This would suggest that it may be useful as a refinement method to examine the most likely fault conditions as produced by the diagnostic model processor. In this manner, the single-fault hypothesis can be examined (noting its application) with

hypotheses as follows:

$$F(a_1) = 1/\sqrt{3} \approx 0.58$$

$$F(a_2) = 1/\sqrt{3} \approx 0.58$$

$$F(a_3) = 1$$

secondary constraints, his correction is statistically accurate. Equation 6 in Petti et al. is used as a more conservative estimate of the secondary tolerance to
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To the Editor:

This letter is in response to the paper by Skogestad, Lundström, and Jacobsen titled "Selecting the Best Distillation Control Configuration" (May 1990, p. 753). In the paper titled "Control Configuration Selection for Distillation Columns" (October 1987, p. 1620), Skogestad and Morari used a spectrum of distillation column cases to analyze control performance since the dynamic behavior and resulting control performance can change significantly with relative volatility, product purity, and feed composition. I like the idea very much.

Distillation columns with a large relative volatility (e.g., $\alpha > 3$) are characterized by very fast dynamic responses since the reflux to feed ratios are usually low. Fortunately, a large relative volatility is accompanied by a relatively large temperature drop across the column which facilitates the use of tray temperature to infer product composition. As a result, the detrimental effect of composition analyzer deadtime (3 to 12 minutes considering industrial sample transport delays) can be negated by controlling tray temperature (which responds quickly) and choosing the setpoints for the tray temperatures from the composition analyzer updates.

Columns with a low relative volatility (e.g., $\alpha < 1.3$) typically use relatively high reflux to feed ratios and are quite slow responding processes. These columns do not have a large enough temperature drop across them to allow the effective use of tray temperatures to infer product composition. But the analyzer deadtime to process time constant ratio is sufficiently low so that the deadtime does not adversely affect control performance.

In light of the above perspective, I have a number of questions about several of the modeling assumptions made in this paper.

I. It is assumed that the molar holdup on each stage to feed rate ratio is a constant 0.5 minute for all cases considered. At a constant feed rate and constant product compositions, as the relative volatility is decreased, the resulting column diameter will increase due to the increase in vapor-liquid traffic. As a result, the molar holdup to feed rate ratio will be larger for low relative volatility columns than for high relative volatility columns. In addition, it is common practice to design a high relative volatility distillation column with a rectifying section with a smaller diameter than the stripping section since the liquid traffic in the rectifying section is typically lower. For example, I have seen a C_2/C_3 splitter for which the molar holdup in the rectifying section was about half of that in the stripping section. The variation in M_i/F between low and high relative volatility columns is significantly larger than the latter example and could amount to an order of magnitude variation in M_i/F for the cases considered in this paper. This difference in M_i/F would cause very significant differences in column dynamics between low and high relative volatility columns unlike the results shown in the paper.

II. The authors assumed that the liquid holdup in the reboiler and the overhead accumulator were also each equal to the molar holdup on a single tray. The authors offer the following reasons for that assumption. "1) If temperatures are used to estimate composition, then the dynamic response *inside* the column matters and this response is usually weakly affected by the reboiler and condenser holdups (Skogestad and Morari, 1988a). 2) The holdups (levels) in the condenser and reboiler may vary with time and it is then safest (from a robustness viewpoint) to use the smallest holdup when designing the controllers (Skogestad and Morari, 1988b)."

With regard to the first reason, when tray temperatures are used to infer product compositions they are usually located a number of trays away from the product end. Also, the residence time of an industrial accumulator can be substantial (as much as an hour). Then, for example, after the reflux rate is changed, the composition on the tray used to infer overhead composition will not reach its final value until the composition of the accumulator has adequately approached its new value. As a result, the molar holdup in the accumulator and reboiler will introduce a major lag into the dynamic response of the process. Simulations based upon a low holdup in the accumulator and reboiler would seem to have a considerably faster response than an industrial column.

For the second point, while a fast responding system is more sensitive to deadtime and nonstationary behavior, distillation columns, which industrially have large reboiler and accumulator holdups, exhibit a highly lagged, overdamped response which is a formidable control problem in their own right. Consider the results for Column D shown in Figure 8 of their paper for a feed composition upset and a feed flow rate change. For the reflux ratio/boilup ratio control, the largest relative error in y was about 2% while the time to return to setpoint was about 15 minutes. Since Column D has a relative volatility of 1.12, one would have to rely upon composition analyzer updates for composition measurements and there would be about a 5-minute analyzer deadtime. In industrial columns of this type, with a normal-sized accumulator, it would take at least 15 minutes before the feedback control was aware that there was a significant deviation from setpoint. The simulated low holdups for the accumulator and little or no deadtime probably contributed to the use of the high controller gains (the L/D, V/B controller gains were each 113 or 30 times that used by the other controllers). With the lag of the accumulator and reboiler, I would expect that the control response for this case would be considerably different since using a high controller gain on a system with a lot of lag can cause oscillatory or even unstable behavior.

III. It was stated that the controller settings developed for setpoint changes were somewhat sluggish for disturbance rejection. This is a common observance with linear controllers. In theory, a SISO PID controller could be applied for disturbance rejection by raising the controller gain in order to keep the process at the point for which it was tuned. In practice, if a PID controller is highly tuned at a point for disturbance rejection, it will sooner or later deviate from setpoint enough to cause poor control performance. As a result, control engineers must "detune" their controllers in order to insure reliable performance.

