Operation of Integrated Three-Product (Petlyuk) Distillation Columns

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In this paper, we consider operation and control of the Petlyuk design. The degrees of freedom are analyzed especially at steady state. These results together with steady-state solution curves help describe the complex plant dynamics and possible optimization strategies. We also propose control schemes for controlling three and four product compositions. The results indicate that there may be serious problems involved in operating the Petlyuk column, at least for high-purity separations.

1. Introduction

The separation of more than two components by continuous distillation has traditionally been accomplished by arranging columns in series. Several alternative configurations exist, most notably the direct and indirect sequences (where light or heavy components are removed first, respectively).

Almost 50 years ago Wright (1949) proposed a promising design alternative for separating a ternary feed. This design consists of an ordinary column shell with the feed and sidestream product draw divided by a vertical wall through a set of trays. As compared to the direct or indirect sequence, this implementation offers savings in investment (only one shell and two heat exchangers) as well as operating costs. Although several authors have studied the design of such columns, very little work has been done on the operation and control.

This configuration is usually denoted as a *Petlyuk column* after Petlyuk *et al.* (1965) who later studied the scheme theoretically. Many authors have since predicted considerable savings in energy and capital cost with this design, but still few of these integrated columns have been built. One reason is probably that the Petlyuk column, compared to an ordinary distillation column, has many more degrees of freedom in both operation and design. This undoubtably makes the design of both the column and its control system more complex.

A two-column implementation of the Petlyuk design is shown in Figure 1a. It consists of a prefractionator with reflux and boilup from the downstream three-product column, a setup with only one reboiler and one condenser. As proposed by Wright (1949), practical implementation of such a column can be accomplished in a single shell by inserting a vertical wall through the middle section of the column (Figure 1b), thus separating the feed and side product draw. Petlyuk's main reason for this design was to avoid thermodynamic losses from mixing different streams at the feed tray location. We will hereafter denote the product streams D, S, and B (and feed F), with ternary components 1, 2, and 3. Component 1 is most volatile, followed by components 2 and 3. Mole fractions are denoted x_{ij} where i is the stream and j is the component.

A similar design, but with a condenser and reboiler also for the prefractionator, was proposed even earlier by Brugma (1937). We will denote this as a pseudo-Petlyuk design.

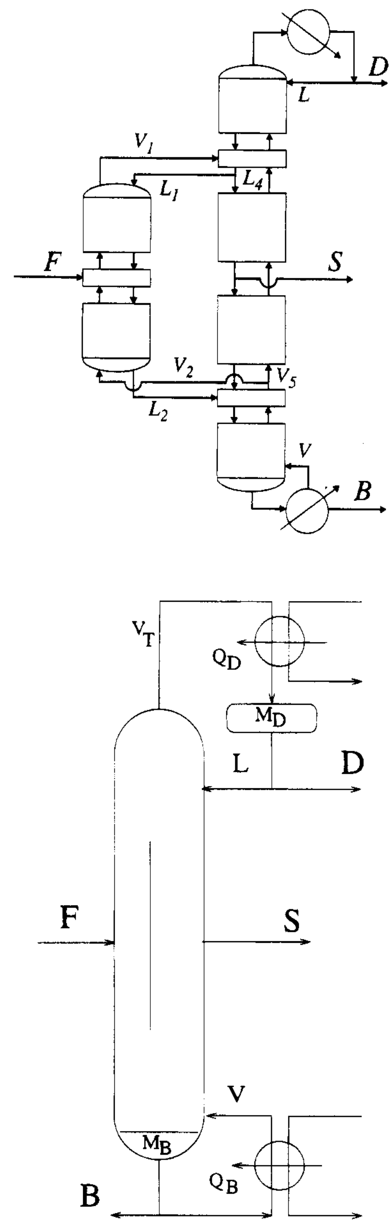

Figure 1. Petlyuk column representations: (a, top) Stream notation for Petlyuk design with prefractionator and main column. (b, bottom) Practical implementation integrating prefractionator and main column.

Cahn and Di Micelli (1962) have also proposed a Petlyuk design, although this was for separating four components by introducing *two* product sidestreams.

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Stupin and Lockhart (1972) claimed that Fenske–Underwood design computations overestimated the stage requirements and found the performance of the Petlyuk column to be rather insensitive to changes in trays and internal flows.

Tedder and Rudd (1978) were among the first to study the optimal separation of a given ternary feed. The alternatives included the direct and indirect sequence, columns with side draws, columns with side strippers and side rectifiers, and a pseudo-Petlyuk design. They found the pseudo-Petlyuk design to be preferable when the fraction of intermediate component 2 in the feed is large (40%–80%).

Cerda and Westerberg (1981) derived simple methods for estimating the operating parameters at limiting flow conditions.

Fidowski and Krolikowski (1986) compared the optimal (minimum) vapor flow rates for the direct and indirect sequence with both the Petlyuk and the pseudo-Petlyuk designs. The Petlyuk design showed significant savings. They developed analytical expressions for the minimum vapor flow rate based on the Underwood formulas.

Chavez *et al.* (1986) discussed the possibility for multiple steady states in complex columns, concentrating their work on a Petlyuk design. They found that the Petlyuk design has five degrees of freedom at steady state, and they found that four different steady-state solutions may occur when specifying three purities (in each of the products) plus bottom rate and reboiler duty. They explain this in terms of matching specifications in interlinked columns.

Glinos and Malone (1988) also derived an analytical expression for various alternative designs, including the Petlyuk design. Their recommendations are to use the Petlyuk design when the fraction of intermediate component 2 in the feed is small, and they found that the maximum vapor savings compared to simple sequences were about 50% when $x_{F2} \rightarrow 0$. They found that columns with side rectifiers may be equally well suited when the fraction of component 2 in the feed is less than 30%. However, they concluded that Petlyuk columns may also have a significant advantage for moderate or high x_{F2} values but that the conclusion depends on the relative volatilities.

Faravalli *et al.* (1989) built on the work of Chavez *et al.* and looked at which of the steady states is most resilient to changing internal flows. They applied "control" to the column, but only to aid in finding the steady-state solutions.

Triantagyllou and Smith (1992) presented a good overview of the design of Petlyuk columns and explained how it may be approximated as a regular column with two side strippers which are joined together.

The only report of an industrial implementation of a Petlyuk design is from BASF in Germany (as reported by Rudd, 1992). The Department for Process Integration at UMIST is currently investigating operational aspects of a Petlyuk pilot plant.

In this work we will study the dynamic behavior of a Petlyuk column and propose suitable controller structures. The original motivation of this project was to study composition control of the three product streams of a Petlyuk column. The results from this study, which are presented at the end of this paper, show that from a linear point of view there are no major problems.

