.0E-5*API+C9*1.0E-6*(API**2.0)+C10*1.0E-8*(API
     1**3.0)+C11*UOPK+C12*(UOPK**2.0)+C13*1.0E-4*(UOPK**3.0)
      HC=C14*1.0E-4+C15*1.0E-8*API
      HD=C16*1.0E-8+C17*1.0E-11*API
      HHV(I)=WMOL(I)*(HA+HB*T+HC*(T**2.0)+HD*
     1(T**3.0))
      RETURN
      END
```

```
C
C     SUBROUTINE CALCULATES VAPOR ENTHALPY TEMPERATURE
C          DERIVATIVES
C
      SUBROUTINE VENHDT(I,T)
      PARAMETER (NMAX=50,NSMAX=50)
      COMMON/BATCH/ID(NMAX),WMOL(NMAX)
      COMMON/KCALC/AK(NMAX),BK(NMAX),CK(NMAX),PK(NMAX),EQK(NMAX)
      COMMON/ENTH1/AH(NMAX),BH(NMAX),CH(NMAX)
      COMMON/ENTH2/DH(NMAX),EH(NMAX),FH(NMAX),DELH(NMAX)
      COMMON/GRAVIT/GSG(NMAX),BP(NMAX)
      COMMON/FEED/C(105,105),D(105,225)
      COMMON/FEED2/E(105,105,50)
      COMMON/MTRX1/X(105,105),Y(105,105)
      COMMON/MTRX2/A(105,105),B(105,105)
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
C
C  BASE OF HHV = 0.0, AT T.= -200.0 FAHRENHEIT
C
C  GPA*K
      DATA C0,C1,C2,C3,C4,C5,C6/311.30335,3.542502,-.052716
     A944,3.5353777,43.401618,-8.9530784,.35118581/
      DATA C7,C8,C9,C10,C11,C12/-.81862040,-2.8568658,
     A1.7024654,-1.4886832,.08395563,.0042934703/
      DATA C13,C14,C15,C16,C17,C18/-3.0836492,3.1073165
     1,-5.8323815,-5.6334748,9.1745019,-5.9563889/
      TABS=T+459.6
      IF(ID(I).GT.NMXPRE) GO TO 100
      DHVDT(I)=BH(I)+TABS*(2.0*CH(I)*1.0E-4+TABS*(3.0*DH(I)
     A*1.0E-7+TABS*(4.0*EH(I)*1.0E-11+TABS*(5.0*FH(I)*1.0E-15))))
      DHVDT(I)=DHVDT(I)*WMOL(I)
      RETURN
C  GPA*K
  100 UOPK=((BP(I)+459.6)**(1.0/3.0))/GSG(I)
      API=141.5/GSG(I)-131.5
      HB=C7+C8*1.0E-5*API+C9*1.0E-6*(API**2.0)+C10*1.0E-8*(API
     1**3.0)+C11*UOPK+C12*(UOPK**2.0)+C13*1.0E-4*(UOPK**3.0)
      HC=C14*1.0E-4+C15*1.0E-8*API
      HD=C16*1.0E-8+C17*1.0E-11*API
      DHVDT(I)=WMOL(I)*(HB+2*HC*T+3*HD*
     1(T**2.0))
      RETURN
      END
C
C     SUBROUTINE CALCULATES LIQUID ENTHALPIES
C
      SUBROUTINE VLNTH(I,T)
      PARAMETER (NMAX=50,NSMAX=50)
      COMMON/BATCH/ID(NMAX),WMOL(NMAX)
      COMMON/KCALC/AK(NMAX),BK(NMAX),CK(NMAX),PK(NMAX),EQK(NMAX)
      COMMON/ENTH1/AH(NMAX),BH(NMAX),CH(NMAX)
      COMMON/ENTH2/DH(NMAX),EH(NMAX),FH(NMAX),DLTH(NMAX)
      COMMON/GRAVIT/GSG(NMAX),BP(NMAX)
C
C
C
      COMMON/FEED/C(105,105),D(105,225)
      COMMON/FEED2/E(105,105,50)
      COMMON/MTRX1/X(105,105),Y(105,105)
      COMMON/MTRX2/A(105,105),B(105,105)
      COMMON/RATE/AL(50),V(50),G(105),SAVE(100,50)
      COMMON/EQB/EQUIL(100),EQDT(100),ERR(50)
      COMMON/DENTH/DHVDT(100),HHL(100),HHV(100)
```

```
      COMMON/RUN/TOL,PTOT,NS,NC,JSTAR,NMXPRE
      DIMENSION DLBTU(25)
C
C  BASE OF HHV = 0.0, AT T = -200.0 FAHRENHEIT
C
C  ENTHALPY CALCULATIONS FOR THE HYPOTHETICAL FRACTIONS ARE
C     FROM JOHN ERBAR'S AND THE GPSA GPA*K COMPUTER PROGRAM
C       (AUGUST, 1974)
C
      DATA C0,C1,C2,C3,C4,C5,C6/311.30335,3.542502,-.052716
     A944,3.5353777,43.401618,-8.9530784,.35118581/
      DATA C7,C8,C9,C10,C11,C12/-.81862040,-2.8568658,
     A1.7024654,-1.4886832,.08395563,.0042934703/
      DATA C13,C14,C15,C16,C17,C18/-3.0836492,3.1073165
     1,-5.8323815,-5.6334748,9.1745019,-5.9563889/
      TABS=T+459.6
      IF(ID(I).GT.NMXPRE) GO TO 100
      HHL(I)=AH(I)+TABS*(BH(I)+TABS*(CH(I)*1.0E-4+TABS*(DH(I)
     1*1.0E-7+TABS*(EH(I)*1.0E-11+TABS*(FH(I)*1.0E-15)))))
      HHL(I)=HHL(I)*WMOL(I)
      DLBTU(I)=DLTH(I)*1000.0*3.968321/2.20462
      HHL(I)=HHL(I)-DLBTU(I)
      RETURN
C  GPA*K
  100 UOPK=((BP(I)+459.6)**(1.0/3.0))/GSG(I)
      API=141.5/GSG(I)-131.5
      HA=C0+C1*API+C2*(API**2.0)+C3*1.0E-4*(API**3.0)+C4*UOPK
     1+C5*(UOPK**2.0)+C6*(UOPK**3.0)+C18*1.0E-3*API*UOPK
      HB=C7+C8*1.0E-5*API+C9*1.0E-6*(API**2.0)+C10*1.0E-8*(API
     1**3.0)+C11*UOPK+C12*(UOPK**2.0)+C13*1.0E-4*(UOPK**3.0)
      HC=C14*1.0E-4+C15*1.0E-8*API
      HD=C16*1.0E-8+C17*1.0E-11*API
      HHL(I)=WMOL(I)*(HA+HB*T+HC*(T**2.0)+HD*
     1(T**3.0))
C  LATENT HEAT OF VAPORIZATION
C       TC=768.07121+1.7133693*BP(I)-.0010834003*BP(I)**2-.0089212579
C     $**BP(I)**3+3.8890584E-07*API*BP(I)+5.3094920E-06*API*BP(I)
C     $**2+3.2711600E-08*API*BP(I)**3
      BPR=BP(I)+459.6
      DLTH(I)=BPR*(7.58+4.571*ALOG10(BPR))
C       DLTH(I)=DLTH(I)*((TC-537.)/(TC-BPR))**.38
C       DLBTU(I)=DLTH(I)*1000.0*3.968321/2.20462
      HHL(I)=HHL(I)-DLTH(I)
      RETURN
      END
C
```

APPENDIX B

CASE STUDY RESULTS

CASE STUDY I

HYPOTHETICAL TWO-PHASE DISTILLATION

PROGRAM OUTPUT

FEEDS

FEED #	1
TRAY	7
COMP	
C3H8	1.0000
N-C4H10	2.0000
N-C5H12	3.0000
N-C7H16	5.0000
N-C10H22	22.0000
N-C13H28	30.0000
N-C15H32	40.0000
RATE, LBMOLS	103.0000
TEMP, F	300.0000
ENTHALPY, BTU	4674859.50
L/F	0.9382
MOL WT.	175.1080
DENSITY,LB/FT3	46.4505

STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS 0.0000
TEMPERATURE, F : 0.0000

TEST1

PRODUCTS

	BOTTOMS	DISTILLATE
COMP		
C3H8	0.0000	1.0007
N-C4H10	0.0000	2.0014
N-C5H12	0.0016	3.0006
N-C7H16	3.0001	2.0012
N-C10H22	21.9996	0.0000
N-C13H28	29.9994	0.0000
N-C15H32	39.9993	0.0000
RATE, LBMOLS	95.0000	8.0000
TEMP, F	367.5810	127.1180
L/F	1.0000	0.0000
MOL WT.	183.7760	72.1862
DENSITY,LB/FT3	47.1134	

REBOILER DUTY, BTU : 708743.0000
CONDENSER DUTY, BTU : -152742.5781

SOUR WATER PRODUCED, LBMOLS: 0.0000
OVERHEAD STEAM, LBMOLS: 0.0000

TEST1

TEMPERATURE, LIQUID AND VAPOR PROFILES

STREAMS LEAVING TRAY

TRAY NO	TEMP DEGF	PRES mmHg	LIQUID LB-MOLS	MOL WT	DENS LB/FT3	VAPOR LB-MOLS	MOL WT
13	127.12	517.00	10.00	94.39	42.15	8.00	72.19
12	159.67	517.00	10.47	98.00	42.65	18.00	84.53
11	164.86	517.00	10.54	98.57	42.72	18.47	86.82
10	166.12	517.00	10.27	99.08	42.76	18.54	87.19
9	171.48	517.00	8.40	103.70	43.08	18.27	87.27
8	208.08	517.00	4.78	130.47	44.71	16.40	88.22
7	300.55	517.00	100.50	179.43	46.86	12.78	93.99
6	306.85	517.00	104.24	177.30	46.78	5.50	104.34
5	311.15	517.00	106.83	176.01	46.73	9.24	110.81
4	313.54	517.00	108.11	175.52	46.72	11.83	113.69
3	317.10	517.00	109.39	175.49	46.73	13.11	115.75
2	329.11	517.00	112.71	176.45	46.80	14.39	120.89
1	367.58	517.00	95.00	183.78	47.11	17.71	137.22

TEST1

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
13	0.0048	0.0325	0.1489	0.8138	0.0000	0.0000	0.0000
12	0.0016	0.0109	0.0593	0.9282	0.0001	0.0000	0.0000
11	0.0014	0.0090	0.0431	0.9455	0.0008	0.0000	0.0000
10	0.0014	0.0088	0.0403	0.9390	0.0106	0.0000	0.0000
9	0.0013	0.0083	0.0374	0.8374	0.1132	0.0022	0.0002
8	0.0011	0.0061	0.0248	0.3811	0.4558	0.0936	0.0375
7	0.0007	0.0033	0.0110	0.0563	0.2292	0.3006	0.3989
6	0.0001	0.0012	0.0085	0.0846	0.2290	0.2914	0.3852
5	0.0000	0.0003	0.0039	0.1045	0.2294	0.2856	0.3764
4	0.0000	0.0000	0.0014	0.1114	0.2319	0.2830	0.3723
3	0.0000	0.0000	0.0004	0.1068	0.2419	0.2819	0.3689
2	0.0000	0.0000	0.0001	0.0816	0.2675	0.2862	0.3645
1	0.0000	0.0000	0.0000	0.0316	0.2316	0.3158	0.4210

CASE STUDY I (Continued)

HYPOTHETICAL TWO-PHASE DISTILLATION

MAXI*SIM OUTPUT

SCI MAXI*SIM JOB ID:
VERSION #2.2 DATE: 19-AUG-1988
PAGE 3 USER:

HYPOTHETICAL 2 PHASE COLUMN

UNIT OPERATION NO 3 IS A DTXT UNIT***

FEEDS>>>>>>>PRODUCTS>>>>>>>>>>>>>>>>

STREAM FLOW RATES ARE LB-MOLS

STREAM NO / NAME / COMPONENT	1	2	3
C3H8	1.0000	1.0000	0.0000
NC4H10	2.0000	2.0000	0.0000
NC5H12	3.0000	2.9984	0.0016
NC7H16	5.0000	2.0047	2.9953
NC10H22	22.0000	0.0000	22.0000
NC13H28	30.0000	0.0000	30.0000
NC15H32	40.0000	0.0000	40.0000
TOTAL	103.0000	8.0031	94.9969
T,DEG F	300.00	126.91	373.30
P,PSIA	10.00	10.00	10.00
H,KBTU	1968.28	92.29	2660.74
S,KBTU/R	16.2519	0.7166	16.5016
MOL WEIGHT	175.1042	72.1622	183.7768
D,LB/FT3	3.2569	0.1172	39.1975
L/F(MOLAR)	0.93819	0.00001	1.00000

COLUMN CONDENSER,REBOILER & SIDE HEATER/COOLER DUTIES

CONDENSER	-163.128 KBTU
REBOILER	947.920 KBTU

ESTIMATED COLUMN DIAMETERS

TOP SECTION	0.96 FT
BOTTOM SECTION	1.92 FT

(BASED ON 75% VAPOR FLOOD VELOCITY &)
(5 SEC LIQ RESID IN DOWNCOMER)

SCI MAXI*SIM JOB ID:
VERSION #2.2 DATE: 19-AUG-1988
PAGE 4 USER:

HYPOTHETICAL 2 PHASE COLUMN

TEMPERATURE,PRESSURE,LIQUID AND VAPOR PROFILES

TRAY			STREAMS LEAVING TRAY					
			LIQUID			VAPOR		
NO	TEMP DEG F	PRES PSIA	LB-MOL	MOL WT	DENSITY LB/FT3	LB-MOL	MOL WT	DENSITY LB/FT3
12	126.91	10.0	10.0	94.1	40.149	8.0	72.2	0.117
11	159.17	10.0	10.6	97.8	39.505	18.0	84.3	0.130
10	164.63	10.0	10.7	98.4	39.395	18.6	86.7	0.133
9	166.01	10.0	10.4	98.9	39.415	18.7	87.1	0.133
8	171.51	10.0	8.5	103.5	39.797	18.4	87.3	0.132
7	208.13	10.0	4.8	129.9	41.069	16.5	88.3	0.126
6	301.88	10.0	104.8	176.6	40.940	6.4	99.4	0.124
5	306.86	10.0	108.4	174.8	40.731	9.8	107.4	0.133
4	309.65	10.0	110.4	174.0	40.619	13.4	111.5	0.138
3	311.76	10.0	111.5	173.7	40.550	15.4	113.4	0.140
2	316.51	10.0	113.3	173.8	40.423	16.5	115.7	0.142
1	331.75	10.0	117.8	175.3	40.052	18.3	122.1	0.148
0	373.30	10.0	95.0	183.8	39.197	22.8	139.9	0.161

SCI MAXI*SIM JOB ID:
VERSION #2.2 DATE: 19-AUG-1988
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HYPOTHETICAL 2 PHASE COLUMN

LIQUID COMPOSITION PROFILE

TRAY NO	COMPONENT NUMBER 1	2	3	4	5	6	7
12	0.00587	0.03553	0.15406	0.80453	0.00000	0.00000	0.00000
11	0.00206	0.01259	0.06392	0.92135	0.00007	0.00000	0.00000
10	0.00184	0.01041	0.04652	0.94034	0.00089	0.00000	0.00000
9	0.00181	0.01010	0.04321	0.93405	0.01082	0.00002	0.00000
8	0.00175	0.00962	0.04018	0.83328	0.11275	0.00223	0.00018
7	0.00148	0.00738	0.02763	0.38640	0.44594	0.09423	0.03694
6	0.00041	0.00289	0.01324	0.08334	0.22740	0.28972	0.38300
5	0.00006	0.00086	0.00759	0.11198	0.22717	0.28152	0.37082
4	0.00001	0.00019	0.00322	0.12659	0.22805	0.27739	0.36455
3	0.00000	0.00004	0.00119	0.12959	0.23233	0.27557	0.36128
2	0.00000	0.00001	0.00040	0.11984	0.24803	0.27467	0.35706
1	0.00000	0.00000	0.00011	0.08570	0.27955	0.28194	0.35270
0	0.00000	0.00000	0.00002	0.03153	0.23159	0.31580	0.42107

SCI MAXI*SIM JOB ID:
VERSION #2.2 DATE: 19-AUG-1988
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HYPOTHETICAL 2 PHASE COLUMN

VAPOR COMPOSITION PROFILE

TRAY NO	COMPONENT NUMBER 1	2	3	4	5	6	7
12	0.12495	0.24990	0.37466	0.25049	0.00000	0.00000	0.00000
11	0.05881	0.13082	0.25212	0.55825	0.00000	0.00000	0.00000