However, during this work it became clear that there are serious problems related to the steady-state behavior that can make practical operation very difficult. The main problem is that there exist "holes" in the operating region for which it is not possible to achieve the desired product specifications. This behavior has no equivalence in ordinary two-product distillation columns.

2. Degrees of Freedom

We here consider the *degrees of freedom* (DOF) in a given column with fixed stages, feed locations, etc. Starting with binary distillation and considering a steady state where it is assumed that the holdups (condenser level, reboiler level, and pressure) are already controlled, two independent (manipulated) variables remain, for example reflux (L) and boilup (V).

In a Petlyuk column we get at steady state three additional degrees of freedom, one for each of the three additional streams leaving the main column (see Figure 2). These are the sidestream S plus the streams L_1 and V_2 sent back to the prefractionator (we will use the fractions $R_L = L_1/L$ and $R_V = V_2/V$ as DOFs in the further analysis).

Note that in this analysis the prefractionator itself does not have any degrees of freedom at steady state. The five DOFs for the Petlyuk design may be used to specify (control) the top and bottom composition (x_{D1} and x_{B3}) and one or two compositions in the sidestream. This leaves one or two degrees of freedom for optimization purposes, which we in this paper select to be minimizing the energy consumption in terms of the boilup to feed rate, V/F.

There are also possibilities for increasing the number of DOFs, for example, by taking off several sidestreams (e.g., a vapor and liquid sidestream, S_V and S_L) and by using a triple-wall solution as suggested in the figure in the paper of Petlyuk *et al.* (1965).

In a usual two-product distillation column one can at most control one specification for each product (two-point control). Simpler alternatives are no control (relying on self-regulation) or one-point control. Since the high-purity distillation columns it is critical that the overall product split is adjusted correctly (such that D/F is approximately equal to the fraction of light component), one generally finds that no control is unacceptable. One-point control, with the composition in the other end being self-regulated, is usually satisfactory if some over-refluxing (increased energy consumption) is allowed for.

For a Petlyuk scheme one must at least adjust two product splits correctly (e.g., D/F to match the light component and S/F to match the intermediate component), thus at least two-point control is required. Such a control scheme is not treated in detail here, but it is clear that it will at least require increased energy consumption compared to controlling three or four compositions. Additionally, there will be no way to adjust the separation in the prefractionator, as determined by the recycle fractions R_L and R_V.

In this paper we first study three-point control where one composition in each product is controlled, for example x_{D1}, x_{S2}, and x_{B3}. This may be an adequate control scheme. However, with only one degree of freedom to control the sidestream composition, we will not be able to adjust the ratio between the sidestream impurities, x_{S1} and x_{S3}, which may constitute an additional product specification. Thus, we finally consider four-point control with four product composition specifications (two in the sidestream).

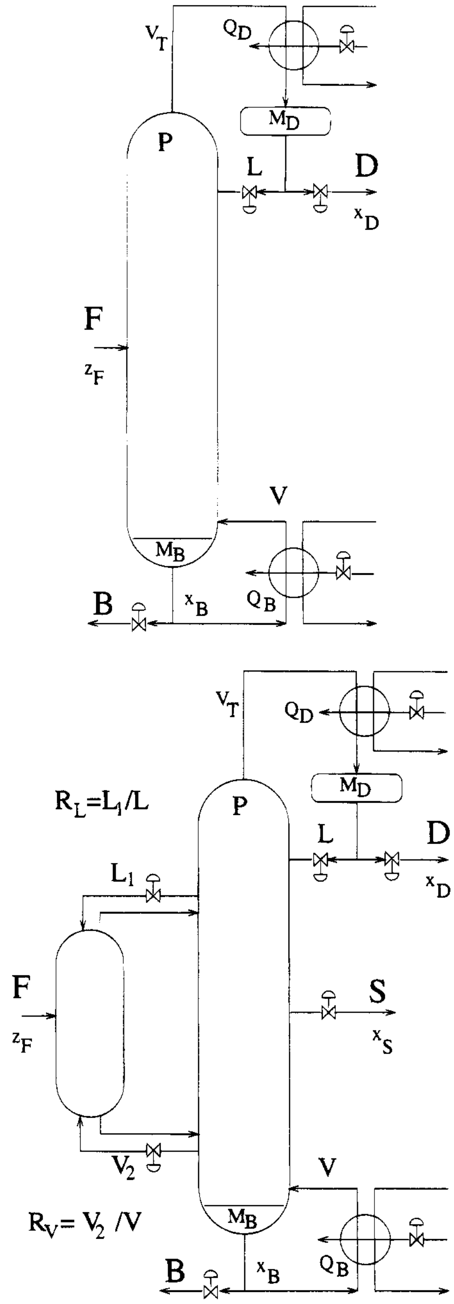

Figure 2. Degrees of freedom in distillation (indicated by valve position): (a, top) Binary distillation column. (b, bottom) Petlyuk column, implemented as prefractionator and main column.

3. Case Study

Previous authors have looked at a variety of ternary systems, from close boiling C_4 isomers to component sets spanning C_1-C_6. We have chosen the system ethanol, propanol, and butanol for the examples. The model description considers only the material balance of the system, assuming constant molar flows and constant relative volatility. A relative volatility of 4:2:1 was used for the three components, based on the geometric average of ASPENPLUS data for the top, feed, and bottom trays. The model incorporates linearized flow dynamics with a time lag from the top to the bottom of 3.6 min. The results are primarily generated with SPEEDUP, an equation-based solver. This enabled optimization, linearization, and dynamic simulations to be performed within the same environment. Steady-state simulations were also done with ASPENPLUS, a steady-state flow sheet simulator, to check the integrity of the results. Here we used the Redlich–Kwong–UNIFAC thermodynamic property set and no pressure drop.

Table 1. Relative Energy Consumption

design	boilup (%)	design	boilup %
direct	100	stripper	91
indirect	110	pseudo-Petlyuk	96
rectifier	92	Petlyuk	**87**

We have used the same number of trays in the center sections of the "main" column as in the prefractionator. This is in line with the assumed industrial implementation with a dividing wall in the shell. The "main" column consists of 40 stages, and there are 20 stages in the prefractionator. The feed is liquid at its boiling point with a flow rate of 60 kmol/min and a feed composition $x_F = [0.333, 0.333, 0.334]$. We demand 99% pure products in the top and bottom, and the design purity in the sidestream is around 99%. When having four specifications, we also demand $x_{S1} = x_{S3}$.

3.1. Comments on Model Structure and Calculations. SPEEDUP is an equation-oriented solver, i.e., all model equations are solved simultaneously and not in a modular fashion. This makes it easy to change specifications, but the solution (often) depends on good start estimates. This, together with the fact that including pressure makes distillation an even more difficult (stiff) problem to solve, largely determined the level of complexity of the model.