10	0.05502	0.11486	0.19783	0.63225	0.00004	0.00000	0.00000
9	0.05463	0.11311	0.18723	0.64452	0.00051	0.00000	0.00000
8	0.05543	0.11450	0.18751	0.63644	0.00611	0.00001	0.00000
7	0.06163	0.12638	0.20269	0.55011	0.05795	0.00115	0.00009
6	0.02896	0.10090	0.23706	0.41190	0.17277	0.03481	0.01360
5	0.00439	0.03086	0.14112	0.58433	0.18688	0.03756	0.01487
4	0.00049	0.00694	0.06116	0.68064	0.19595	0.03919	0.01562
3	0.00005	0.00136	0.02299	0.71248	0.20622	0.04063	0.01626
2	0.00000	0.00025	0.00795	0.69273	0.23663	0.04451	0.01793
1	0.00000	0.00004	0.00238	0.57823	0.33338	0.06117	0.02481
0	0.00000	0.00000	0.00048	0.31104	0.47909	0.14108	0.06831

SCI MAXI*SIM JOB ID:
VERSION #2.2 DATE: 19-AUG-1988
PAGE 7 USER:

HYPOTHETICAL 2 PHASE COLUMN

K VALUE PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
12	21.2695	7.03421	2.43184	0.31134	0.01536	0.00077	0.00011
11	28.5333	10.3877	3.94404	0.60591	0.03913	0.00258	0.00043
10	29.8538	11.0349	4.25236	0.67237	0.04531	0.00312	0.00053
9	30.1951	11.2031	4.33305	0.69003	0.04699	0.00327	0.00056
8	31.5973	11.8946	4.66610	0.76383	0.05419	0.00393	0.00069
7	41.5976	17.1286	7.33433	1.42383	0.13002	0.01217	0.00254
6	70.3637	34.8858	17.9012	4.94240	0.75991	0.12020	0.03553
5	71.7531	35.8973	18.5821	5.21830	0.82266	0.13343	0.04010
4	72.5287	36.4659	18.9674	5.37648	0.85933	0.14132	0.04286
3	73.1399	36.9066	19.2642	5.49800	0.88764	0.14746	0.04502
2	74.5788	37.9323	19.9526	5.78033	0.95405	0.16204	0.05020
1	79.1794	41.2748	22.2352	6.74734	1.19252	0.21693	0.07033
0	91.2972	50.6163	28.9462	9.86459	2.06873	0.44672	0.16223

CASE STUDY II

H_2S STRIPPER

PROGRAM OUTPUT

FEEDS

FEED #	1	2
TRAY	2	1
COMP		
H2S	5.1700	1.8200
N2	0.0000	7.0700
CO2	1.2300	30.2300
CH4	1.3800	274.9100
C2H6	15.4200	90.6500
C3H8	114.6000	59.4000
I-C4H10	58.6400	7.4600
N-C4H10	211.2500	18.1700
I-C5H12	157.2100	4.6200
N-C5H12	358.8500	8.2300
3N-C17H36	5006.5000	4.9800
RATE, LBMOLS	5930.2500	507.5400
TEMP, F	130.0000	130.0000
ENTHALPY, BTU	221364144.00	4694450.50
L/F	1.0000	0.0000
MOL WT.	212.8582	29.4743
DENSITY,LB/FT3	46.6167	25.0469

STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS 0.0000
TEMPERATURE, F : 0.0000

TEST 2

PRODUCTS

	BOTTOMS	DISTILLATE
COMP		
H2S	1.6523	5.3380
N2	0.2215	6.8488
CO2	0.0036	31.4577
CH4	17.3098	258.9911
C2H6	28.2460	77.8275
C3H8	85.1854	88.8192
I-C4H10	46.2360	19.8652
N-C4H10	175.4587	53.9654
I-C5H12	144.4132	17.4190
N-C5H12	335.9069	31.1779
3N-C17H36	5011.5293	0.0023
RATE, LBMOLS	5846.1631	591.6270
TEMP, F	128.3595	129.6415
L/F	1.0000	0.0000
MOL WT.	215.0655	33.7501
DENSITY,LB/FT3	46.6937	

REBOILER DUTY, BTU : 0.0000
CONDENSER DUTY, BTU : 11856.0000

SOUR WATER PRODUCED, LBMOLS: 0.0000
OVERHEAD STEAM, LBMOLS: 0.0000

TEST 2

TEMPERATURE, LIQUID AND VAPOR PROFILES

STREAMS LEAVING TRAY

TRAY NO	TEMP DEGF	PRES mmHg	LIQUID LB-MOLS	MOL WT	DENS LB/FT3	VAPOR LB-MOLS	MOL WT
2	129.64	1551.00	5920.96	212.99	46.57	591.63	33.75
1	128.36	1551.00	5846.16	215.07	46.69	582.33	32.17

TEST 2

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
2	0.0005	0.0000	0.0000	0.0027	0.0042	0.0169	0.0093
1	0.0003	0.0000	0.0000	0.0030	0.0048	0.0146	0.0079

TEST 2

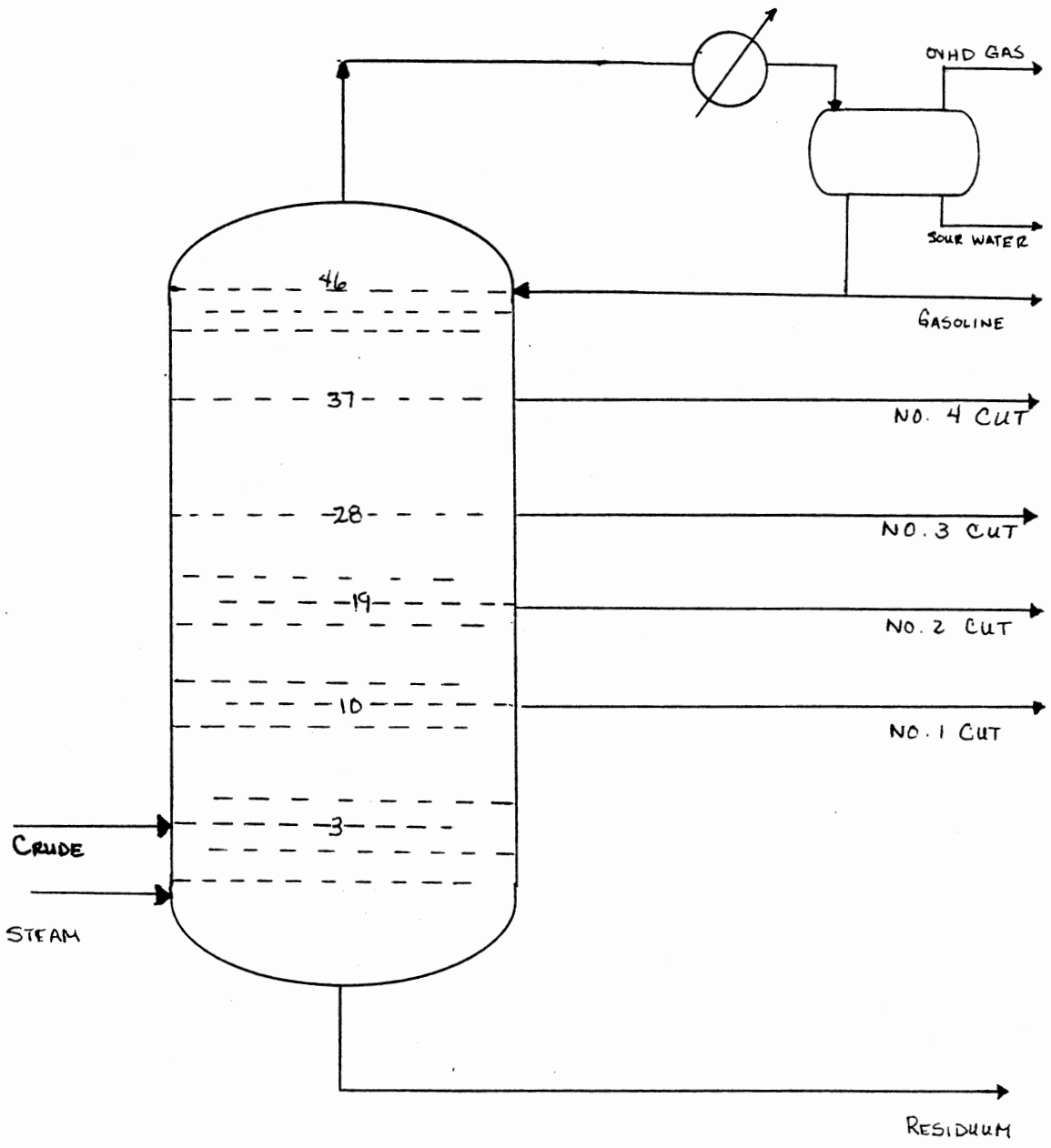
LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER			
NO	8	9	10	11
2	0.0343	0.0263	0.0602	0.8456
1	0.0300	0.0247	0.0575	0.8572

CASE STUDY III

STRIPPED-DOWN CRUDE COLUMN

PROGRAM OUTPUT



STRIPPED DOWN CRUDE COLUMN

FEEDS

FEED #	1
TRAY	3
COMP	
C2H6	0.0303
C3H8	0.4670
I-C4H10	0.1980
N-C4H10	0.8620
I-C5H12	0.6810
N-C5H12	0.9240
N-C6H14	1.6700
FR1	0.7220
FR2	2.7200
FR3	2.5220
FR4	2.8620
FR5	2.4210
FR6	2.2000
FR7	1.8980
FR8	1.6990
FR9	1.3570
FR10	0.8510
FR11	0.7570
FR12	0.6780
FR13	0.6580
FR14	0.5860
H2O	0.0000
RATE, LBMOLS	26.7633
TEMP, F	584.9500
ENTHALPY, BTU	926972.06
L/F	0.2880
MOL WT.	204.2151
DENSITY,LB/FT3	49.5434

STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS 5.8100
TEMPERATURE, F : 700.0000

IRANIAN CRUDE

PRODUCTS

	BOTTOMS	DISTILLATE
COMP		
C2H6	0.0000	0.0280
C3H8	0.0001	0.3602
I-C4H10	0.0001	0.1139
N-C4H10	0.0006	0.4294
I-C5H12	0.0013	0.1975
N-C5H12	0.0019	0.2221
N-C6H14	0.0084	0.1561
FR1	0.0068	0.0501
FR2	0.0393	0.0481
FR3	0.0701	0.0000
FR4	0.1661	0.0000
FR5	0.2763	0.0000
FR6	0.6304	0.0000
FR7	1.7811	0.0000
FR8	1.6989	0.0000
FR9	1.3570	0.0000
FR10	0.8510	0.0000
FR11	0.7570	0.0000
FR12	0.6780	0.0000
FR13	0.6580	0.0000
FR14	0.5860	0.0000
RATE, LBMOLS	9.5684	1.6060
TEMP, F	649.4727	123.6684
L/F	1.0000	0.0000
MOL WT.	357.5368	63.2228
DENSITY,LB/FT3	56.5475	

REBOILER DUTY, BTU : 0.0000
CONDENSER DUTY, BTU : -526626.5000

SOUR WATER PRODUCED, LBMOLS: 5.7030
OVERHEAD STEAM, LBMOLS: 0.1070

IRANIAN CRUDE

SIDE DRAWS

DRAW #	1	2	3
TRAY	46	37	10
COMP #			
C2H6	0.0022	0.0000	0.0000
C3H8	0.1036	0.0016	0.0007
I-C4H10	0.0815	0.0014	0.0005
N-C4H10	0.4188	0.0076	0.0026
I-C5H12	0.4625	0.0121	0.0034
N-C5H12	0.6690	0.0194	0.0051
N-C6H14	1.3787	0.0881	0.0163
FR1	0.5786	0.0626	0.0101
FR2	1.2075	1.2855	0.0557
FR3	0.0005	2.1620	0.0998
FR4	0.0000	0.6945	0.2788
FR5	0.0000	0.0000	1.3294
FR6	0.0000	0.0000	1.5695
FR7	0.0000	0.0000	0.1169
FR8	0.0000	0.0000	0.0001
FR9	0.0000	0.0000	0.0000
FR10	0.0000	0.0000	0.0000
FR11	0.0000	0.0000	0.0000
FR12	0.0000	0.0000	0.0000
FR13	0.0000	0.0000	0.0000
FR14	0.0000	0.0000	0.0000
RATE, LBMOLS	4.9030	4.3350	3.4890
TEMP, F	123.6684	286.2103	460.6146
L/F	1.0000	1.0000	1.0000
MOL WT.	84.0073	114.8376	177.2745
DENSITY,LB/FT3	41.7048	47.0323	50.8577

IRANIAN CRUDE

SIDE DRAWS

DRAW #	4	5
TRAY	19	28
COMP #		
C2H6	0.0000	0.0000
C3H8	0.0003	0.0004
I-C4H10	0.0003	0.0003
N-C4H10	0.0013	0.0016
I-C5H12	0.0018	0.0023
N-C5H12	0.0028	0.0036
N-C6H14	0.0095	0.0127
FR1	0.0059	0.0079
FR2	0.0347	0.0490
FR3	0.0726	0.1169
FR4	0.4868	1.2357
FR5	0.8147	0.0006
FR6	0.0001	0.0000
FR7	0.0000	0.0000
FR8	0.0000	0.0000
FR9	0.0000	0.0000
FR10	0.0000	0.0000
FR11	0.0000	0.0000
FR12	0.0000	0.0000
FR13	0.0000	0.0000
FR14	0.0000	0.0000
RATE, LBMOLS	1.4310	1.4310
TEMP, F	404.5782	364.5188
L/F	1.0000	1.0000
MOL WT.	154.0909	135.6471
DENSITY,LB/FT3	49.3448	48.4826

IRANIAN CRUDE

TEMPERATURE, LIQUID AND VAPOR PROFILES

STREAMS LEAVING TRAY

TRAY NO	TEMP DEGF	PRES mmHg	LIQUID LB-MOLS	MOL WT	DENS LB/FT3	VAPOR LB-MOLS	MOL WT
46	123.67	1292.50	27.30	84.01	41.70	1.61	63.22
45	197.12	1298.24	27.01	94.68	44.07	33.81	83.02
44	222.60	1303.99	26.29	98.90	44.98	33.52	91.61
43	232.75	1309.73	25.87	100.82	45.36	32.80	94.93
42	237.79	1315.48	25.63	101.85	45.55	32.38	96.42
41	241.34	1321.22	25.47	102.64	45.68	32.14	97.20
40	245.62	1326.97	25.26	103.69	45.82	31.98	97.80
39	252.93	1332.71	24.89	105.58	46.07	31.77	98.61
38	265.82	1338.46	24.21	109.02	46.47	31.40	100.04
37	286.21	1344.20	18.80	114.84	47.03	30.72	102.63
36	313.49	1349.94	17.70	122.81	47.66	29.65	106.94
35	336.31	1355.69	16.86	129.07	48.08	28.54	111.58
34	350.50	1361.43	16.39	132.67	48.31	27.71	115.06
33	357.71	1367.18	16.17	134.39	48.41	27.24	116.98
32	361.09	1372.92	16.07	135.13	48.45	27.01	117.88
31	362.70	1378.67	16.03	135.44	48.47	26.92	118.26
30	363.56	1384.41	16.01	135.57	48.48	26.87	118.42
29	364.10	1390.16	16.00	135.62	48.48	26.85	118.49
28	364.52	1395.90	14.56	135.65	48.48	26.85	118.51
27	364.91	1401.64	14.56	135.67	48.48	26.84	118.52
26	365.33	1407.39	14.56	135.73	48.49	26.84	118.53
25	365.90	1413.13	14.55	135.88	48.49	26.83	118.56
24	366.83	1418.88	14.53	136.26	48.51	26.82	118.64
23	368.68	1424.62	14.48	137.17	48.55	26.80	118.83
22	372.50	1430.37	14.37	139.20	48.65	26.76	119.29
21	379.88	1436.11	14.14	143.05	48.83	26.65	120.32
20	391.41	1441.86	13.79	148.55	49.09	26.42	122.21
19	404.58	1447.60	11.97	154.09	49.34	26.06	124.84
18	415.35	1453.34	11.71	158.10	49.53	25.68	127.37
17	421.49	1459.09	11.57	160.18	49.63	25.42	128.95
16	424.62	1464.83	11.51	161.16	49.68	25.28	129.74
15	426.35	1470.58	11.47	161.68	49.70	25.21	130.11
14	427.77	1476.32	11.44	162.18	49.74	25.18	130.30
13	429.94	1482.07	11.39	163.15	49.81	25.15	130.49
12	434.55	1487.81	11.26	165.38	49.98	25.09	130.86
11	444.20	1493.56	10.98	169.97	50.34	24.97	131.70
10	460.61	1499.30	7.04	177.27	50.86	24.69	133.36
9	482.04	1505.04	6.67	186.67	51.42	24.23	135.85
8	499.11	1510.79	6.38	194.95	51.80	23.86	137.83
7	513.82	1516.53	6.13	202.57	52.09	23.57	139.47
6	526.34	1522.28	5.93	209.10	52.32	23.33	140.89
5	537.24	1528.02	5.71	215.65	52.57	23.12	142.02
4	551.73	1533.77	5.04	229.35	53.05	22.91	143.02
3	609.38	1539.51	9.81	352.77	56.36	22.23	143.92
2	632.24	1545.26	9.71	354.83	56.45	0.24	160.23
1	649.47	1551.00	9.57	357.54	56.55	0.14	172.14

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
46	0.0005	0.0211	0.0166	0.0854	0.0943	0.1364	0.2812
45	0.0000	0.0015	0.0020	0.0128	0.0278	0.0484	0.2383
44	0.0000	0.0006	0.0006	0.0037	0.0091	0.0170	0.1449
43	0.0000	0.0005	0.0005	0.0027	0.0053	0.0094	0.0870
42	0.0000	0.0005	0.0005	0.0026	0.0045	0.0077	0.0586
41	0.0000	0.0005	0.0005	0.0025	0.0043	0.0072	0.0455
40	0.0000	0.0005	0.0004	0.0024	0.0041	0.0068	0.0385
39	0.0000	0.0005	0.0004	0.0023	0.0038	0.0063	0.0331
38	0.0000	0.0004	0.0004	0.0021	0.0034	0.0055	0.0270
37	0.0000	0.0004	0.0003	0.0018	0.0028	0.0045	0.0203
36	0.0000	0.0003	0.0003	0.0015	0.0022	0.0035	0.0144
35	0.0000	0.0003	0.0002	0.0013	0.0019	0.0029	0.0112
34	0.0000	0.0003	0.0002	0.0012	0.0017	0.0027	0.0098
33	0.0000	0.0003	0.0002	0.0011	0.0017	0.0025	0.0092
32	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0090
31	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
30	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
29	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
28	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
27	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
26	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
25	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
24	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
23	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0088
22	0.0000	0.0003	0.0002	0.0011	0.0016	0.0024	0.0085
21	0.0000	0.0003	0.0002	0.0011	0.0015	0.0023	0.0080