3.2. Economic Gain in the Petlyuk Design. Earlier work has showed that the Petlyuk design often is more energy efficient. This was confirmed for our mixture: the savings in energy compared to the standard "direct sequence" with two columns was 13% (calculations performed with ASPENPLUS). Table 1 gives an economic comparison with other ternary separation schemes and shows that the Petlyuk column operates most economically for the given components and feed composition. The other schemes were the indirect sequence, removing the heaviest component first, pseudo-Petlyuk, as described earlier, and a binary column with a side rectifier or side stripper attached, respectively. All designs consist of 60 stages, optimally distributed between the different column sections to give the least boilup.

How to determine the number of trays in a Petlyuk column for an economic comparison may be disputed. Here we have chosen to include all trays according to the model structures derived from Figure 1. This should be a conservation estimate. Counting only the trays in the main column section (assuming the prefractionator to be an integral part) leads to a relative boilup of 69% with an optimally distributed prefractionator. Both implementations (dividing wall or separate prefractionator) may be preferred at hitherto unknown operating conditions. Without going into detail, it seems reasonable to prefer one shell at high pressure where shell costs constitute more of the total.

4. Steady-State Optimal Operating Point

This section is concerned with exploiting any extra DOF to achieve optimal economic operation in terms of boilup rate. Three and four compositions are specified in the optimization problem.

4.1. Four Compositions Specified. As noted above the column has five degrees of freedom at steady state. We first study the steady-state behavior with four compositions specified: $x_{D1} = 0.99$, $x_{B3} = 0.99$, $x_{S2} =$

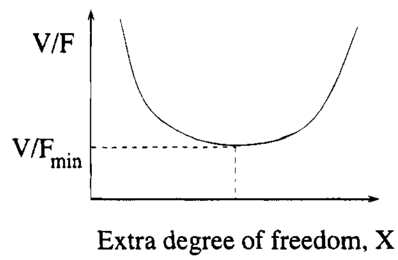

Figure 3. Expected relationship for energy usage (boilup rate, *V/F*) as function of extra DOF.

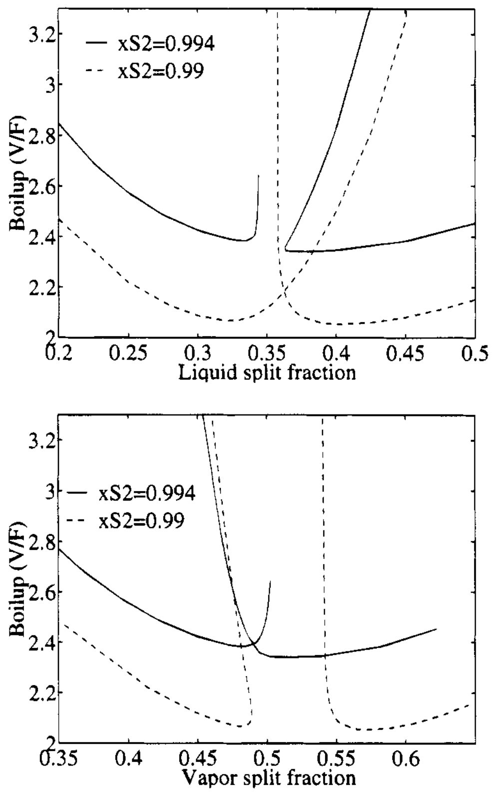

Figure 4. Boilup *V/F* as a function of internal stream splits R_L and R_V. Four product compositions specified. (SPEEDUP calculations.)

0.99, and $x_{S1}/x_{S3} = 1$ (equal distribution of the impurities in the sidestream). The purity of the sidestream (x_{S2}) is nominally 0.99, but the results are also given for other values.

One degree of freedom (DOF) then remains to be specified (denoted X in the following). For operation it is important to make a good choice of X since this variable will be kept constant or changed only slowly to minimize the operation costs, which is here selected to be given by the boilup rate (*V/F*). At first we expected to find a relationship as given in Figure 3, where *V/F* has a minimum as a function of X. Ideally, we would like the plot to be as "flat" as possible such that the exact value of X was not too important.

Unfortunately, the picture is not quite as simple in practice. This is illustrated in Figure 4a which shows boilup *V/F* as a function of $X = R_L$ (the internal reflux ratio to the prefractionator) for two values of x_{S2}. The first thing to note is that there are two possible solutions for some values of R_L. One of these corresponds to a higher value of *V/F* and should be avoided. These results are similar to those of Chavez (1986). Thus, if R_L is used as the DOF to be kept constant, the first challenge for operation and control would be to stay at the "lower" solution corresponding to the smallest *V/F*.

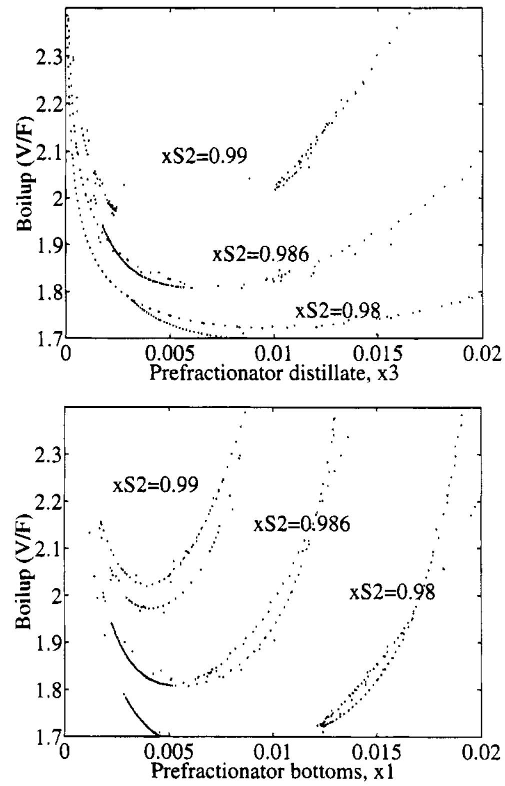

Figure 5. Boilup *V/F* as a function of prefractionator compositions. Four compositions specified. (ASPEN calculations.) (a, top) Component 3 in prefractionator distillate. (b, bottom): Component 1 in prefractionator bottoms.

Assuming that this could be done, we find for $x_{S2} = 0.99$ (dotted line) that keeping R_L at about 0.35 would be a good choice and that *V/F* would not depend too strongly on the exact value. However, for increased sidestream purity, $x_{S2} = 0.994$, there is a "hole" in the operating region, and for $R_L = 0.35$ it is not possible to achieve the desired product specifications even with infinite reflux.

It is then clear that R_L is not a good choice for the remaining DOF. To see if other choices are better we prepared similar plots for other choices ($X = R_V$ or bottom or distillate compositions for the prefractionator) shown in Figures 4b and 5a,b. However, we find that none of these are acceptable. For example, with R_V fixed we find a hole in the operating range for *low* values of x_{S2}. We also find similar problems when specifying compositions in the prefractionator.

When x_{S2} is increased slowly from 0.99 and upwards, we see that the "hole" in Figure 4a *develops*, while a "hole" simultaneously *closes* in Figure 4b. A "hole" is present for all values of x_{S2}, either for $V/F(R_V)$ or $V/F(R_L)$.

The data have been compared with results generated with ASPENPLUS. As we see from Figure 6, we get behavior very similar to the results from SPEEDUP, with holes for different x_{S2} specifications. ASPENPLUS uses more complex thermodynamics (Redlich−Kwong−UNIFAC was chosen here) with, for example, composition-dependent K-values. This and other more rigorous aspects in the ASPENPLUS calculations lead to the difference in x_{S2} values for which the holes appear. The difference in *V/F* between the two programs is approximately 4% for equal specifications. Slight changes to the relative volatilities in SPEEDUP would probably give better correspondence, both in terms of *V/F* and the specifications for which the holes develop.

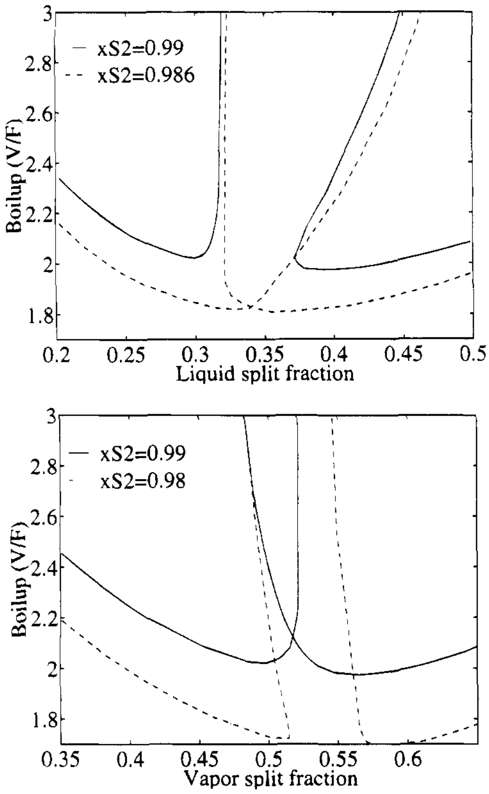

Figure 6. Boilup V/F as a function of internal stream splits R_L and R_V. Four product compositions specified. (ASPEN calculations.)

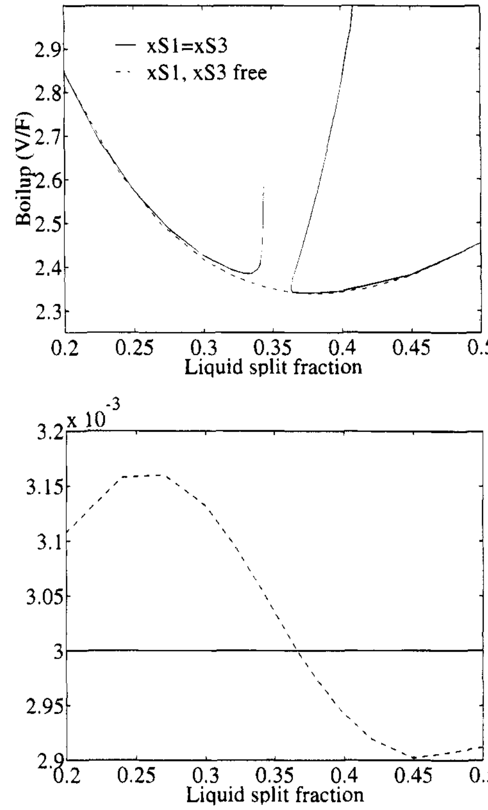

Figure 7. (a, top) Boilup rate V/F and (b, bottom) sidestream impurity x_{S1} as a function of R_L when relaxing the constraint $x_{S1} = x_{S3}$ and instead using the additional degree of freedom to minimize boilup (V/F). $x_{D1} = 0.99$, $x_{B3} = 0.99$, and $x_{S2} = 0.994$.

4.2. Three Compositions Specified. The conclusion from the above plots is that "holes" in the operating range will make it very difficult to control four compositions. A possibly better alternative is to control only three compositions, that is, to let the ratio of the impurities in the sidestream vary freely (and not specify $x_{S1}/x_{S3} = 1$ as above). This yields another DOF that must be specified, for example, one may select $X_1 = R_L$ and $X_2 = R_V$. This is the choice made in the control part later.

Obviously the removal of one specification "loosens up" the problem somewhat, letting the column profile vary more freely. For example, while it was impossible to achieve $x_{S2} = 0.994$ with $R_L = 0.35$ and $x_{S1}/x_{S3} = 1$, we find that we can achieve $x_{S2} = 0.994$ with $R_L = 0.35$ and $R_V = 0.49$, for example, giving $x_{S1}/x_{S3} = 1.02$ and $V/F = 2.35$. Figure 7a shows how the hole has disappeared when the constraint $x_{S1} = x_{S3}$ is relaxed. No solutions corresponding to another branch were found. Figure 7b shows how one of the sidestream impurities x_{S1} has varied to accomplish the minimum boilup in Figure 7a.

Specifying three compositions gives a lower minimum boilup than with four compositions specified. The difference between the two cases in Figure 7a is small. This results from the column "naturally" preferring $x_{S1} \approx x_{S3}$ with the given tray distribution and side draw position (also evident from Figure 7b).

To summarize: the Petlyuk column exhibits holes in the operating space when four product compositions, including two in the side draw, are specified. The location of these holes in terms of values of R_L and R_V varies with design and specifications. The lower bound on the boilup for the four-specification set is given by the three-specification solution.

5. Control of the Petlyuk Column

In the remaining part of the paper we consider control of the column using decentralized control. The reflux (L) is used to control the top composition (x_{D1}), the boilup (V) is used to control the bottom composition (x_{B3}), and the sidestream flow rate (S) is used to control sidestream composition (x_{S2}). For "three-point" control, R_L and R_V are fixed. For "four-point" control, R_L is used to control the impurity ratio (x_{S1}/x_{S3}) with R_V fixed. We also consider using multiple side draws in control of four compositions.

The nominal operating point studied has $R_L = 0.35$, $R_V = 0.49$, $x_{D1} = 0.99$, $x_{B3} = 0.99$, and $x_{S2} = 0.994$.

5.1. Linear Analysis Tools. In the following we will use a plant description of the form

$$y(s) = G(s)\,u(s) + G_d(s)\,d(s) \qquad (1)$$

where G and G_d denote the process and disturbance plant model and y, u, and d are the measurements, manipulated inputs, and disturbances, respectively.