20	0.0000	0.0002	0.0002	0.0010	0.0014	0.0021	0.0073
19	0.0000	0.0002	0.0002	0.0009	0.0013	0.0020	0.0066
18	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0062
17	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0059
16	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
15	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
14	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
13	0.0000	0.0002	0.0002	0.0009	0.0012	0.0017	0.0057
12	0.0000	0.0002	0.0002	0.0008	0.0011	0.0017	0.0055
11	0.0000	0.0002	0.0002	0.0008	0.0011	0.0016	0.0052
10	0.0000	0.0002	0.0002	0.0007	0.0010	0.0015	0.0047
9	0.0000	0.0002	0.0001	0.0007	0.0009	0.0013	0.0041
8	0.0000	0.0002	0.0001	0.0006	0.0008	0.0012	0.0037
7	0.0000	0.0002	0.0001	0.0006	0.0008	0.0011	0.0035
6	0.0000	0.0002	0.0001	0.0006	0.0007	0.0011	0.0032
5	0.0000	0.0002	0.0001	0.0006	0.0007	0.0010	0.0031
4	0.0000	0.0002	0.0001	0.0005	0.0007	0.0010	0.0029
3	0.0000	0.0001	0.0001	0.0004	0.0005	0.0008	0.0022
2	0.0000	0.0000	0.0000	0.0002	0.0003	0.0004	0.0014
1	0.0000	0.0000	0.0000	0.0001	0.0001	0.0002	0.0009

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	8	9	10	11	12	13	14
46	0.1180	0.2463	0.0001	0.0000	0.0000	0.0000	0.0000
45	0.1346	0.5338	0.0007	0.0000	0.0000	0.0000	0.0000
44	0.1057	0.7159	0.0025	0.0000	0.0000	0.0000	0.0000
43	0.0758	0.8117	0.0071	0.0000	0.0000	0.0000	0.0000
42	0.0549	0.8520	0.0187	0.0000	0.0000	0.0000	0.0000
41	0.0418	0.8511	0.0466	0.0002	0.0000	0.0000	0.0000
40	0.0332	0.8039	0.1089	0.0012	0.0000	0.0000	0.0000
39	0.0266	0.6918	0.2274	0.0079	0.0000	0.0000	0.0000
38	0.0204	0.5082	0.3906	0.0422	0.0000	0.0000	0.0000
37	0.0144	0.2965	0.4987	0.1602	0.0000	0.0000	0.0000
36	0.0097	0.1344	0.4386	0.3951	0.0000	0.0000	0.0000
35	0.0072	0.0663	0.2956	0.6130	0.0000	0.0000	0.0000
34	0.0062	0.0441	0.1860	0.7479	0.0000	0.0000	0.0000
33	0.0057	0.0374	0.1275	0.8143	0.0000	0.0000	0.0000
32	0.0056	0.0353	0.1007	0.8437	0.0000	0.0000	0.0000
31	0.0055	0.0346	0.0892	0.8561	0.0000	0.0000	0.0000
30	0.0055	0.0343	0.0844	0.8611	0.0001	0.0000	0.0000
29	0.0055	0.0343	0.0824	0.8630	0.0001	0.0000	0.0000
28	0.0055	0.0343	0.0817	0.8635	0.0004	0.0000	0.0000
27	0.0055	0.0343	0.0814	0.8630	0.0011	0.0000	0.0000
26	0.0055	0.0343	0.0812	0.8613	0.0030	0.0000	0.0000
25	0.0055	0.0342	0.0810	0.8569	0.0077	0.0000	0.0000
24	0.0055	0.0341	0.0804	0.8461	0.0194	0.0000	0.0000
23	0.0054	0.0336	0.0789	0.8200	0.0476	0.0000	0.0000
22	0.0053	0.0325	0.0755	0.7622	0.1104	0.0000	0.0000
21	0.0050	0.0304	0.0691	0.6533	0.2290	0.0000	0.0000
20	0.0045	0.0273	0.0599	0.4974	0.3986	0.0000	0.0000
19	0.0041	0.0243	0.0508	0.3402	0.5693	0.0001	0.0000
18	0.0038	0.0222	0.0443	0.2259	0.6930	0.0002	0.0000
17	0.0037	0.0211	0.0410	0.1667	0.7565	0.0008	0.0000
16	0.0036	0.0206	0.0396	0.1395	0.7844	0.0022	0.0000
15	0.0036	0.0204	0.0389	0.1275	0.7933	0.0064	0.0000
14	0.0036	0.0203	0.0384	0.1216	0.7884	0.0177	0.0000
13	0.0035	0.0200	0.0378	0.1172	0.7636	0.0480	0.0001
12	0.0034	0.0193	0.0362	0.1104	0.6989	0.1212	0.0009
11	0.0032	0.0180	0.0332	0.0979	0.5682	0.2643	0.0062
10	0.0029	0.0160	0.0286	0.0799	0.3810	0.4498	0.0335
9	0.0025	0.0138	0.0238	0.0615	0.2060	0.5579	0.1268
8	0.0023	0.0123	0.0207	0.0508	0.1258	0.5068	0.2724
7	0.0021	0.0113	0.0185	0.0438	0.0910	0.3875	0.4303
6	0.0020	0.0105	0.0170	0.0390	0.0739	0.2765	0.5431
5	0.0019	0.0099	0.0157	0.0355	0.0638	0.1997	0.5731
4	0.0018	0.0091	0.0143	0.0315	0.0542	0.1441	0.4857
3	0.0013	0.0067	0.0099	0.0204	0.0313	0.0674	0.1853
2	0.0010	0.0054	0.0088	0.0192	0.0305	0.0671	0.1861
1	0.0007	0.0041	0.0073	0.0174	0.0289	0.0659	0.1861

IRANIAN CRUDE

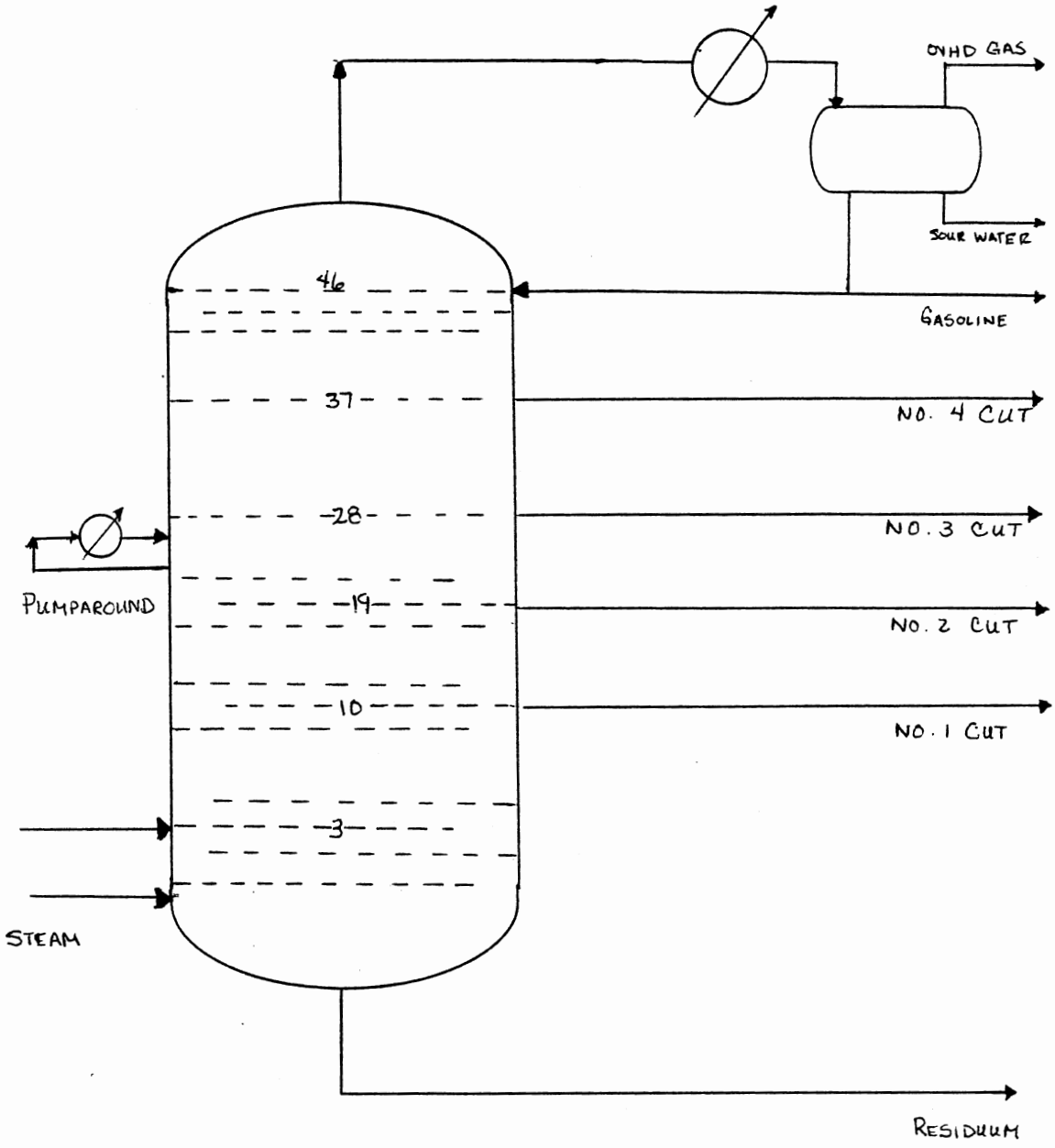
LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	15	16	17	18	19	20	21
46	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
45	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
44	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
43	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
42	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
41	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
40	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
39	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
38	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
37	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
36	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
35	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
34	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
33	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
32	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
31	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
30	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
29	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
28	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
27	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
25	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
23	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
21	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
19	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
17	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
16	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
15	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
13	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
11	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
9	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
8	0.0021	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
7	0.0092	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
6	0.0314	0.0006	0.0000	0.0000	0.0000	0.0000	0.0000
5	0.0882	0.0062	0.0003	0.0000	0.0000	0.0000	0.0000
4	0.1946	0.0487	0.0087	0.0018	0.0001	0.0000	0.0000
3	0.1747	0.1388	0.0869	0.0772	0.0691	0.0671	0.0598
2	0.1760	0.1400	0.0877	0.0780	0.0698	0.0678	0.0603
1	0.1776	0.1418	0.0889	0.0791	0.0709	0.0688	0.0612

CASE STUDY IV

CRUDE COLUMN WITH PUMPAROUND

PROGRAM OUTPUT



CRUDE COLUMN WITH PUMPAROUND

FEEDS

FEED #	1
TRAY	3
COMP	
C2H6	0.0303
C3H8	0.4670
I-C4H10	0.1980
N-C4H10	0.8620
I-C5H12	0.6810
N-C5H12	0.9240
N-C6H14	1.6700
FR1	0.7220
FR2	2.7200
FR3	2.5220
FR4	2.8620
FR5	2.4210
FR6	2.2000
FR7	1.8980
FR8	1.6990
FR9	1.3570
FR10	0.8510
FR11	0.7570
FR12	0.6780
FR13	0.6580
FR14	0.5860
H2O	0.0000
RATE, LBMOLS	26.7633
TEMP, F	584.9500
ENTHALPY, BTU	926972.06
L/F	0.2880
MOL WT.	204.2151
DENSITY,LB/FT3	49.5434

STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS 5.8100
TEMPERATURE, F : 700.0000

IRANIAN CRUDE

PRODUCTS

COMP	BOTTOMS	DISTILLATE
C2H6	0.0000	0.0279
C3H8	0.0001	0.3552
I-C4H10	0.0001	0.1108
N-C4H10	0.0006	0.4154
I-C5H12	0.0012	0.1881
N-C5H12	0.0019	0.2106
N-C6H14	0.0084	0.1458
FR1	0.0068	0.0467
FR2	0.0395	0.0426
FR3	0.0708	0.0000
FR4	0.1683	0.0000
FR5	0.2810	0.0000
FR6	0.6518	0.0000
FR7	1.8147	0.0000
FR8	1.6989	0.0000
FR9	1.3570	0.0000
FR10	0.8510	0.0000
FR11	0.7570	0.0000
FR12	0.6780	0.0000
FR13	0.6580	0.0000
FR14	0.5860	0.0000
RATE, LBMOLS	9.6311	1.5432
TEMP, F	648.5800	122.0831
L/F	1.0000	0.0000
MOL WT.	356.5687	62.8894
DENSITY,LB/FT3	56.5205	

REBOILER DUTY, BTU : 0.0000
CONDENSER DUTY, BTU : -526913.5625

SOUR WATER PRODUCED, LBMOLS: 5.7119
OVERHEAD STEAM, LBMOLS: 0.0981

IRANIAN CRUDE

SIDE DRAWS

DRAW #	1	2	3
TRAY	46	37	10
COMP #			
C2H6	0.0024	0.0000	0.0000
C3H8	0.1086	0.0017	0.0007
I-C4H10	0.0845	0.0015	0.0005
N-C4H10	0.4326	0.0077	0.0026
I-C5H12	0.4717	0.0123	0.0035
N-C5H12	0.6801	0.0198	0.0052
N-C6H14	1.3866	0.0902	0.0165
FR1	0.5800	0.0643	0.0102
FR2	1.1559	1.3409	0.0563
FR3	0.0005	2.1569	0.1012
FR4	0.0000	0.6397	0.2840
FR5	0.0000	0.0000	1.3770
FR6	0.0000	0.0000	1.5480
FR7	0.0000	0.0000	0.0833
FR8	0.0000	0.0000	0.0001
FR9	0.0000	0.0000	0.0000
FR10	0.0000	0.0000	0.0000
FR11	0.0000	0.0000	0.0000
FR12	0.0000	0.0000	0.0000
FR13	0.0000	0.0000	0.0000
FR14	0.0000	0.0000	0.0000
RATE, LBMOLS	4.9030	4.3350	3.4890
TEMP, F	122.0831	284.6060	459.0972
L/F	1.0000	1.0000	1.0000
MOL WT.	83.6210	114.3481	176.4546
DENSITY,LB/FT3	41.6203	46.9903	50.8111

IRANIAN CRUDE

SIDE DRAWS

DRAW #	4	5
TRAY	19	28
COMP #		
C2H6	0.0000	0.0000
C3H8	0.0003	0.0004
I-C4H10	0.0003	0.0003
N-C4H10	0.0014	0.0016
I-C5H12	0.0019	0.0023
N-C5H12	0.0029	0.0036
N-C6H14	0.0097	0.0128
FR1	0.0060	0.0079
FR2	0.0357	0.0492
FR3	0.0755	0.1173
FR4	0.5355	1.2350
FR5	0.7618	0.0005
FR6	0.0001	0.0000
FR7	0.0000	0.0000
FR8	0.0000	0.0000
FR9	0.0000	0.0000
FR10	0.0000	0.0000
FR11	0.0000	0.0000
FR12	0.0000	0.0000
FR13	0.0000	0.0000
FR14	0.0000	0.0000
RATE, LBMOLS	1.4310	1.4310
TEMP, F	401.6382	364.4160
L/F	1.0000	1.0000
MOL WT.	152.8936	135.6307
DENSITY,LB/FT3	49.2888	48.4814

PUMPAROUND 1 FROM TRAY 22 TO 23
RATE, LBMOLS: 2.0000 COOLING, BTU: 1000.0000

IRANIAN CRUDE

TEMPERATURE, LIQUID AND VAPOR PROFILES

STREAMS LEAVING TRAY

TRAY NO	TEMP DEGF	PRES mmHg	LIQUID LB-MOLS	MOL WT	DENS LB/FT3	VAPOR LB-MOLS	MOL WT
46	122.08	1292.50	27.30	83.62	41.62	1.54	62.89
45	195.90	1298.24	27.04	94.42	44.01	33.75	82.67
44	221.88	1303.99	26.32	98.74	44.95	33.48	91.39
43	232.32	1309.73	25.88	100.73	45.35	32.76	94.79
42	237.50	1315.48	25.64	101.79	45.54	32.33	96.33
41	241.07	1321.22	25.48	102.57	45.67	32.09	97.14
40	245.23	1326.97	25.28	103.59	45.81	31.92	97.74
39	252.26	1332.71	24.92	105.40	46.05	31.72	98.52
38	264.72	1338.46	24.26	108.71	46.44	31.36	99.90
37	284.61	1344.20	18.88	114.35	46.99	30.71	102.40
36	311.57	1349.94	17.78	122.24	47.62	29.66	106.60
35	334.70	1355.69	16.93	128.63	48.05	28.56	111.21
34	349.49	1361.43	16.43	132.42	48.29	27.71	114.78
33	357.17	1367.18	16.19	134.26	48.40	27.21	116.82
32	360.79	1372.92	16.09	135.07	48.45	26.97	117.78
31	362.52	1378.67	16.04	135.41	48.47	26.87	118.20
30	363.43	1384.41	16.02	135.55	48.48	26.82	118.37
29	363.99	1390.16	16.01	135.60	48.48	26.80	118.44
28	364.42	1395.90	14.57	135.63	48.48	26.79	118.47
27	364.80	1401.64	14.57	135.66	48.48	26.79	118.48
26	365.22	1407.39	14.56	135.71	48.48	26.78	118.49
25	365.77	1413.13	14.56	135.85	48.49	26.78	118.52
24	366.65	1418.88	14.54	136.19	48.51	26.77	118.58
23	368.38	1424.62	16.57	137.04	48.55	26.75	118.75
22	371.31	1430.37	14.47	138.53	48.62	26.78	119.12
21	377.69	1436.11	14.28	141.90	48.77	26.69	119.97
20	388.36	1441.86	13.94	147.13	49.02	26.49	121.64
19	401.64	1447.60	12.12	152.89	49.29	26.16	124.17
18	413.39	1453.34	11.83	157.38	49.50	25.76	126.85
17	420.45	1459.09	11.67	159.82	49.61	25.47	128.64
16	424.13	1464.83	11.59	160.99	49.67	25.31	129.58
15	426.10	1470.58	11.55	161.60	49.70	25.23	130.03
14	427.60	1476.32	11.51	162.11	49.73	25.19	130.26
13	429.71	1482.07	11.46	163.03	49.80	25.16	130.46
12	434.09	1487.81	11.34	165.14	49.97	25.10	130.81
11	443.31	1493.56	11.07	169.52	50.30	24.98	131.62
10	459.10	1499.30	7.14	176.45	50.81	24.71	133.22
9	479.40	1505.04	6.80	185.06	51.36	24.28	135.60
8	495.15	1510.79	6.52	192.56	51.71	23.93	137.45
7	509.21	1516.53	6.28	200.03	52.00	23.66	138.97
6	522.41	1522.28	6.05	207.11	52.25	23.41	140.41
5	534.58	1528.02	5.81	214.32	52.52	23.18	141.68
4	550.05	1533.77	5.12	228.33	53.02	22.95	142.83
3	608.29	1539.51	9.87	351.74	56.33	22.26	143.84
2	631.34	1545.26	9.78	353.83	56.42	0.24	160.28
1	648.58	1551.00	9.63	356.57	56.52	0.15	172.17

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
46	0.0005	0.0222	0.0172	0.0882	0.0962	0.1387	0.2828
45	0.0000	0.0015	0.0021	0.0133	0.0287	0.0499	0.2443
44	0.0000	0.0006	0.0006	0.0038	0.0093	0.0175	0.1496
43	0.0000	0.0005	0.0005	0.0028	0.0054	0.0096	0.0896
42	0.0000	0.0005	0.0005	0.0026	0.0046	0.0077	0.0600
41	0.0000	0.0005	0.0005	0.0025	0.0043	0.0072	0.0462
40	0.0000	0.0005	0.0004	0.0024	0.0041	0.0068	0.0390
39	0.0000	0.0005	0.0004	0.0023	0.0038	0.0063	0.0335
38	0.0000	0.0004	0.0004	0.0021	0.0034	0.0056	0.0275
37	0.0000	0.0004	0.0003	0.0018	0.0028	0.0046	0.0208
36	0.0000	0.0003	0.0003	0.0015	0.0023	0.0036	0.0148
35	0.0000	0.0003	0.0002	0.0013	0.0019	0.0030	0.0115
34	0.0000	0.0003	0.0002	0.0012	0.0017	0.0027	0.0099
33	0.0000	0.0003	0.0002	0.0012	0.0017	0.0026	0.0093
32	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0090
31	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
30	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
29	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
28	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
27	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
26	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
25	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
24	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
23	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0088
22	0.0000	0.0003	0.0002	0.0011	0.0016	0.0024	0.0086
21	0.0000	0.0003	0.0002	0.0011	0.0015	0.0023	0.0082
20	0.0000	0.0003	0.0002	0.0010	0.0014	0.0022	0.0075
19	0.0000	0.0002	0.0002	0.0009	0.0013	0.0020	0.0068
18	0.0000	0.0002	0.0002	0.0009	0.0012	0.0019	0.0063
17	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0060
16	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0059
15	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
14	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
13	0.0000	0.0002	0.0002	0.0009	0.0012	0.0017	0.0057
12	0.0000	0.0002	0.0002	0.0008	0.0011	0.0017	0.0056
11	0.0000	0.0002	0.0002	0.0008	0.0011	0.0016	0.0052
10	0.0000	0.0002	0.0002	0.0007	0.0010	0.0015	0.0047
9	0.0000	0.0002	0.0001	0.0007	0.0009	0.0013	0.0042
8	0.0000	0.0002	0.0001	0.0006	0.0008	0.0012	0.0038
7	0.0000	0.0002	0.0001	0.0006	0.0008	0.0012	0.0035
6	0.0000	0.0002	0.0001	0.0006	0.0007	0.0011	0.0033
5	0.0000	0.0002	0.0001	0.0006	0.0007	0.0010	0.0031
4	0.0000	0.0002	0.0001	0.0005	0.0007	0.0010	0.0029
3	0.0000	0.0001	0.0001	0.0004	0.0005	0.0008	0.0022
2	0.0000	0.0000	0.0000	0.0002	0.0003	0.0004	0.0014
1	0.0000	0.0000	0.0000	0.0001	0.0001	0.0002	0.0009

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	8	9	10	11	12	13	14
46	0.1183	0.2357	0.0001	0.0000	0.0000	0.0000	0.0000
45	0.1375	0.5219	0.0006	0.0000	0.0000	0.0000	0.0000
44	0.1088	0.7075	0.0022	0.0000	0.0000	0.0000	0.0000
43	0.0781	0.8070	0.0065	0.0000	0.0000	0.0000	0.0000
42	0.0565	0.8503	0.0174	0.0000	0.0000	0.0000	0.0000
41	0.0428	0.8523	0.0436	0.0001	0.0000	0.0000	0.0000
40	0.0339	0.8090	0.1027	0.0010	0.0000	0.0000	0.0000
39	0.0272	0.7019	0.2171	0.0069	0.0000	0.0000	0.0000
38	0.0209	0.5220	0.3800	0.0377	0.0000	0.0000	0.0000
37	0.0148	0.3093	0.4976	0.1476	0.0000	0.0000	0.0000
36	0.0099	0.1417	0.4499	0.3757	0.0000	0.0000	0.0000
35	0.0074	0.0693	0.3083	0.5968	0.0000	0.0000	0.0000
34	0.0062	0.0452	0.1943	0.7383	0.0000	0.0000	0.0000
33	0.0058	0.0378	0.1318	0.8094	0.0000	0.0000	0.0000
32	0.0056	0.0355	0.1028	0.8413	0.0000	0.0000	0.0000
31	0.0056	0.0347	0.0902	0.8548	0.0000	0.0000	0.0000
30	0.0055	0.0345	0.0850	0.8603	0.0000	0.0000	0.0000
29	0.0055	0.0344	0.0828	0.8625	0.0001	0.0000	0.0000
28	0.0055	0.0344	0.0820	0.8631	0.0004	0.0000	0.0000
27	0.0055	0.0344	0.0817	0.8627	0.0010	0.0000	0.0000
26	0.0055	0.0344	0.0815	0.8612	0.0027	0.0000	0.0000
25	0.0055	0.0344	0.0813	0.8571	0.0070	0.0000	0.0000
24	0.0055	0.0342	0.0807	0.8471	0.0178	0.0000	0.0000
23	0.0055	0.0338	0.0794	0.8229	0.0440	0.0000	0.0000
22	0.0053	0.0329	0.0768	0.7810	0.0898	0.0000	0.0000
21	0.0051	0.0310	0.0711	0.6858	0.1935	0.0000	0.0000
20	0.0047	0.0281	0.0623	0.5377	0.3547	0.0000	0.0000
19	0.0042	0.0249	0.0527	0.3742	0.5324	0.0001	0.0000
18	0.0039	0.0225	0.0454	0.2464	0.6709	0.0002	0.0000
17	0.0037	0.0213	0.0416	0.1770	0.7455	0.0007	0.0000
16	0.0036	0.0207	0.0398	0.1442	0.7796	0.0020	0.0000
15	0.0036	0.0205	0.0390	0.1296	0.7916	0.0058	0.0000
14	0.0036	0.0203	0.0385	0.1226	0.7887	0.0163	0.0000
13	0.0035	0.0200	0.0379	0.1178	0.7663	0.0445	0.0001
12	0.0034	0.0194	0.0364	0.1112	0.7055	0.1139	0.0006
11	0.0032	0.0181	0.0335	0.0991	0.5795	0.2533	0.0043
10	0.0029	0.0161	0.0290	0.0814	0.3947	0.4437	0.0239
9	0.0026	0.0140	0.0243	0.0634	0.2180	0.5751	0.0949
8	0.0024	0.0127	0.0214	0.0529	0.1351	0.5491	0.2181
7	0.0022	0.0116	0.0192	0.0458	0.0977	0.4373	0.3723
6	0.0020	0.0107	0.0174	0.0405	0.0783	0.3147	0.5024
5	0.0019	0.0100	0.0160	0.0364	0.0662	0.2212	0.5549
4	0.0018	0.0092	0.0145	0.0319	0.0554	0.1534	0.4840
3	0.0013	0.0067	0.0100	0.0205	0.0316	0.0692	0.1875
2	0.0010	0.0054	0.0088	0.0194	0.0308	0.0689	0.1883
1	0.0007	0.0041	0.0073	0.0175	0.0292	0.0677	0.1884

IRANIAN CRUDE

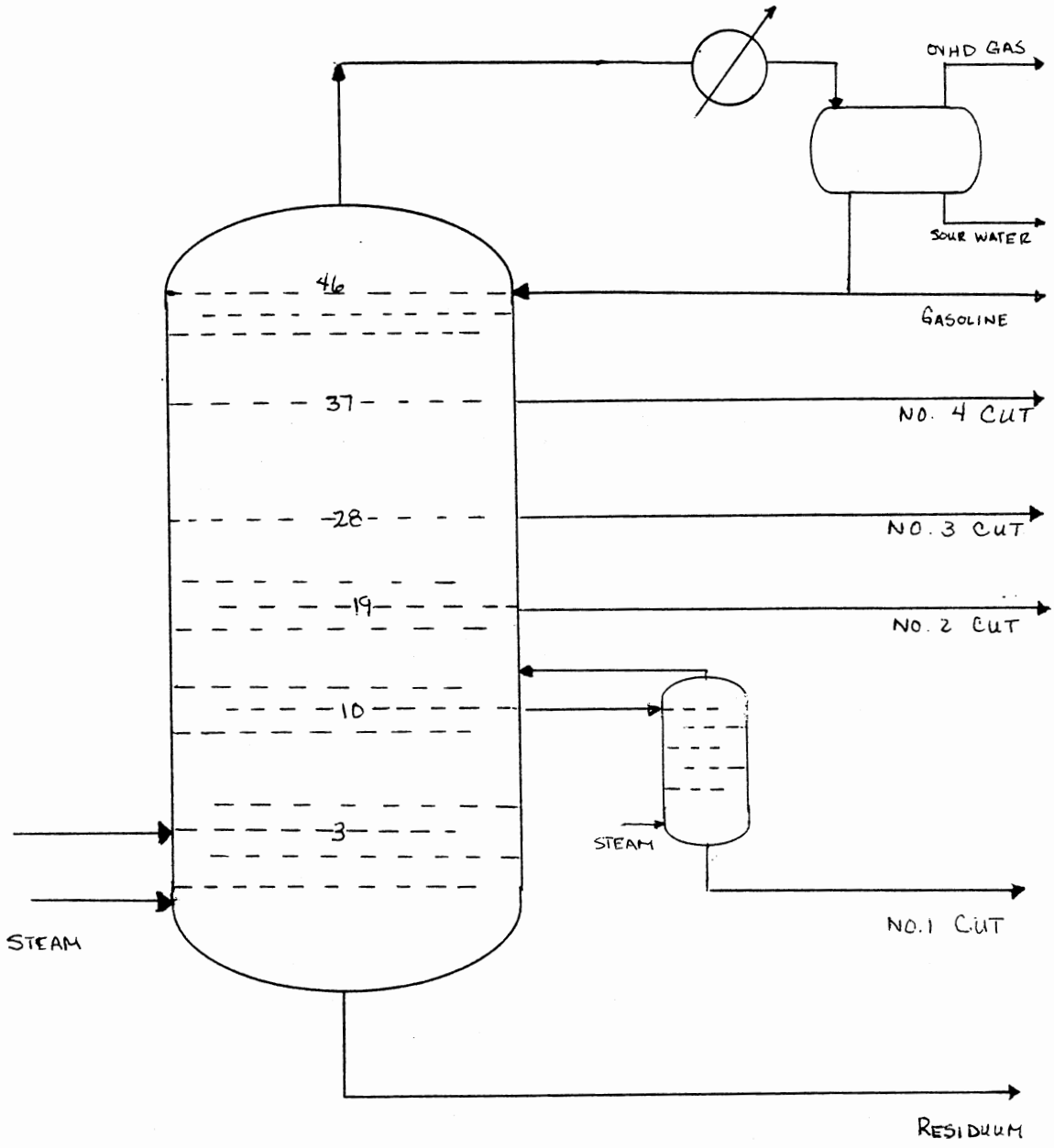
LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	15	16	17	18	19	20	21
46	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
45	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
44	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
43	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
42	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
41	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
40	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
39	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
38	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
37	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
36	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
35	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
34	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
33	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
32	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
31	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
30	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
29	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
28	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
27	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
25	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
23	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
21	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
19	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
17	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
16	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
15	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
13	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
11	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
9	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
8	0.0016	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
7	0.0074	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
6	0.0274	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000
5	0.0816	0.0057	0.0003	0.0000	0.0000	0.0000	0.0000
4	0.1875	0.0468	0.0083	0.0017	0.0001	0.0000	0.0000
3	0.1735	0.1378	0.0863	0.0767	0.0687	0.0666	0.0593
2	0.1749	0.1391	0.0871	0.0774	0.0694	0.0673	0.0599
1	0.1764	0.1409	0.0884	0.0786	0.0704	0.0683	0.0608

CASE STUDY V

CRUDE COLUMN WITH STRIPPER

PROGRAM OUTPUT



CRUDE COLUMN WITH STRIPPER

FEEDS

FEED #	1
TRAY	3
COMP	
C2H6	0.0303
C3H8	0.4670
I-C4H10	0.1980
N-C4H10	0.8620
I-C5H12	0.6810
N-C5H12	0.9240
N-C6H14	1.6700
FR1	0.7220
FR2	2.7200
FR3	2.5220
FR4	2.8620
FR5	2.4210
FR6	2.2000
FR7	1.8980
FR8	1.6990
FR9	1.3570
FR10	0.8510
FR11	0.7570
FR12	0.6780
FR13	0.6580
FR14	0.5860
H2O	0.0000
RATE, LBMOLS	26.7633
TEMP, F	584.9500
ENTHALPY, BTU	926972.06
L/F	0.2880
MOL WT.	204.2151
DENSITY,LB/FT3	49.5434

STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS 5.8100
TEMPERATURE, F : 700.0000

IRANIAN CRUDE

PRODUCTS

	BOTTOMS	DISTILLATE
COMP		
C2H6	0.0000	0.0284
C3H8	0.0001	0.3763
I-C4H10	0.0001	0.1240
N-C4H10	0.0006	0.4765
I-C5H12	0.0013	0.2312
N-C5H12	0.0019	0.2641
N-C6H14	0.0084	0.1959
FR1	0.0067	0.0634
FR2	0.0387	0.0710
FR3	0.0684	0.0000
FR4	0.1607	0.0000
FR5	0.2649	0.0000
FR6	0.5883	0.0000
FR7	1.6683	0.0000
FR8	1.6987	0.0000
FR9	1.3570	0.0000
FR10	0.8510	0.0000
FR11	0.7570	0.0000
FR12	0.6780	0.0000
FR13	0.6580	0.0000
FR14	0.5860	0.0000
RATE, LBMOLS	9.3942	1.8308
TEMP, F	651.6321	129.0918
L/F	1.0000	0.0000
MOL WT.	360.2112	64.4844
DENSITY,LB/FT3	56.6215	

REBOILER DUTY, BTU : 0.0000
CONDENSER DUTY, BTU : -547262.5000

SOUR WATER PRODUCED, LBMOLS: 6.8772
OVERHEAD STEAM, LBMOLS: 0.1428

IRANIAN CRUDE

SIDE DRAWS

DRAW #	1	2	3
TRAY	46	37	19
COMP #			
C2H6	0.0019	0.0000	0.0000
C3H8	0.0882	0.0016	0.0003
I-C4H10	0.0717	0.0014	0.0003
N-C4H10	0.3736	0.0073	0.0013
I-C5H12	0.4309	0.0115	0.0018