In this paper we mainly use the relative gain array (RGA or Λ) to study interaction in the distillation column. The properties of the RGA are well-known (e.g., Grosdidier *et al.*, 1985). The most important for our purpose are (1) no two-way interaction is present when $\Lambda = I$, (2) the RGA is independent of scaling in inputs or outputs, and (3) the rows and columns both sum to 1. To evaluate the disturbance sensitivity, we consider the closed loop disturbance gain (CLDG) which is the appropriate measure when we use decentralized control (Hovd and Skogestad, 1992). The CLDG is defined as $\Delta = G_{\mathrm{diag}}G^{-1}G_d$, where G_{diag} consists of the diagonal elements of G. For decentralizd control, frequency-dependent plots of $\Delta = \delta_{ik}$ are used to evaluate the necessary bandwidth requirements in loop i, that is, at low frequencies the loop gain $L_i = g_{ii}c_i$ must be larger than δ_{ik} in magnitude to get acceptable performance. We also look at the singular value decomposition $G = U\Sigma V^T$ and examine the elements of G.

The disturbances considered are changes in the feed flow and feed composition. All variables have been scaled with respect to the maximum allowed change for judging input constraints and performance: $\Delta L = \Delta V = 30\%$, $\Delta R_L = \Delta R_V = 0.2$, $\Delta S = 25\%$, $\Delta x_{ij} = 0.01$, $\Delta F = 17\%$, and $\Delta z_{Fi} = 0.2$.

5.2. Analysis of Three-Point Control: The *LVS* Configuration. In this case R_L and R_V are fixed and the outputs and inputs are

$$y = \begin{pmatrix} x_{D1} \\ x_{B3} \\ x_{S2} \end{pmatrix} \qquad u = \begin{pmatrix} L \\ V \\ S \end{pmatrix} \tag{2}$$

The Petlyuk column at the operating point with minimum energy use ($R_L = 0.376$ and $R_V = 0.517$) has transmission zeros in the right half plane (RHP). The most dominant zero is at a frequency of 2.53 rad/min. Thus, there may exist fundamental problems with achieving a fast response (high bandwidth).

A RHP zero occurs between S and x_{S2} at this operating point for the following reason: changes in S will behave toward x_{S3} almost as an integrator through changes in S initially only affecting the liquid below the tray. The initial change in x_{S1} is larger than that for x_{S3}, but it will then level off after a short time. The relation between these two responses will determine if an inverse response occurs or not. The subsequent analysis will reveal that these RHP zeros will only slightly impair the high-frequency behavior and not prove a control problem.

The steady-state gain matrix G is

$$G(0) = \begin{pmatrix} 153.45 & -179.34 & 0.03 \\ -157.67 & 184.75 & 21.63 \\ -4.80 & 6.09 & -2.41 \end{pmatrix}$$

We see that the sidestream S mainly affects the middle and bottom product, while both L and V have a large effect on x_{S2}. We see quite readily that there will be interaction between the top and bottom composition, in line with ordinary binary distillation.

The singular value decomposition $G = U\Sigma V^T$ (at steady state) will allow us some conclusions on the high and low gain directions of the plant. The output and input directions are given (as the columns) in U and V, respectively, and the singular values are $\Sigma = \mathrm{diag}[339.12\ 15.33\ 0.34]$.

$$U = \begin{pmatrix} 0.70 & -0.70 & 0.14 \\ -0.72 & -0.69 & 0.11 \\ -0.02 & 0.18 & 0.98 \end{pmatrix}$$

$$V = \begin{pmatrix} 0.65 & -0.04 & 0.76 \\ -0.76 & 0.03 & 0.65 \\ -0.05 & -1.00 & -0.01 \end{pmatrix}$$

We see that the high gain output direction (column 1 of U) corresponds to moving the top and bottom compositions in opposite directions or moving the column composition profile up or down. (Remember that we look at three compositions, not only the composition of light component in the top and bottom as in binary distillation.) x_{S2} is not affected much in this case,

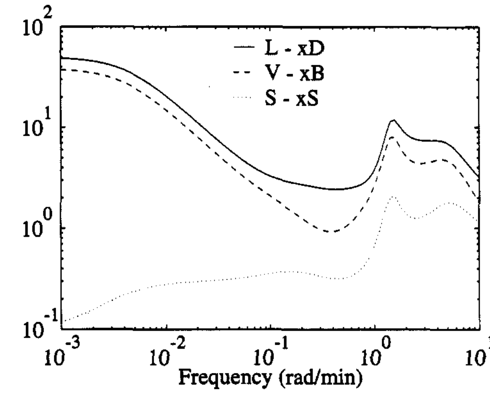

Figure 8. Relative gain array, λ_{ii}, for the *LVS* configuration.

corresponding to trading x_{S1} with x_{S3}. The low gain direction (column 3 of U) corresponds to moving them in the same direction, i.e., making both D and B more or less pure. This is in accordance with ordinary binary distillation. The medium gain direction corresponds almost entirely to changing S (column 2 in V) and moves x_{S2} opposite to x_{D1} and x_{B3}. This is reasonable; for example reducing S will give a buildup of component 2 around the side draw, giving more impurities in both the top and bottom sections of the column.

We can predict from this that using S for control will require countermeasures to keep both x_{D1} and x_{B3} at their setpoints. Moving the column profile in the high gain direction will leave x_{S2} quite immobile, whereas the low gain direction (making D and B more pure) will significantly affect x_{S2}.

We then study the interaction and disturbance rejection properties. The steady-state RGA values

$$\Lambda(0) = \begin{pmatrix} 49.83 & -48.83 & 0.00 \\ -38.34 & 38.44 & 0.90 \\ -10.48 & 11.38 & 0.10 \end{pmatrix}$$

show again that the control of x_{D1} and x_{B3} interact. The same trend is evident from the frequency-dependent RGA as shown in Figure 8. The interaction tapers off at higher frequencies, showing that the control having an effect around the bandwidth of the plant will not be much affected by interaction. Pairing on very low RGA values is generally not advisable, thus questioning a pairing on $\lambda_{33} = 0.1$. Again, however, the medium frequency values are better (Figure 8), giving confidence to the indicated pairing.

The RGA indicates that another feasible pairing than the one mentioned exists, namely, using boilup to control the sidestream purity and the sidestream to control the bottom composition (denoted as the *LSV* configuration). This again indicates that changes in S, through mainly affecting the liquid flow below the side draw, primarily interact with the lower part of the column.

The closed loop disturbance gain, CLDG, is shown in Figure 9 for the pairing indicated in eq 2. The most difficult disturbance to reject is changes in F on x_D and x_B requiring a bandwidth of about 0.25 rad/min (time constant of 4 min) in these loops. On the other hand, the required bandwidth for controlling x_{S2} is significantly smaller (less than 0.1 rad/min). This gives us that side draw control loop can be tuned more loosely than the top and bottom composition loops. We also note that the bandwidth requirements are much lower than for the RHP zero at 2.53 rad/min, indicating that the RHP zero is not a practical limitation.