N-C5H12	0.6302	0.0184	0.0026
N-C6H14	1.3502	0.0816	0.0089
FR1	0.5732	0.0571	0.0055
FR2	1.3823	1.0992	0.0318
FR3	0.0008	2.1825	0.0639
FR4	0.0000	0.8744	0.3355
FR5	0.0000	0.0000	0.9790
FR6	0.0000	0.0000	0.0002
FR7	0.0000	0.0000	0.0000
FR8	0.0000	0.0000	0.0000
FR9	0.0000	0.0000	0.0000
FR10	0.0000	0.0000	0.0000
FR11	0.0000	0.0000	0.0000
FR12	0.0000	0.0000	0.0000
FR13	0.0000	0.0000	0.0000
FR14	0.0000	0.0000	0.0000
RATE, LBMOLS	4.9030	4.3350	1.4310
TEMP, F	129.0918	291.7382	414.2675
L/F	1.0000	1.0000	1.0000
MOL WT.	85.2922	116.4504	157.8131
DENSITY,LB/FT3	41.9854	47.1708	49.5189

IRANIAN CRUDE

SIDE DRAWS

DRAW #	4
TRAY	28
COMP #	
C2H6	0.0000
C3H8	0.0004
I-C4H10	0.0003
N-C4H10	0.0016
I-C5H12	0.0023
N-C5H12	0.0036
N-C6H14	0.0127
FR1	0.0079
FR2	0.0489
FR3	0.1163
FR4	1.2354
FR5	0.0016
FR6	0.0000
FR7	0.0000
FR8	0.0000
FR9	0.0000
FR10	0.0000
FR11	0.0000
FR12	0.0000
FR13	0.0000
FR14	0.0000
RATE, LBMOLS	1.4310
TEMP, F	364.6549
L/F	1.0000
MOL WT.	135.6852
DENSITY,LB/FT3	48.4848

SIDESTREAM STRIPPERS

STRIPPER #	1
TRAY	10
NO. TRAYS	3
COMP #	
C2H6	0.0000
C3H8	0.0001
I-C4H10	0.0002
N-C4H10	0.0011
I-C5H12	0.0021
N-C5H12	0.0033
N-C6H14	0.0125
FR1	0.0083
FR2	0.0480
FR3	0.0899
FR4	0.2558
FR5	1.1748
FR6	1.6116
FR7	0.2305
FR8	0.0003
FR9	0.0000
FR10	0.0000
FR11	0.0000
FR12	0.0000
FR13	0.0000
FR14	0.0000
TEMP, F	475.5463
RATE, LBMOLS	3.4384
L/F	1.0000
MOL WT.	180.4853
DENSITY,LB/FT3	51.0524
STEAM, LBMOLS	1.2100
TEMP, F	500.0000
PUMPBACK	0.0000
TRAY	0

IRANIAN CRUDE

TEMPERATURE, LIQUID AND VAPOR PROFILES

STREAMS LEAVING TRAY

TRAY NO	TEMP DEGF	PRES mmHg	LIQUID LB-MOLS	MOL WT	DENS LB/FT3	VAPOR LB-MOLS	MOL WT
46	129.09	1292.50	27.30	85.29	41.99	1.83	64.48
45	201.05	1298.24	26.90	95.48	44.23	34.03	84.17
44	224.80	1303.99	26.18	99.37	45.07	33.64	92.31
43	234.03	1309.73	25.78	101.11	45.42	32.91	95.34
42	238.69	1315.48	25.56	102.07	45.59	32.52	96.66
41	242.32	1321.22	25.39	102.89	45.71	32.30	97.39
40	247.18	1326.97	25.15	104.10	45.88	32.12	98.01
39	255.63	1332.71	24.70	106.28	46.16	31.88	98.93
38	270.00	1338.46	23.93	110.15	46.59	31.44	100.58
37	291.74	1344.20	18.45	116.45	47.17	30.67	103.45
36	319.31	1349.94	17.35	124.48	47.78	29.52	108.05
35	340.66	1355.69	16.58	130.23	48.16	28.41	112.63
34	352.98	1361.43	16.18	133.29	48.34	27.65	115.75
33	358.95	1367.18	16.00	134.68	48.43	27.25	117.35
32	361.69	1372.92	15.92	135.27	48.46	27.07	118.07
31	363.01	1378.67	15.89	135.51	48.48	26.99	118.37
30	363.74	1384.41	15.88	135.61	48.48	26.96	118.49
29	364.24	1390.16	15.87	135.65	48.48	26.94	118.54
28	364.65	1395.90	14.43	135.69	48.48	26.94	118.56
27	365.10	1401.64	14.43	135.75	48.49	26.93	118.57
26	365.68	1407.39	14.42	135.91	48.49	26.92	118.61
25	366.65	1413.13	14.39	136.30	48.51	26.91	118.68
24	368.54	1418.88	14.34	137.24	48.56	26.89	118.88
23	372.40	1424.62	14.23	139.30	48.65	26.84	119.35
22	379.74	1430.37	14.00	143.12	48.83	26.73	120.37
21	391.04	1436.11	13.65	148.51	49.08	26.50	122.22
20	403.82	1441.86	13.27	153.90	49.34	26.15	124.76
19	414.27	1447.60	11.56	157.81	49.52	25.77	127.19
18	420.89	1453.34	11.41	160.07	49.62	25.49	128.88
17	424.20	1459.09	11.34	161.11	49.67	25.34	129.73
16	425.88	1464.83	11.31	161.59	49.70	25.27	130.11
15	427.00	1470.58	11.28	161.92	49.72	25.24	130.29
14	428.30	1476.32	11.25	162.41	49.75	25.21	130.41
13	430.75	1482.07	11.19	163.55	49.84	25.19	130.59
12	436.15	1487.81	11.03	166.20	50.05	25.12	131.01
11	447.14	1493.56	10.72	171.46	50.44	24.96	131.98
10	465.38	1499.30	6.72	179.84	51.00	24.60	133.82
9	489.67	1505.04	6.31	191.12	51.61	24.09	136.59
8	508.81	1510.79	6.01	200.40	52.00	23.68	138.85
7	523.01	1516.53	5.80	207.32	52.25	23.38	140.56
6	533.06	1522.28	5.65	212.42	52.44	23.17	141.75
5	541.63	1528.02	5.47	218.01	52.66	23.02	142.56
4	554.97	1533.77	4.80	231.67	53.14	22.84	143.34
3	612.06	1539.51	9.62	355.60	56.44	22.17	144.07
2	634.48	1545.26	9.53	357.58	56.53	0.22	160.07
1	651.63	1551.00	9.39	360.21	56.62	0.13	171.93

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
46	0.0004	0.0180	0.0146	0.0762	0.0879	0.1285	0.2754
45	0.0000	0.0013	0.0018	0.0112	0.0249	0.0436	0.2197
44	0.0000	0.0006	0.0006	0.0035	0.0083	0.0155	0.1311
43	0.0000	0.0005	0.0005	0.0027	0.0051	0.0090	0.0794
42	0.0000	0.0005	0.0005	0.0025	0.0044	0.0075	0.0548
41	0.0000	0.0005	0.0005	0.0025	0.0042	0.0070	0.0433
40	0.0000	0.0005	0.0004	0.0024	0.0040	0.0067	0.0369
39	0.0000	0.0005	0.0004	0.0022	0.0037	0.0061	0.0315
38	0.0000	0.0004	0.0004	0.0020	0.0032	0.0052	0.0253
37	0.0000	0.0004	0.0003	0.0017	0.0026	0.0042	0.0188
36	0.0000	0.0003	0.0003	0.0014	0.0021	0.0033	0.0135
35	0.0000	0.0003	0.0002	0.0012	0.0018	0.0028	0.0107
34	0.0000	0.0003	0.0002	0.0012	0.0017	0.0026	0.0096
33	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0091
32	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
31	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
30	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
29	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
28	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
27	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
26	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
25	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0088
24	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0087
23	0.0000	0.0003	0.0002	0.0011	0.0016	0.0024	0.0085
22	0.0000	0.0003	0.0002	0.0011	0.0015	0.0023	0.0080
21	0.0000	0.0002	0.0002	0.0010	0.0014	0.0021	0.0073
20	0.0000	0.0002	0.0002	0.0009	0.0013	0.0020	0.0067
19	0.0000	0.0002	0.0002	0.0009	0.0012	0.0019	0.0062
18	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0059
17	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
16	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
15	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
14	0.0000	0.0002	0.0002	0.0009	0.0012	0.0017	0.0057
13	0.0000	0.0002	0.0002	0.0008	0.0011	0.0017	0.0057
12	0.0000	0.0002	0.0002	0.0008	0.0011	0.0017	0.0055
11	0.0000	0.0002	0.0002	0.0008	0.0011	0.0016	0.0051
10	0.0000	0.0002	0.0001	0.0007	0.0010	0.0014	0.0045
9	0.0000	0.0002	0.0001	0.0007	0.0009	0.0013	0.0039
8	0.0000	0.0002	0.0001	0.0006	0.0008	0.0012	0.0035
7	0.0000	0.0002	0.0001	0.0006	0.0007	0.0011	0.0033
6	0.0000	0.0002	0.0001	0.0006	0.0007	0.0010	0.0031
5	0.0000	0.0002	0.0001	0.0005	0.0007	0.0010	0.0030
4	0.0000	0.0002	0.0001	0.0005	0.0007	0.0010	0.0028
3	0.0000	0.0001	0.0001	0.0004	0.0005	0.0008	0.0021
2	0.0000	0.0000	0.0000	0.0002	0.0003	0.0004	0.0014
1	0.0000	0.0000	0.0000	0.0001	0.0001	0.0002	0.0009

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	8	9	10	11	12	13	14
46	0.1169	0.2819	0.0002	0.0000	0.0000	0.0000	0.0000
45	0.1256	0.5709	0.0010	0.0000	0.0000	0.0000	0.0000
44	0.0962	0.7407	0.0034	0.0000	0.0000	0.0000	0.0000
43	0.0688	0.8246	0.0095	0.0000	0.0000	0.0000	0.0000
42	0.0503	0.8549	0.0245	0.0000	0.0000	0.0000	0.0000
41	0.0388	0.8432	0.0598	0.0003	0.0000	0.0000	0.0000
40	0.0310	0.7809	0.1352	0.0019	0.0000	0.0000	0.0000
39	0.0248	0.6504	0.2686	0.0118	0.0000	0.0000	0.0000
38	0.0188	0.4562	0.4299	0.0586	0.0000	0.0000	0.0000
37	0.0132	0.2536	0.5035	0.2017	0.0000	0.0000	0.0000
36	0.0089	0.1127	0.4064	0.4511	0.0000	0.0000	0.0000
35	0.0068	0.0583	0.2621	0.6556	0.0000	0.0000	0.0000
34	0.0060	0.0414	0.1654	0.7716	0.0000	0.0000	0.0000
33	0.0057	0.0364	0.1172	0.8258	0.0000	0.0000	0.0000
32	0.0055	0.0349	0.0959	0.8490	0.0000	0.0000	0.0000
31	0.0055	0.0344	0.0869	0.8585	0.0001	0.0000	0.0000
30	0.0055	0.0342	0.0833	0.8623	0.0001	0.0000	0.0000
29	0.0055	0.0342	0.0819	0.8635	0.0004	0.0000	0.0000
28	0.0055	0.0342	0.0813	0.8633	0.0011	0.0000	0.0000
27	0.0055	0.0342	0.0810	0.8616	0.0031	0.0000	0.0000
26	0.0055	0.0341	0.0807	0.8570	0.0081	0.0000	0.0000
25	0.0055	0.0339	0.0801	0.8457	0.0203	0.0000	0.0000
24	0.0054	0.0334	0.0786	0.8189	0.0493	0.0000	0.0000
23	0.0052	0.0323	0.0752	0.7604	0.1129	0.0000	0.0000
22	0.0050	0.0302	0.0688	0.6521	0.2307	0.0000	0.0000
21	0.0045	0.0273	0.0598	0.4992	0.3969	0.0000	0.0000
20	0.0041	0.0243	0.0510	0.3460	0.5633	0.0000	0.0000
19	0.0038	0.0223	0.0447	0.2344	0.6841	0.0001	0.0000
18	0.0037	0.0211	0.0411	0.1698	0.7538	0.0004	0.0000
17	0.0036	0.0206	0.0396	0.1404	0.7847	0.0011	0.0000
16	0.0036	0.0204	0.0389	0.1278	0.7963	0.0030	0.0000
15	0.0036	0.0203	0.0386	0.1223	0.7969	0.0084	0.0000
14	0.0035	0.0202	0.0382	0.1191	0.7861	0.0229	0.0000
13	0.0035	0.0199	0.0374	0.1153	0.7537	0.0602	0.0002
12	0.0034	0.0191	0.0357	0.1079	0.6767	0.1457	0.0020
11	0.0031	0.0176	0.0323	0.0943	0.5321	0.2982	0.0134
10	0.0028	0.0154	0.0274	0.0754	0.3410	0.4639	0.0659
9	0.0024	0.0131	0.0223	0.0566	0.1748	0.5063	0.2166
8	0.0022	0.0116	0.0192	0.0461	0.1054	0.4099	0.3952
7	0.0020	0.0107	0.0173	0.0401	0.0786	0.2958	0.5350
6	0.0019	0.0101	0.0162	0.0367	0.0670	0.2165	0.6028
5	0.0019	0.0096	0.0153	0.0341	0.0601	0.1682	0.5909
4	0.0017	0.0089	0.0140	0.0306	0.0521	0.1301	0.4764
3	0.0013	0.0066	0.0098	0.0200	0.0305	0.0641	0.1769
2	0.0010	0.0054	0.0087	0.0189	0.0298	0.0638	0.1776
1	0.0007	0.0041	0.0073	0.0171	0.0282	0.0626	0.1776

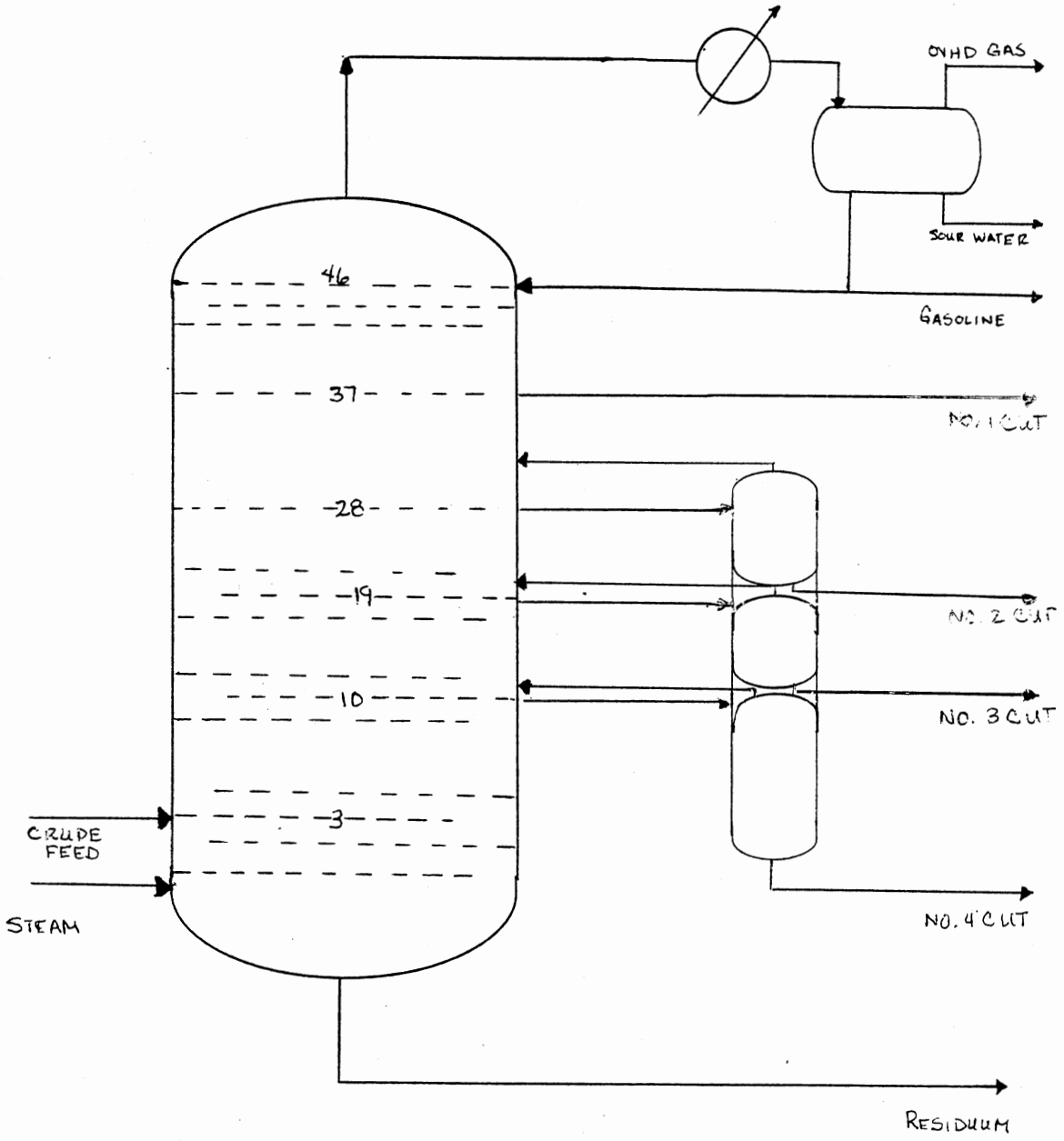
IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	15	16	17	18	19	20	21
46	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
45	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
44	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
43	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
42	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
41	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
40	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
39	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
38	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
37	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
36	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
35	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
34	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
33	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
32	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
31	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