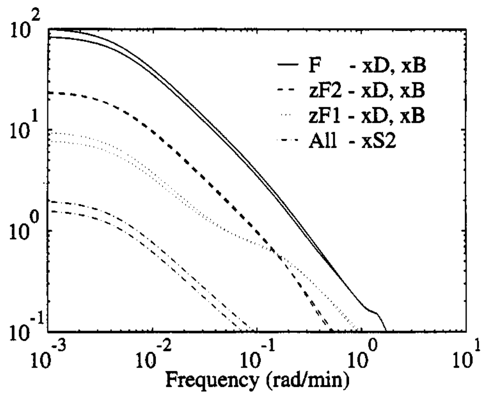

Figure 9. Closed loop disturbance gain, δ_{ij}, for the *LVS* configuration.

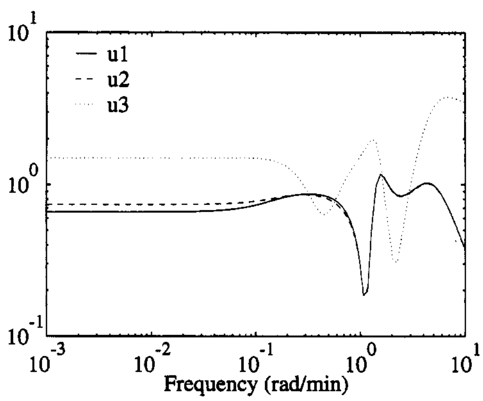

Figure 10. Necessary control action for perfect control, $(G^{-1}G_d)_i$.

Table 2. Controller Tunings, *LVS* Configuration

controller	$K \times 100$	τ_I (min)
x_{D1}	50	18
x_{B3}	50	18
x_{S2}	−5	30

Examining the CLDG for the reversed pairing mentioned above gives approximately the same bandwidth requirements (not shown). In addition, x_{S2} becomes more sensitive to disturbances in general and x_B less sensitive to changes in F.

Figure 10 shows the necessary control action for perfect disturbance rejection at all frequencies. It is worth noticing that $\|G^{-1}G_d\|_\infty$ (the largest row sum) is only slightly over 1, meaning that good control is possible without violating the constraints. The curve also shows that the required input changes are largest for u_3, the sidestream flow rate.

The CLDG and $G^{-1}G_d$ complement each other nicely; the first indicates the most difficult disturbance, the other which actuator is the primary limitation to perfect control under disturbances.

5.3. Nonlinear Simulations with Three-Point Control. The conclusion is that from a linear point of view the process is easy to control at this operating point. There might even be several good pairings for control. We next perform nonlinear simulations to test these conclusions within a "normal" operating range. Frequency response based tuning rules (Ziegler–Nichols) were not applicable due to the low high frequency phase shift ($\angle G(j\omega) \le \pi$). The controller tunings in Table 2 were found to give acceptable performance and actuator behavior.

The simulation results in Figure 11 show the closed loop response to disturbances in F (60 → 50) and z_F

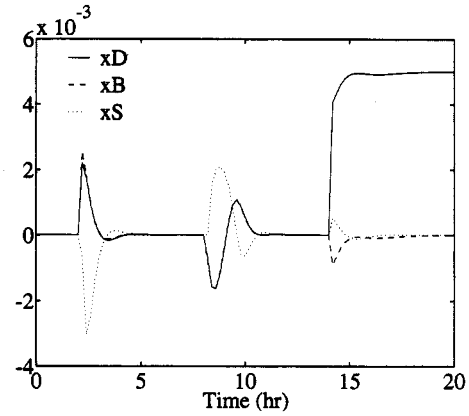

Figure 11. Time response to disturbance and setpoint change, *LVS* configuration. ΔF at $t = 2$, Δx_F at $t = 8$, and $\Delta y_{D1,s}$ at $t = 14$. Plot shows deviations from steady-state values.

([0.33, 0.33, 0.33] → [0.33, 0.40, 0.27]) and a distillate purity setpoint change (0.99 → 0.995), respectively. The Petlyuk column handles disturbances and setpoint changes well.

Although the system seems resilient, a setpoint change in x_{S2} of 0.994 → 0.996 is infeasible for this operating point (L and V increase >100%), showing that three-point control may have problems from some range of R_L and R_V.

The reason is probably that the column does not have enough stages to achieve this degree of separation, at least not without adjusting R_L and R_V.

It is also worth noting that binary columns, using a design rule like $N_{\mathrm{trays}} = 2N_{\mathrm{trays,min}}$, can handle purity increases of almost an order of magnitude, i.e., 0.99 → 0.999 without reaching the limit of $R_{\max}$. This does not look equally possible for the sidestream S in the Petlyuk column, where a near stationary x_{S1} limits the possible increase in x_{S2} to at most the absolute value of x_{S3}. This will, however, depend on the design.

The pairing with V and S actuators exchanged (*LSV* configuration) seemed equally feasible from the linear analysis. However, the control system with this pairing failed against the mentioned perturbation set. The reason for this failure is the strong nonlinearity from V to x_{S2}.

5.4. Analysis of Four-Point Control: The LVR_LS Configuration. R_L is added as a manipulated variable and is used to control x_{S1}. The set of measurements and manipulated variables is thus

$$y = \begin{pmatrix} x_{D1} \\ x_{B3} \\ x_{S1} \\ x_{S2} \end{pmatrix} \qquad u = \begin{pmatrix} L \\ V \\ R_L \\ S \end{pmatrix}$$

The process gain and RGA at steady-state operating conditions are

$$G(0) = \begin{pmatrix} 153.45 & -179.34 & 0.23 & 0.03 \\ -157.67 & 184.75 & -0.10 & 21.63 \\ 24.63 & -28.97 & -0.23 & -0.10 \\ -4.80 & 6.09 & 0.13 & -2.41 \end{pmatrix}$$

$$\Lambda(0) = \begin{pmatrix} 23.84 & -22.95 & 0.11 & 0.00 \\ -49.01 & 49.09 & 0.02 & 0.90 \\ 39.23 & -39.31 & 1.08 & -0.00 \\ -13.06 & 14.18 & -0.21 & 0.10 \end{pmatrix}$$

We see that although a suitable pairing exists ($L \rightarrow x_{D1}$,

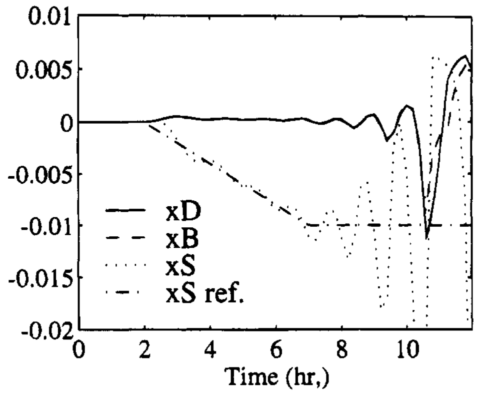

Figure 12. $x_{S2} = 0.994 \rightarrow 0.992$ setpoint ramp decrease from $t = 2$ to $t = 6$, LVR_LS configuration.

$V \rightarrow x_{B3}$, $R_L \rightarrow x_{S1}$, and $S \rightarrow x_{S2}$) the manipulated variable R_L has a very low gain toward *all* control objectives. The largest gain is about 0.23, which means that the input signal needed to reject disturbances will be approximately 4–5 times the assigned bounds. The closed loop disturbance gain is nearly identical to the three-point control case. The additional measurement x_{S1} is insensitive to all disturbances, having values below 0.2 at all frequencies and thus not needing control at all for disturbance rejection. Conclusion: x_{S1} is insensitive to both inputs and disturbances.

5.5. Problems with Four-Point Control. It was predicted earlier from a nonlinear steady-state analysis that four-point control may experience problems with fixed values of either R_L or R_V. This is indeed the case, and Figure 12 shows the result of a setpoint *decrease* in x_{S2} with fixed $R_V = 0.517$; the column becomes unstable. Reflux and boilup reach the imposed constraints (+100%) without managing to hold the specifications. The reason for this is that R_L, when used for control, is reduced to comply with the specification on x_{S1}. This in turn brings R_L into the hole, causing the specifications set to become infeasible. The difficulties with operating in areas corresponding to "holes" in the $V/F(X)$ plots seriously limit four-point control, despite good disturbance rejection properties.

5.6. Four-Point Control with Multiple Sidestreams: The LVR_SS Configuration. As mentioned in section 2, several sidestreams can be withdrawn from the column to add degrees of freedom for optimization or control. Here two liquid sidestreams S_1 and S_2 (which are then mixed to a single side product) are considered for improved control over the sidestream impurities.

Consider the column composition profile in Figure 13. We see that component 2 creates a bell-shaped curve with a maximum around the side draw. If it is of interest to keep the impurities at a prescribed level, having two separate product streams will increase the flexibility (through the choice of blending). Envision (for clarity) that the product streams are withdrawn from trays 15 and 25. If the blend of these two streams is to follow the specification $x_{S2} = 99.4\%$, then obviously x_2 must be equal or above 99.4% (i.e., overfractionation) over a wider span of trays around the side draws. The point here is that increased flexibility in the side draw product specification gives increased utility use through some overfractionation. This holds for the comparison with an optimal side draw placement but might be reversed for bad designs where the introduction of a second draw point relieves a "stressed" profile.

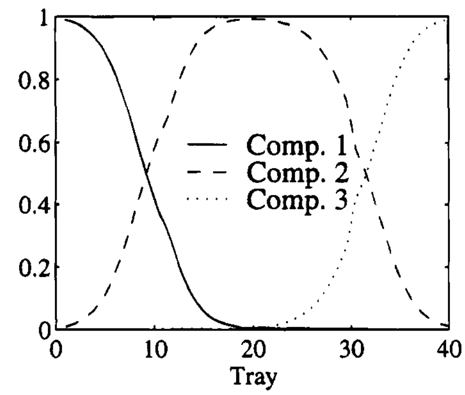

Figure 13. Composition profile in main column section.

Table 3. Economic Sensitivity to Placement of Multiple Side Draws

configuration	V/F	highest x_2
base case	2.34	0.994
1 tray separating side draws	2.44	0.995
3 trays separating side draws	2.72	0.996

To make the discussion conceptually simpler, the two product streams are mixed and the fractional content of the upper side draw, denoted R_S, is introduced. Thus, the set of actuators becomes $u = (L\ V\ R_S\ S)^T$. This is closely related to the LVR_LS configuration in that a stream distribution is used to manipulate the sidestream impurities, only the streams are external, not internal to the column.

The flexibility of x_{S1}/x_{S3} is limited by the compositions of the separate draws, that is, the cases $R_S = 0$ or $R_S = 1$. Clearly, side draws far from each other will allow more leverage with respect to x_{S1}/x_{S3} than side draws next to each other.

We first examine the economic penalty of having separate draws (within the same total number of trays). Table 3 shows the added boilup from having 1 and 3 trays between S_1 and S_2 compared with the base case data. We see that within the allotted amount of trays, attaining the required separation quickly becomes expensive. Table 3 also shows the highest x_2 between the side draws. As mentioned earlier, some overfractionation is necessary in the column for the side product to reach $x_{S2} = 0.994$.

In the following, S_1 and S_2 are separated by one tray (combined, the total side product is again S), giving $x_{S1} = 0.0046$ and 0.0016, respectively, when $R_S = 0$ or $R_S = 1$ and $x_{S2} = 0.994$ is specified. The analysis of this configuration shows that R_S resembles R_L with respect to gain and interaction and is thus not suited for control. The gain matrix G,

$$G = \begin{pmatrix} 141.56 & -166.91 & -0.20 & 0.02 \\ -145.22 & 171.63 & 0.13 & 21.88 \\ 22.91 & -27.17 & 0.21 & -0.09 \\ -4.19 & 5.35 & -0.07 & -2.55 \end{pmatrix}$$

which corresponds to the following measurements and actuators

$$y = \begin{pmatrix} x_{D1} \\ x_{B3} \\ x_{S1} \\ x_{S2} \end{pmatrix} \qquad u = \begin{pmatrix} L \\ V \\ R_S \\ S \end{pmatrix}$$

is visibly similar to the gain for the LVR_LS configuration. Thus, the LVR_SS configuration cannot be recom-

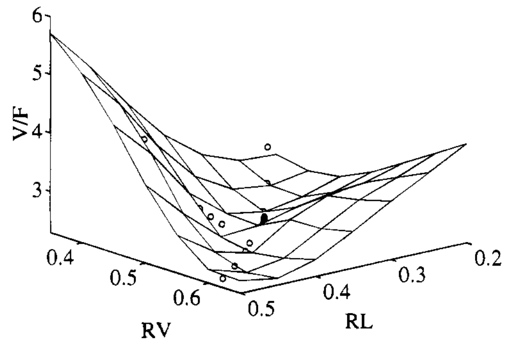

Figure 14. Surface plot of solution points for boilup as a function of R_L and R_V.

mended for four-point control. The solution with multiple sidestreams will give increased flexibility, though.

6. Discussion

6.1. Possible Operational Problems. Seider *et al.* (1990) discussed complex nonlinearities and questioned the stability of the Petlyuk column on the different solution branches (the work only uses this as an illustrative example). This work has not found any indications of this property when looking at the system eigenvalues for the linearized plant on all solution branches.