30	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
29	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
28	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
27	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
25	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
23	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
21	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
19	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
17	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
16	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
15	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
13	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
11	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
10	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
9	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
8	0.0041	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
7	0.0144	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
6	0.0423	0.0008	0.0000	0.0000	0.0000	0.0000	0.0000
5	0.1064	0.0076	0.0003	0.0000	0.0000	0.0000	0.0000
4	0.2149	0.0541	0.0097	0.0020	0.0001	0.0000	0.0000
3	0.1781	0.1415	0.0886	0.0787	0.0705	0.0684	0.0609
2	0.1794	0.1427	0.0894	0.0795	0.0712	0.0691	0.0615
1	0.1808	0.1445	0.0906	0.0806	0.0722	0.0700	0.0624

APPENDIX C

IRANIAN CRUDE COLUMN DESIGN



IRANIAN CRUDE COLUMN DESIGN

FEEDS

FEED #	1
TRAY	3
COMP	
C2H6	0.0303
C3H8	0.4670
I-C4H10	0.1980
N-C4H10	0.8620
I-C5H12	0.6810
N-C5H12	0.9240
N-C6H14	1.6700
FR1	0.7220
FR2	2.7200
FR3	2.5220
FR4	2.8620
FR5	2.4210
FR6	2.2000
FR7	1.8980
FR8	1.6990
FR9	1.3570
FR10	0.8510
FR11	0.7570
FR12	0.6780
FR13	0.6580
FR14	0.5860
H2O	0.0000
RATE, LBMOLS	26.7633
TEMP, F	584.9500
ENTHALPY, BTU	926972.06
L/F	0.2880
MOL WT.	204.2151
DENSITY,LB/FT3	49.5434

STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS 5.8100
TEMPERATURE, F : 700.0000

IRANIAN CRUDE

PRODUCTS

	BOTTOMS	DISTILLATE
COMP		
C2H6	0.0000	0.0286
C3H8	0.0001	0.3863
I-C4H10	0.0001	0.1308
N-C4H10	0.0006	0.5084
I-C5H12	0.0013	0.2559
N-C5H12	0.0020	0.2954
N-C6H14	0.0084	0.2278
FR1	0.0067	0.0742
FR2	0.0384	0.0920
FR3	0.0675	0.0000
FR4	0.1578	0.0000
FR5	0.2587	0.0000
FR6	0.5689	0.0000
FR7	1.5972	0.0000
FR8	1.6985	0.0000
FR9	1.3570	0.0000
FR10	0.8510	0.0000
FR11	0.7570	0.0000
FR12	0.6780	0.0000
FR13	0.6580	0.0000
FR14	0.5860	0.0000
RATE, LBMOLS	9.2931	2.0031
TEMP, F	652.7953	132.7272
L/F	1.0000	0.0000
MOL WT.	361.7840	65.2490
DENSITY,LB/FT3	56.6649	

REBOILER DUTY, BTU : 0.0000
CONDENSER DUTY, BTU : -561666.6875

SOUR WATER PRODUCED, LBMOLS: 7.6965
OVERHEAD STEAM, LBMOLS: 0.1735

IRANIAN CRUDE

SIDE DRAWS

DRAW #	1	2
TRAY	46	37
COMP #		
C2H6	0.0017	0.0000
C3H8	0.0788	0.0015
I-C4H10	0.0654	0.0013
N-C4H10	0.3436	0.0071
I-C5H12	0.4084	0.0110
N-C5H12	0.6019	0.0176
N-C6H14	1.3278	0.0771
FR1	0.5685	0.0533
FR2	1.5058	0.9656
FR3	0.0012	2.1991
FR4	0.0000	1.0012
FR5	0.0000	0.0000
FR6	0.0000	0.0000
FR7	0.0000	0.0000
FR8	0.0000	0.0000
FR9	0.0000	0.0000
FR10	0.0000	0.0000
FR11	0.0000	0.0000
FR12	0.0000	0.0000
FR13	0.0000	0.0000
FR14	0.0000	0.0000
RATE, LBMOLS	4.9030	4.3350
TEMP, F	132.7272	295.8848
L/F	1.0000	1.0000
MOL WT.	86.1758	117.5912
DENSITY,LB/FT3	42.1779	47.2688

IRANIAN CRUDE

SIDESTREAM STRIPPERS

STRIPPER #	1	2	3
TRAY	10	19	28
NO. TRAYS	3	3	3
COMP #			
C2H6	0.0000	0.0000	0.0000
C3H8	0.0001	0.0000	0.0000
I-C4H10	0.0002	0.0001	0.0001
N-C4H10	0.0012	0.0004	0.0006
I-C5H12	0.0022	0.0009	0.0013
N-C5H12	0.0035	0.0014	0.0022
N-C6H14	0.0129	0.0062	0.0100
FR1	0.0084	0.0042	0.0066
FR2	0.0483	0.0264	0.0440
FR3	0.0890	0.0555	0.1109
FR4	0.2493	0.2476	1.2167
FR5	1.1040	1.0631	0.0040
FR6	1.6307	0.0002	0.0000
FR7	0.3007	0.0000	0.0000
FR8	0.0005	0.0000	0.0000
FR9	0.0000	0.0000	0.0000
FR10	0.0000	0.0000	0.0000
FR11	0.0000	0.0000	0.0000
FR12	0.0000	0.0000	0.0000
FR13	0.0000	0.0000	0.0000
FR14	0.0000	0.0000	0.0000
RATE, LBMOLS	3.4511	1.4060	1.3965
TEMP, F	472.7130	427.5630	373.7302
L/F	1.0000	1.0000	1.0000
MOL WT.	181.8458	160.5384	136.2065
DENSITY,LB/FT3	51.1220	49.6749	48.5427
STEAM, LBMOLS	1.2100	0.4500	0.4000
TEMP, F	500.0000	475.0000	460.0000
PUMPBACK	0.0000	0.0000	0.0000
TRAY	0	0	0

IRANIAN CRUDE

TEMPERATURE, LIQUID AND VAPOR PROFILES

STREAMS LEAVING TRAY

TRAY NO	TEMP DEGF	PRES mmHg	LIQUID LB-MOLS	MOL WT	DENS LB/FT3	VAPOR LB-MOLS	MOL WT
46	132.73	1292.50	27.30	86.18	42.18	2.00	65.25
45	203.65	1298.24	26.82	96.00	44.34	34.21	84.95
44	226.18	1303.99	26.10	99.67	45.12	33.73	92.74
43	234.84	1309.73	25.73	101.30	45.45	33.01	95.58
42	239.35	1315.48	25.51	102.23	45.61	32.63	96.81
41	243.17	1321.22	25.33	103.11	45.74	32.42	97.52
40	248.66	1326.97	25.04	104.49	45.93	32.23	98.18
39	258.13	1332.71	24.53	106.93	46.24	31.95	99.22
38	273.57	1338.46	23.70	111.08	46.69	31.44	101.04
37	295.88	1344.20	18.19	117.59	47.27	30.61	104.10
36	323.03	1349.94	17.11	125.49	47.85	29.43	108.80
35	343.11	1355.69	16.41	130.85	48.20	28.36	113.23
34	354.27	1361.43	16.05	133.61	48.36	27.65	116.10
33	359.55	1367.18	15.89	134.83	48.44	27.29	117.52
32	361.97	1372.92	15.82	135.34	48.47	27.13	118.15
31	363.14	1378.67	15.79	135.54	48.48	27.06	118.40
30	363.82	1384.41	15.78	135.63	48.48	27.03	118.50
29	364.30	1390.16	15.77	135.68	48.48	27.02	118.55
28	364.77	1395.90	14.33	135.75	48.49	26.97	118.57
27	365.36	1401.64	14.32	135.90	48.50	26.96	118.60
26	366.32	1407.39	14.30	136.29	48.51	26.95	118.68
25	368.17	1413.13	14.25	137.21	48.56	26.93	118.87
24	371.93	1418.88	14.14	139.21	48.65	26.88	119.33
23	379.06	1424.62	13.92	142.93	48.82	26.77	120.31
22	390.09	1430.37	13.57	148.22	49.07	26.55	122.10
21	402.70	1436.11	13.20	153.58	49.32	26.20	124.57
20	413.13	1441.86	12.92	157.52	49.51	25.83	126.97
19	419.83	1447.60	11.32	159.81	49.61	25.51	128.67
18	423.55	1453.34	11.24	160.99	49.67	25.34	129.63
17	425.36	1459.09	11.21	161.51	49.69	25.26	130.06
16	426.39	1464.83	11.19	161.77	49.71	25.23	130.25
15	427.27	1470.58	11.17	162.03	49.72	25.21	130.34
14	428.56	1476.32	11.14	162.54	49.76	25.19	130.44
13	431.22	1482.07	11.07	163.80	49.86	25.16	130.63
12	437.12	1487.81	10.90	166.70	50.08	25.09	131.09
11	448.93	1493.56	10.56	172.37	50.50	24.92	132.13
10	468.19	1499.30	6.55	181.34	51.08	24.54	134.10
9	493.74	1505.04	6.12	193.39	51.70	24.02	137.01
8	513.14	1510.79	5.83	202.67	52.08	23.59	139.34
7	526.44	1516.53	5.65	209.01	52.31	23.30	140.99
6	535.34	1522.28	5.51	213.57	52.48	23.12	142.04
5	543.26	1528.02	5.33	219.01	52.70	22.98	142.74
4	556.50	1533.77	4.67	232.93	53.18	22.81	143.47
3	613.57	1539.51	9.51	357.26	56.49	22.14	144.14
2	635.75	1545.26	9.42	359.20	56.57	0.21	159.97
1	652.80	1551.00	9.29	361.78	56.66	0.13	171.76

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
46	0.0003	0.0161	0.0133	0.0701	0.0833	0.1228	0.2708
45	0.0000	0.0012	0.0016	0.0103	0.0230	0.0405	0.2077
44	0.0000	0.0005	0.0006	0.0034	0.0079	0.0146	0.1226
43	0.0000	0.0005	0.0005	0.0027	0.0050	0.0087	0.0748
42	0.0000	0.0005	0.0005	0.0025	0.0044	0.0074	0.0524
41	0.0000	0.0005	0.0004	0.0025	0.0042	0.0069	0.0419
40	0.0000	0.0005	0.0004	0.0023	0.0039	0.0065	0.0357
39	0.0000	0.0004	0.0004	0.0022	0.0036	0.0059	0.0301
38	0.0000	0.0004	0.0004	0.0019	0.0031	0.0050	0.0240
37	0.0000	0.0004	0.0003	0.0016	0.0025	0.0041	0.0178
36	0.0000	0.0003	0.0003	0.0014	0.0021	0.0032	0.0129
35	0.0000	0.0003	0.0002	0.0012	0.0018	0.0028	0.0105
34	0.0000	0.0003	0.0002	0.0012	0.0017	0.0026	0.0094
33	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0091
32	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
31	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0089
30	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0088
29	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0088
28	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0088
27	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0088
26	0.0000	0.0003	0.0002	0.0011	0.0016	0.0025	0.0088
25	0.0000	0.0003	0.0002	0.0011	0.0016	0.0024	0.0087
24	0.0000	0.0003	0.0002	0.0011	0.0016	0.0024	0.0085
23	0.0000	0.0003	0.0002	0.0011	0.0015	0.0023	0.0080
22	0.0000	0.0002	0.0002	0.0010	0.0014	0.0021	0.0073
21	0.0000	0.0002	0.0002	0.0009	0.0013	0.0020	0.0067
20	0.0000	0.0002	0.0002	0.0009	0.0012	0.0019	0.0062
19	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0060
18	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
17	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
16	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
15	0.0000	0.0002	0.0002	0.0009	0.0012	0.0018	0.0058
14	0.0000	0.0002	0.0002	0.0009	0.0012	0.0017	0.0057
13	0.0000	0.0002	0.0002	0.0008	0.0011	0.0017	0.0056
12	0.0000	0.0002	0.0002	0.0008	0.0011	0.0017	0.0054
11	0.0000	0.0002	0.0002	0.0008	0.0010	0.0016	0.0050
10	0.0000	0.0002	0.0001	0.0007	0.0009	0.0014	0.0045
9	0.0000	0.0002	0.0001	0.0006	0.0008	0.0012	0.0038
8	0.0000	0.0002	0.0001	0.0006	0.0008	0.0011	0.0035
7	0.0000	0.0002	0.0001	0.0006	0.0007	0.0011	0.0032
6	0.0000	0.0002	0.0001	0.0006	0.0007	0.0010	0.0031
5	0.0000	0.0002	0.0001	0.0005	0.0007	0.0010	0.0030
4	0.0000	0.0001	0.0001	0.0005	0.0007	0.0009	0.0028
3	0.0000	0.0001	0.0001	0.0004	0.0005	0.0007	0.0021
2	0.0000	0.0000	0.0000	0.0002	0.0003	0.0004	0.0014
1	0.0000	0.0000	0.0000	0.0001	0.0001	0.0002	0.0009

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	8	9	10	11	12	13	14
46	0.1159	0.3071	0.0002	0.0000	0.0000	0.0000	0.0000
45	0.1197	0.5946	0.0013	0.0000	0.0000	0.0000	0.0000
44	0.0905	0.7555	0.0044	0.0000	0.0000	0.0000	0.0000
43	0.0646	0.8312	0.0121	0.0000	0.0000	0.0000	0.0000
42	0.0475	0.8540	0.0307	0.0000	0.0000	0.0000	0.0000
41	0.0369	0.8329	0.0734	0.0004	0.0000	0.0000	0.0000
40	0.0296	0.7571	0.1613	0.0026	0.0000	0.0000	0.0000
39	0.0234	0.6120	0.3064	0.0156	0.0000	0.0000	0.0000
38	0.0175	0.4133	0.4621	0.0722	0.0000	0.0000	0.0000
37	0.0123	0.2227	0.5073	0.2310	0.0000	0.0000	0.0000
36	0.0085	0.0993	0.3873	0.4848	0.0000	0.0000	0.0000
35	0.0066	0.0539	0.2439	0.6788	0.0000	0.0000	0.0000
34	0.0059	0.0401	0.1549	0.7838	0.0000	0.0000	0.0000
33	0.0056	0.0360	0.1120	0.8315	0.0000	0.0000	0.0000
32	0.0055	0.0347	0.0935	0.8516	0.0000	0.0000	0.0000
31	0.0055	0.0343	0.0859	0.8596	0.0001	0.0000	0.0000
30	0.0055	0.0341	0.0828	0.8626	0.0004	0.0000	0.0000
29	0.0055	0.0341	0.0816	0.8632	0.0010	0.0000	0.0000
28	0.0055	0.0341	0.0811	0.8619	0.0028	0.0000	0.0000
27	0.0055	0.0341	0.0807	0.8575	0.0077	0.0000	0.0000
26	0.0055	0.0339	0.0800	0.8464	0.0197	0.0000	0.0000
25	0.0054	0.0334	0.0785	0.8202	0.0481	0.0000	0.0000
24	0.0052	0.0323	0.0752	0.7633	0.1099	0.0000	0.0000
23	0.0050	0.0303	0.0690	0.6578	0.2247	0.0000	0.0000
22	0.0045	0.0274	0.0602	0.5077	0.3878	0.0000	0.0000
21	0.0041	0.0245	0.0514	0.3555	0.5532	0.0000	0.0000
20	0.0038	0.0224	0.0451	0.2431	0.6749	0.0000	0.0000
19	0.0037	0.0212	0.0415	0.1772	0.7460	0.0001	0.0000
18	0.0036	0.0206	0.0397	0.1434	0.7822	0.0004	0.0000
17	0.0036	0.0204	0.0389	0.1290	0.7968	0.0013	0.0000
16	0.0036	0.0203	0.0386	0.1230	0.8009	0.0036	0.0000
15	0.0036	0.0203	0.0384	0.1202	0.7978	0.0098	0.0000
14	0.0035	0.0201	0.0381	0.1180	0.7840	0.0264	0.0000
13	0.0035	0.0198	0.0373	0.1143	0.7470	0.0681	0.0004
12	0.0034	0.0190	0.0353	0.1065	0.6628	0.1608	0.0029
11	0.0031	0.0174	0.0317	0.0922	0.5110	0.3172	0.0186
10	0.0028	0.0151	0.0267	0.0730	0.3193	0.4689	0.0863
9	0.0024	0.0127	0.0216	0.0542	0.1600	0.4782	0.2629
8	0.0021	0.0113	0.0186	0.0441	0.0972	0.3693	0.4456
7	0.0020	0.0104	0.0169	0.0388	0.0744	0.2643	0.5694
6	0.0019	0.0100	0.0159	0.0359	0.0649	0.1981	0.6176
5	0.0018	0.0095	0.0151	0.0336	0.0589	0.1588	0.5899
4	0.0017	0.0089	0.0138	0.0302	0.0512	0.1253	0.4660
3	0.0013	0.0066	0.0097	0.0198	0.0301	0.0626	0.1712
2	0.0010	0.0054	0.0086	0.0187	0.0293	0.0623	0.1719
1	0.0007	0.0041	0.0073	0.0170	0.0278	0.0612	0.1719

IRANIAN CRUDE

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	15	16	17	18	19	20	21
46	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
45	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
44	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
43	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
42	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
41	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
40	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
39	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
38	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
37	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
36	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
35	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
34	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
33	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
32	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
31	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
30	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
29	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
28	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
27	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
25	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
23	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
21	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
19	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
17	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
16	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
15	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
13	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
11	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
10	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
9	0.0012	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
8	0.0054	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
7	0.0178	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
6	0.0491	0.0010	0.0000	0.0000	0.0000	0.0000	0.0000
5	0.1180	0.0085	0.0004	0.0000	0.0000	0.0000	0.0000
4	0.2276	0.0575	0.0103	0.0022	0.0001	0.0000	0.0000
3	0.1801	0.1431	0.0896	0.0796	0.0713	0.0692	0.0616
2	0.1814	0.1443	0.0904	0.0804	0.0720	0.0698	0.0622
1	0.1828	0.1460	0.0916	0.0815	0.0730	0.0708	0.0631

APPENDIX D

OPERATING DATA COMPARISON

FEEDS

FEED #	1	2
TRAY	1	3
COMP		
C2H6	0.0220	0.0000
C3H8	0.1400	0.0000
I-C4H10	0.1180	0.0000
N-C4H10	0.1690	0.0000
I-C5H12	0.2590	0.0000
N-C5H12	0.2370	0.0000
CYC-C5	0.0100	0.0000
FR1	0.7140	0.0000
FR2	0.9480	0.0000
FR3	1.6830	0.0000
FR4	1.5860	0.1322
FR5	1.4770	0.4930
FR6	1.3640	0.4550
FR7	1.2580	0.4200
FR8	1.7400	0.0000
FR9	1.5920	0.0000
FR10	1.4500	0.0000
FR11	1.3370	0.0000
FR12	1.2300	0.0000
FR13	0.7680	0.0000
FR14	0.7330	0.0000
FR15	0.6930	0.0000
FR16	0.9600	0.0000
FR17	0.8670	0.0000
FR18	0.7800	0.0000
FR19	0.6940	0.0000
FR20	0.6230	0.0000
FR21	0.2800	0.0000
H20	0.0000	0.0000
RATE, LBMOLS	23.7320	1.5002
TEMP, F	479.0000	319.0000
ENTHALPY, BTU	1866410.38	49977.88
L/F	0.6945	1.0000
MOL WT.	220.6098	139.8312
DENSITY,LB/FT3	49.9172	48.2695

STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS 0.0000
TEMPERATURE, F : 0.0000

LOUSIANNA CRUDE PREFLASH TOWER

PRODUCTS

	BOTTOMS	DISTILLATE
COMP		
C2H6	0.0003	0.0217
C3H8	0.0037	0.1363
I-C4H10	0.0054	0.1126
N-C4H10	0.0085	0.1605
I-C5H12	0.0208	0.2382
N-C5H12	0.0206	0.2164
CYC-C5	0.0011	0.0089
FR1	0.0890	0.6250
FR2	0.1467	0.8013
FR3	0.3557	1.3273
FR4	0.4708	1.2474
FR5	0.5825	1.3875
FR6	0.6872	1.1318
FR7	0.7558	0.9222
FR8	1.2432	0.4968
FR9	1.3065	0.2855
FR10	1.3134	0.1366
FR11	1.2830	0.0540
FR12	1.2169	0.0131
FR13	0.7659	0.0021
FR14	0.7324	0.0006
FR15	0.6929	0.0001
FR16	0.9600	0.0000
FR17	0.8670	0.0000
FR18	0.7800	0.0000
FR19	0.6940	0.0000
FR20	0.6230	0.0000
FR21	0.2800	0.0000
H2O	0.0000	0.0000
RATE, LBMOLS	15.9062	9.3260
TEMP, F	487.7769	428.9308
L/F	1.0000	0.0000
MOL WT.	271.6258	120.6033
DENSITY,LB/FT3	52.5387	

REBOILER DUTY, BTU : 0.0000
CONDENSER DUTY, BTU : 0.5000

SOUR WATER PRODUCED, LBMOLS: 0.0000
OVERHEAD STEAM, LBMOLS: 0.0000

LOUSIANNA CRUDE PREFLASH TOWER

TEMPERATURE, LIQUID AND VAPOR PROFILES

STREAMS LEAVING TRAY

TRAY NO	TEMP DEGF	PRES mmHg	LIQUID LB-MOLS	MOL WT	DENS LB/FT3	VAPOR LB-MOLS	MOL WT
3	428.93	2207.60	0.80	178.13	49.88	9.33	120.60
2	465.92	2207.60	0.29	205.72	50.94	8.62	122.58
1	487.78	2207.60	15.91	271.63	52.54	8.12	120.09

LOUSIANNA CRUDE PREFLASH TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
3	0.0000	0.0003	0.0004	0.0006	0.0016	0.0016	0.0001
2	0.0000	0.0002	0.0004	0.0006	0.0014	0.0014	0.0001
1	0.0000	0.0002	0.0003	0.0005	0.0013	0.0013	0.0001

LOUSIANNA CRUDE PREFLASH TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	8	9	10	11	12	13	14
3	0.0069	0.0117	0.0297	0.0462	0.0829	0.1098	0.1414
2	0.0060	0.0100	0.0248	0.0338	0.0441	0.0552	0.0667
1	0.0056	0.0092	0.0224	0.0296	0.0366	0.0432	0.0475

LOUSIANNA CRUDE PREFLASH TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	15	16	17	18	19	20	21
3	0.1305	0.1373	0.1279	0.0977	0.0497	0.0144	0.0063
2	0.1078	0.1249	0.1408	0.1441	0.1155	0.0518	0.0341
1	0.0782	.0.0821	0.0826	0.0807	0.0765	0.0482	0.0460

LOUSIANNA CRUDE PREFLASH TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	22	23	24	25	26	27	28
3	0.0022	0.0006	0.0001	0.0000	0.0000	0.0000	0.0000
2	0.0197	0.0125	0.0035	0.0007	0.0001	0.0000	0.0000
1	0.0436	0.0604	0.0545	0.0490	0.0436	0.0392	0.0176

LOUSIANNA CRUDE PREFLASH TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER
NO	29
3	0.0000
2	0.0000
1	0.0000

FEEDS

FEED #	1	2	3
TRAY	3	15	14
COMP			
C2H6	0.0003	0.0217	0.0000
C3H8	0.0037	0.1363	0.0000
I-C4H10	0.0054	0.1126	0.0000
N-C4H10	0.0085	0.1605	0.0000
I-C5H12	0.0208	0.2382	0.0000
N-C5H12	0.0206	0.2164	0.0000
CYC-C5	0.0011	0.0089	0.0000
FR1	0.0890	0.6250	0.0001
FR2	0.1467	0.8013	0.0001
FR3	0.3557	1.3273	0.0002
FR4	0.4708	1.2474	0.0003
FR5	0.5825	1.3875	0.0003
FR6	0.6872	1.1318	0.0004
FR7	0.7558	0.9222	0.0005
FR8	1.2432	0.4968	0.0008
FR9	1.3065	0.2855	0.0009
FR10	1.3134	0.1366	0.0013
FR11	1.2830	0.0540	0.0017
FR12	1.2169	0.0131	0.0011
FR13	0.7659	0.0021	0.0004
FR14	0.7324	0.0006	0.0002
FR15	0.6929	0.0000	0.0001
FR16	0.9600	0.0000	0.0000
FR17	0.8670	0.0000	0.0000
FR18	0.7800	0.0000	0.0000
FR19	0.6940	0.0000	0.0000
FR20	0.6230	0.0000	0.0000
FR21	0.2800	0.0000	0.0000
H2O	0.0000	0.0000	0.1900
RATE, LBMOLS	15.9063	9.3258	0.1984
TEMP, F	775.0000	428.9000	124.0000
ENTHALPY, BTU	2441526.75	470324.50	1723.29
L/F	0.3000	0.0000	1.0000
MOL WT.	271.6249	0.0000	614.1817
DENSITY,LB/FT3	52.5387	45.1809	1465.7516

STEAM RATE TO BOTTOM OF THE COLUMN, LBMOLS 2.5600
TEMPERATURE, F : 800.0000

LOUSIANNA CRUDE ATMOSPHERIC TOWER

PRODUCTS

	BOTTOMS	DISTILLATE
COMP		
C2H6	0.0000	0.0211
C3H8	0.0000	0.1234
I-C4H10	0.0000	0.0902
N-C4H10	0.0000	0.1206
I-C5H12	0.0001	0.1375
N-C5H12	0.0001	0.1132
CYC-C5	0.0000	0.0038
FR1	0.0008	0.2887
FR2	0.0015	0.2756
FR3	0.0046	0.2604
FR4	0.0080	0.0764
FR5	0.0128	0.0072
FR6	0.0196	0.0002
FR7	0.0240	0.0000
FR8	0.0533	0.0000
FR9	0.0776	0.0000
FR10	0.1110	0.0000
FR11	0.1519	0.0000
FR12	0.2075	0.0000
FR13	0.1725	0.0000
FR14	0.2056	0.0000
FR15	0.2481	0.0000
FR16	0.4765	0.0000
FR17	0.6302	0.0000
FR18	0.7333	0.0000
FR19	0.6926	0.0000
FR20	0.6230	0.0000
FR21	0.2800	0.0000
RATE, LBMOLS	4.7345	1.5104
TEMP, F	779.8489	177.7381
L/F	1.0000	0.0000
MOL WT.	412.6764	78.8356
DENSITY,LB/FT3	56.3013	

REBOILER DUTY, BTU : 0.0000
CONDENSER DUTY, BTU : -686220.3125

SOUR WATER PRODUCED, LBMOLS: 3.5854
OVERHEAD STEAM, LBMOLS: 0.2746

LOUSIANNA CRUDE ATMOSPHERIC TOWER

SIDE DRAWS

DRAW #	1	2
TRAY	27	20
COMP #		
C2H6	0.0009	0.0000
C3H8	0.0160	0.0001
I-C4H10	0.0269	0.0001
N-C4H10	0.0467	0.0002
I-C5H12	0.1168	0.0006
N-C5H12	0.1188	0.0006
CYC-C5	0.0059	0.0000
FR1	0.4023	0.0028
FR2	0.6266	0.0051
FR3	1.2345	0.0150
FR4	0.8596	0.0332
FR5	0.1829	0.1130
FR6	0.0117	0.3034
FR7	0.0003	0.5055
FR8	0.0000	0.3703
FR9	0.0000	0.1229
FR10	0.0000	0.0249
FR11	0.0000	0.0022
FR12	0.0000	0.0000
FR13	0.0000	0.0000
FR14	0.0000	0.0000
FR15	0.0000	0.0000
FR16	0.0000	0.0000
FR17	0.0000	0.0000
FR18	0.0000	0.0000
FR19	0.0000	0.0000
FR20	0.0000	0.0000
FR21	0.0000	0.0000
RATE, LBMOLS	3.6500	1.5000
TEMP, F	177.7381	400.6325
L/F	1.0000	1.0000
MOL WT.	97.8201	158.4073
DENSITY,LB/FT3	42.6359	49.4081

PUMPAROUND 1 FROM TRAY 11 TO 12
 RATE, LBMOLS: 7.4500 COOLING, BTU: 155089.0000