The steady-state results and simulations in this paper are based on the condition that R_L, R_V, or both can be measured (and manipulated) to a certain accuracy. This section discusses possible problems in view of common control limitations and current practice.

R_L and R_V are both ratios although flows are the real static or dynamic variables. As with the $(L/D)(V/B)$ control scheme for binary distillation, ratio control demands more instrumentation than corresponding single flow rate based schemes (LV control in the binary case). For the Petlyuk column this is equivalent to controlling only one stream out of the interconnection trays. For the liquid split at the top of the prefractionator, controlling only one stream is feasible, since tray overflow is self-regulating.

R_V does not easily lend itself for control since this involves manipulating vapor flows in a nonconfined area (i.e., not pipes), and the current implementation at UMIST leaves R_V uncontrolled. Controlling column pressure in distillation can usually be done within 2% (Tolliver and McCune, 1980) allowing for sensor error and actuator uncertainty. Similar results might be attainable with pressure drop "vents" given that the design is suitable.

The economic sensitivity to erroneous values for R_L or R_V is evident from tracing their optimal values in a three-dimensional plot. From the mass balance over the column the relation between R_L and R_V is found to be

$$R_L = (V/L)R_V - (1/L)D' \qquad (3)$$

where D' is the net distillate from the prefractionator ($V_1 - L_1$ in Figure 1a). Thus, the sensitivity to uncertainty in R_L and R_V depends on their mutual distribution. From Figure 14 we see that the worst error is when $\Delta R_L = -\Delta R_V$ over most of the parameter space.

It is apparent that the parameter sensitivity is design dependent, varying with changes in the number of overall trays, the tray distribution or design vapor speed. Excessive overdesign (many more trays than optimal) may introduce "pinch zones" in the column where the composition hardly changes. This will "buffer" disturbances or decouple the column sections above the below the pinch.

This work has so far evaluated direct composition control of three or four compositions. In many distillation systems, however, a single control loop is often implemented to stabilize the column composition profile, typically a tray temperature. See for example the work of Wolff and Skogestad (1993) for a discussion on the effects of using cascade control in distillation. There are more possible alternatives in the choice of which tray temperature to control in the Petlyuk column than in ordinary binary distillation. The obvious choice is to control a tray temperature close to the most critical composition measurement, giving an overall stabilization of the composition profile. Choosing a tray temperature closer to the sidestream may, however, be better since the continuous composition profile of *three* components gives a two-way temperature dependence, not only one-way.

6.2. Control Comparison with Conventional Configurations. The Petlyuk column has been presented as a viable solution for ternary separation. This section compares the Petlyuk column with more common ternary distillation sequences.

6.2.1. Speed of Response. The direct and indirect designs consist of two complete distillation columns, which together with holdups and piping will certainly give a larger dead time through the separation section. This may not be important, just like the size of the dead time in unmeasured disturbances usually does not matter. However, with the increased implementation of supervisory control (in addition to for example a DCS), setpoint changes are more frequently imposed. Quicker plant settling may enable updates to be performed more often and indirectly improve economics of the Petlyuk scheme.

6.2.2. Specifying Four Compositions. It has been demonstrated that the Petlyuk scheme may *not* be suitable if it is desirable to specify both impurities in the sidestream independently (corresponds to specifying two sidestream compositions).

6.2.3. Flexibility. Increased product specifications are probably easier dealt with in ordinary distillation, where common design rules allow a large degree of overfractionation. In the Petlyuk design the sidestream is especially sensitive to increased purity demands. Feed disturbances are handled differently depending upon the control configuration. Using the $(L/D)(V/B)$ configuration for the top and bottom composition control, for example, will probably be more resilient to feed flow disturbances than the LV configuration. The degree of optimality to permanent changes (for example through a debottlenecking) is not clear.

6.2.4. Maintenance. Using a conventional arrangement includes more holdups and thus more control loops. This results in greater maintenance tasks as well as larger changes of problems such as pump failure or actuator wear. It must be noted though that design and operation of actuators for manipulating R_L and R_V may prove difficult.

6.3. Design Suggestions. This section summarizes discoveries important in design of the Petlyuk column.

6.3.1. Internal Flows. Because of the many interconnections in the Petlyuk column, there are also many different liquid and vapor flow rates. Optimal operation will depend on the internal flow regimes so these should

Table 4. Economic Sensitivity to Side Draw Position with 3 and 4 Product Specifications

side draw	V/F (4 prod)	V/F (3 prod)
2 up	2.93	2.67
1 up	2.55	2.43
base case	2.34	2.34
1 down	2.60	2.35
2 down	2.66	2.45

be investigated, also considering possible flooding and weeping. High internal flow rates in the main column section will minimize the disrupting effect on the composition profile from the prefractionator product. Conversely, very low internal flows will easily give an unbalanced system under external influences (for example, a disturbance in feed enthalpy).

6.3.2. Sidestream Placement. To enable fine tuning of the sidestream composition, multiple side draws and blending opportunities are advised. This is especially important where a wide range of feed compositions are treated and when four product specifications are imposed. The side draw placement is also important for economic operation, given that wrong placement of the side draw can make a big difference in the boilup. Table 4 shows how V/F depends on the side draw placement for three and four specifications. The boilup is less sensitive with only three specifications since x_{S1}/x_{S3} can then vary freely. This is in line with the low sensitivity to the internal splits R_L and R_V.

6.3.3. Startup. Equation 3 is also useful for determining R_L and R_V for the startup period (normally total reflux). D' is obviously zero, and L and V are equal, giving $R_L = R_V$. This is reasonable since the best separation is usually when the liquid and vapor flows balance. There may be other startup issues that can significantly reduce the time it takes to establish the correct column composition profile.

7. Conclusion

In this paper, we considered separation of a ternary mixture into three products using a Petlyuk column. The Petlyuk column offers possibilities for savings both in terms of capital and energy costs. However, it has not been widely used due to problems with design and operation.

In this paper we show that with four compositions specified the Petlyuk column may display complicated behavior with multiple internal distributors and "holes" in the operating range corresponding to infeasible specifications. Thus, the Petlyuk design should probably be avoided if it is required to specify the two sidestream impurities independently.

With three product specifications (with only the total impurity in the sidestream specified) the Petlyuk column does not seem to present such problems and acceptable control seems feasible. However, in this case one also needs to adjust two additional degrees of freedom (e.g., R_L and R_V) properly in order to achieve the energy benefits. More work is needed to study how this should be done in practice. Work is also needed to understand the origin of the "holes" in the operating range.

Finally, there is an abundance of DOFs when counting both design and control and all have not been exploited here. The design of the Petlyuk column is a difficult task, and developments here may aid the operational problems encountered.

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