LOUSIANNA CRUDE ATMOSPHERIC TOWER

SIDESTREAM STRIPPERS

STRIPPER #	1	2	3
TRAY	23	18	13
NO. TRAYS	3	2	2
COMP #			
C2H6	0.0000	0.0000	0.0000
C3H8	0.0003	0.0002	0.0000
I-C4H10	0.0005	0.0003	0.0000
N-C4H10	0.0008	0.0005	0.0000
I-C5H12	0.0025	0.0014	0.0001
N-C5H12	0.0026	0.0015	0.0001
CYC-C5	0.0002	0.0001	0.0000
FR1	0.0120	0.0067	0.0006
FR2	0.0260	0.0119	0.0012
FR3	0.1311	0.0327	0.0040
FR4	0.6730	0.0576	0.0079
FR5	1.5166	0.1216	0.0143
FR6	1.1898	0.2689	0.0247
FR7	0.4677	0.6424	0.0362
FR8	0.0631	1.1530	0.0933
FR9	0.0030	1.2010	0.1750
FR10	0.0001	0.8859	0.4060
FR11	0.0000	0.2824	0.8640
FR12	0.0000	0.0187	0.9426
FR13	0.0000	0.0009	0.5313
FR14	0.0000	0.0001	0.4256
FR15	0.0000	0.0000	0.2429
FR16	0.0000	0.0000	0.0243
FR17	0.0000	0.0000	0.0001
FR18	0.0000	0.0000	0.0000
FR19	0.0000	0.0000	0.0000
FR20	0.0000	0.0000	0.0000
FR21	0.0000	0.0000	0.0000
RATE, LBMOLS	4.0890	4.6878	3.7943
TEMP, F	332.4135	452.5168	597.5359
L/F	1.0000	1.0000	1.0000
MOL WT.	133.1460	183.5881	252.1744
DENSITY,LB/FT3	47.5705	50.5559	52.8810
STEAM, LBMOLS	0.2000	0.6200	0.1900
TEMP, F	450.0000	550.0000	650.0000
PUMPBACK	0.0000	0.0000	0.0000
TRAY	0	0	0

LOUSIANNA CRUDE ATMOSPHERIC TOWER

SIDESTREAM STRIPPERS

STRIPPER #	4
TRAY	6
NO. TRAYS	2
COMP #	
C2H6	0.0000
C3H8	0.0000
I-C4H10	0.0000
N-C4H10	0.0000
I-C5H12	0.0000
N-C5H12	0.0000
CYC-C5	0.0000
FR1	0.0000
FR2	0.0000
FR3	0.0001
FR4	0.0002
FR5	0.0005
FR6	0.0012
FR7	0.0017
FR8	0.0056
FR9	0.0114
FR10	0.0216
FR11	0.0367
FR12	0.0619
FR13	0.0630
FR14	0.0993
FR15	0.2071
FR16	0.4573
FR17	0.2369
FR18	0.0470
FR19	0.0014
FR20	0.0000
FR21	0.0000
TEMP, F	720.8229
RATE, LBMOLS	1.2528
L/F	1.0000
MOL WT.	331.5270
DENSITY,LB/FT3	54.7435
STEAM, LBMOLS	0.2900
TEMP, F	750.0000
PUMPBACK	0.0000
TRAY	0

LOUSIANNA CRUDE ATMOSPHERIC TOWER

TEMPERATURE, LIQUID AND VAPOR PROFILES

STREAMS LEAVING TRAY

TRAY NO	TEMP DEGF	PRES mmHg	LIQUID LB-MOLS	MOL WT	DENS LB/FT3	VAPOR LB-MOLS	MOL WT
27	177.74	1448.00	35.00	97.82	42.64	1.51	78.84
26	247.66	1448.00	35.05	109.91	44.50	40.16	97.11
25	282.44	1448.00	33.88	118.28	45.78	40.21	107.65
24	308.36	1448.00	32.40	125.53	46.75	39.04	114.84
23	332.18	1448.00	26.61	133.08	47.56	37.55	120.96
22	356.76	1448.00	25.10	141.44	48.29	35.86	127.21
21	379.17	1448.00	23.67	149.61	48.88	34.35	133.06
20	400.63	1448.00	20.45	158.41	49.41	32.92	138.57
19	424.26	1448.00	18.37	169.51	49.96	31.20	144.15
18	451.69	1448.00	11.51	183.47	50.55	29.11	150.13
17	482.62	1448.00	10.22	199.13	51.14	26.95	156.97
16	504.64	1448.00	9.30	210.59	51.53	25.65	161.87
15	520.84	1448.00	8.66	220.48	51.85	24.73	164.78
14	570.45	1448.00	8.59	236.04	52.41	14.77	196.52
13	597.50	1448.00	3.80	252.01	52.87	14.70	205.52
12	624.81	1448.00	15.08	269.89	53.35	13.70	212.20
11	647.41	1448.00	7.59	279.03	53.61	17.54	224.82
10	667.44	1448.00	6.58	290.30	53.88	17.49	232.57
9	686.71	1606.00	6.00	297.83	54.04	16.48	234.20
8	701.49	1729.00	5.31	304.88	54.18	15.90	234.96
7	715.42	1861.00	4.51	312.71	54.33	15.21	234.56
6	730.49	2002.70	2.15	324.04	54.54	14.30	232.83
5	748.73	2104.70	1.28	344.69	54.93	13.32	229.96
4	764.52	2206.70	0.46	368.59	55.38	12.45	225.51
3	779.61	2404.40	4.74	412.35	56.29	11.63	218.06
2	779.55	2404.40	4.74	412.51	56.30	0.01	217.13
1	779.85	2404.40	4.73	412.68	56.30	0.00	216.83

LOUSIANNA CRUDE ATMOSPHERIC TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	1	2	3	4	5	6	7
27	0.0002	0.0044	0.0074	0.0128	0.0320	0.0326	0.0016
26	0.0000	0.0003	0.0007	0.0014	0.0062	0.0074	0.0005
25	0.0000	0.0001	0.0002	0.0004	0.0015	0.0018	0.0001
24	0.0000	0.0001	0.0001	0.0002	0.0008	0.0009	0.0001
23	0.0000	0.0001	0.0001	0.0002	0.0006	0.0007	0.0000
22	0.0000	0.0001	0.0001	0.0002	0.0005	0.0005	0.0000
21	0.0000	0.0001	0.0001	0.0002	0.0005	0.0005	0.0000
20	0.0000	0.0001	0.0001	0.0001	0.0004	0.0004	0.0000
19	0.0000	0.0001	0.0001	0.0001	0.0004	0.0004	0.0000
18	0.0000	0.0001	0.0001	0.0001	0.0003	0.0003	0.0000
17	0.0000	0.0000	0.0001	0.0001	0.0003	0.0003	0.0000
16	0.0000	0.0000	0.0001	0.0001	0.0003	0.0003	0.0000
15	0.0000	0.0000	0.0001	0.0001	0.0003	0.0003	0.0000
14	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
13	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
12	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
11	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
10	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
9	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
7	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
5	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
3	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
1	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000

LOUSIANNA CRUDE ATMOSPHERIC TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	8	9	10	11	12	13	14
27	0.1102	0.1717	0.3382	0.2355	0.0501	0.0032	0.0001
26	0.0337	0.0775	0.2789	0.4009	0.1696	0.0218	0.0012
25	0.0098	0.0277	0.1539	0.4048	0.3148	0.0761	0.0084
24	0.0043	0.0112	0.0707	0.2942	0.4004	0.1766	0.0375
23	0.0030	0.0064	0.0324	0.1654	0.3710	0.2903	0.1137
22	0.0024	0.0047	0.0176	0.0761	0.2541	0.3420	0.2369
21	0.0021	0.0039	0.0124	0.0370	0.1437	0.2966	0.3314
20	0.0019	0.0034	0.0100	0.0221	0.0753	0.2022	0.3370
19	0.0017	0.0030	0.0084	0.0159	0.0412	0.1128	0.2502
18	0.0015	0.0026	0.0071	0.0124	0.0261	0.0577	0.1375
17	0.0013	0.0023	0.0060	0.0100	0.0187	0.0315	0.0608
16	0.0012	0.0021	0.0054	0.0087	0.0155	0.0233	0.0356
15	0.0012	0.0020	0.0050	0.0079	0.0139	0.0199	0.0273
14	0.0002	0.0004	0.0014	0.0028	0.0052	0.0091	0.0139
13	0.0002	0.0003	0.0011	0.0022	0.0039	0.0066	0.0097
12	0.0002	0.0003	0.0010	0.0020	0.0034	0.0056	0.0079
11	0.0001	0.0002	0.0007	0.0014	0.0024	0.0039	0.0055
10	0.0001	0.0002	0.0007	0.0012	0.0021	0.0034	0.0045
9	0.0001	0.0002	0.0007	0.0013	0.0022	0.0035	0.0046
8	0.0001	0.0002	0.0007	0.0013	0.0022	0.0035	0.0046
7	0.0001	0.0003	0.0008	0.0014	0.0023	0.0036	0.0047
6	0.0001	0.0003	0.0008	0.0015	0.0024	0.0038	0.0048
5	0.0002	0.0003	0.0009	0.0015	0.0025	0.0039	0.0048
4	0.0002	0.0003	0.0009	0.0016	0.0026	0.0040	0.0049
3	0.0002	0.0003	0.0010	0.0018	0.0028	0.0042	0.0052
2	0.0002	0.0003	0.0010	0.0017	0.0027	0.0042	0.0051
1	0.0002	0.0003	0.0010	0.0017	0.0027	0.0041	0.0051

LOUSIANNA CRUDE ATMOSPHERIC TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	15	16	17	18	19	20	21
27	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
25	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
24	0.0026	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
23	0.0153	0.0007	0.0000	0.0000	0.0000	0.0000	0.0000
22	0.0588	0.0055	0.0003	0.0000	0.0000	0.0000	0.0000
21	0.1434	0.0255	0.0026	0.0001	0.0000	0.0000	0.0000
20	0.2469	0.0819	0.0166	0.0015	0.0000	0.0000	0.0000
19	0.3001	0.1814	0.0718	0.0122	0.0004	0.0000	0.0000
18	0.2461	0.2559	0.1884	0.0598	0.0039	0.0002	0.0000
17	0.1370	0.2296	0.3032	0.1741	0.0225	0.0019	0.0003
16	0.0761	0.1589	0.3107	0.2844	0.0656	0.0092	0.0022
15	0.0510	0.1044	0.2492	0.3364	0.1358	0.0310	0.0115
14	0.0348	0.0711	0.1804	0.3239	0.2179	0.0788	0.0434
13	0.0248	0.0464	0.1074	0.2279	0.2479	0.1397	0.1120
12	0.0193	0.0325	0.0596	0.1179	0.1935	0.1668	0.1917
11	0.0135	0.0229	0.0417	0.0835	0.1640	0.1682	0.2214
10	0.0108	0.0175	0.0294	0.0526	0.1097	0.1351	0.2205
9	0.0107	0.0168	0.0267	0.0429	0.0805	0.1009	0.1850
8	0.0106	0.0165	0.0256	0.0391	0.0662	0.0769	0.1432
7	0.0108	0.0165	0.0251	0.0373	0.0590	0.0618	0.1069
6	0.0109	0.0165	0.0247	0.0359	0.0541	0.0519	0.0790
5	0.0108	0.0161	0.0236	0.0335	0.0483	0.0430	0.0572
4	0.0109	0.0159	0.0230	0.0320	0.0447	0.0382	0.0472
3	0.0114	0.0166	0.0236	0.0322	0.0439	0.0365	0.0434
2	0.0113	0.0165	0.0235	0.0321	0.0439	0.0364	0.0434
1	0.0112	0.0164	0.0234	0.0321	0.0438	0.0364	0.0434

LOUSIANNA CRUDE ATMOSPHERIC TOWER

LIQUID COMPOSITION PROFILE

TRAY	COMPONENT NUMBER						
NO	22	23	24	25	26	27	28
27	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
26	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
25	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
24	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
23	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
22	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
21	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
20	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
19	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
18	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
17	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
16	0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
15	0.0026	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
14	0.0158	0.0008	0.0000	0.0000	0.0000	0.0000	0.0000
13	0.0636	0.0064	0.0000	0.0000	0.0000	0.0000	0.0000
12	0.1665	0.0316	0.0002	0.0000	0.0000	0.0000	0.0000
11	0.2207	0.0493	0.0004	0.0000	0.0000	0.0000	0.0000
10	0.2976	0.1125	0.0022	0.0000	0.0000	0.0000	0.0000
9	0.3179	0.1972	0.0088	0.0001	0.0000	0.0000	0.0000
8	0.2911	0.2880	0.0292	0.0007	0.0000	0.0000	0.0000
7	0.2321	0.3512	0.0803	0.0057	0.0000	0.0000	0.0000
6	0.1600	0.3439	0.1741	0.0340	0.0010	0.0000	0.0000
5	0.0920	0.2409	0.2649	0.1392	0.0161	0.0005	0.0000
4	0.0623	0.1463	0.2299	0.2354	0.0851	0.0138	0.0009
3	0.0524	0.1006	0.1329	0.1546	0.1460	0.1314	0.0590
2	0.0524	0.1006	0.1330	0.1548	0.1462	0.1315	0.0591
1	0.0524	0.1006	0.1331	0.1549	0.1463	0.1316	0.0591

VITA

Susan Leigh Parker

Candidate for the Degree of

Master of Science

Thesis: COMPUTER SIMULATION OF A MULTICOMPONENT, MULTISTAGE CRUDE DISTILLATION PROCESS

Major Field: Chemical Engineering

Biographical:

Personal Data: Born in Ponca City, Oklahoma, December 4, 1964, the daughter of Clarence and Patsy Parker.

Education: Graduated from Ponca City Senior High School, Ponca City, Oklahoma, in May 1983; received Bachelor of Science Degree in Chemical Engineering from Oklahoma State University, Stillwater, Oklahoma, in May 1987; completed requirements for the Master of Science Degree in Chemical Engineering at Oklahoma State University, Stillwater, Oklahoma, in December 1988.

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