IV. Four of the cases considered use a relative volatility of 1.5. This corresponds to a ethylene splitter and represents a lower limit on using tray temperatures to infer product compositions. As a result, using tray temperatures to infer product compositions is not as accurate as for the other cases that have higher relative volatilities. It appears to me that case B would pose quite a difficult control problem; and since the feed contains only 10% light material, it would be difficult to use a tray temperature in the bottom part of the column to infer bottoms compositions since there would be a relatively small temperature drop across the stripping section. As a result, one would have to rely upon the composition analyzer for bottom composition feedback. The analyzer deadtime is likely to significantly affect the control performance since the reflux to feed ratio is not particularly high.

V. While I have these questions about the assumptions used in the simulated results, I concur with the author's conclusion that V/B, L/D control should provide the best dual composition control performance. From a McCabe Thiele diagram, one can see that the product purity changes can be affected by changing the slope of the operating line. The slope of the operating line in the stripping section is directly related to V/B while in the rectifying section it is related to L/D. As a result, by maintaining these ratios during upsets, significant decoupling between the ends of the column is provided. For example, if the boilup rate increases, maintaining the vapor/liquid ratio in the rectifying section helps reduce the effect of the boilup rate change on the top of the column. The advantages of this approach can be seen by examining the behavior of LV control to a boilup rate change.

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Reply:

In our paper (Skogestad et al., 1990) we demonstrate the effectiveness of the frequency-dependent RGA as a tool for selecting the best distillation column configuration. These results are general, but in the examples we naturally chose specific values for relative volatility, holdups, etc. Although the example columns were chosen to represent a range of columns, it is not possible to cover all combinations of parameters that may be encountered in practice. In particular, we agree that it is unrealistic that the holdups are the same in all example columns. This was done intentionally to keep the presentation clearer.

Nevertheless, we are happy for the comments made by Dr. Riggs because this gives us the opportunity to clarify some issues which we were not able to discuss in our paper. Riggs considers the two cases: a) large relative volatility columns with fast dynamic response; and b) low relative volatility columns with a relatively slow dynamic response. It should be noted, however, that although it is true that low volatility columns generally have a slow *open-loop* response, it does not imply that the response under feedback control has to be slow. This is a myth which is very often encountered among operators. Note that the overall lag for liquid response, θ_L, which is one of the most important dynamic parameters for feedback control of distillation columns, is not generally large for low volatility columns. We agree that the signal to noise ratio makes it simpler to use tray temperatures to infer compositions in case a, but we do not agree that it is not possible to use temperature measurements in case b where the temperature difference between top and bottom is small. In fact, the signal to noise ratio may be improved by adding more temperature measurements as discussed in the following.

I. Holdup on Stages. We agree that the molar holdup to feed rate (M_i/F) usually is larger for columns with a large reflux. There is, however, no general relationship because the holdup also depends on the particular packing or tray chosen. In fact, we have observed similar values of $M_i/F \approx 0.5$ min in real columns where L/F has been as different as 3 and 20. In any case increasing the holdup will simply amount to changing the time scale in the results described in our paper. To obtain the holdup in a practical situation it is probably simplest to observe the time, θ_L, it takes for a change in reflux to affect reboiler level. For packed columns the overall liquid holdup *inside* the column, M_i, may then be estimated as $M_i/L =$

$\theta_L/0.6$ where L is the average reflux in the column (Skogestad and Morari, 1988). For tray columns this will yield the holdup above the weir, and the "dead volume" under the weir must be added to get the total holdup.

II. Holdup in Condenser (Accumulator) and Reboiler. We certainly agree that the holdups used in our example columns are much lower than will be observed in a real column. We believe, however, that the use of negligible holdup in the reboiler and condenser is justified by the two reasons we stated. In addition, it should be noted that the RGA is almost unaffected by these holdups.

The response in the condenser (y_D) and the reboiler (x_B) will have an additional lag approximately equal to the residence time in the condenser and reboiler. This lag is undesirable for control purposes and it will vary depending on the holdup. It is, therefore, better in such cases to measure the compositions inside the column, for example, in the overhead vapor line. Alternatively, temperature measurements may be used. As mentioned in the paper they have a dynamic response which is usually weakly affected by the condenser and reboiler holdup.

1. It is not correct, as stated by Riggs, that the holdup in the accumulator and reboiler will introduce a major lag into the dynamic response inside the column. The reason is that the compositions inside the column will start changing immediately if, for example, the liquid flow changes.

2. The seemingly high controller gains for the $(L/D)(V/B)$ configuration are mainly a result of input scaling. For example, if the essentially equivalent $[L/(L + D)][V/(V + B)]$ configuration was used instead, the controller gains for column D would be reduced by a factor 413 (top) and 1,110 (bottom).

III. Disturbance Rejection and Set Points. The fact that the controller settings developed for set point changes were somewhat sluggish for disturbance rejection has nothing to do with the use of linear controllers. The same results would apply to nonlinear controllers. The reason is simply that the disturbances are large and to reject them we require fast response by the controller. For set-point changes these tunings give a response that is faster than what we actually want. In such cases, the feedback controller should be designed for disturbance rejection and a prefilter should be added to the set-points to dampen the response (two-degree-of-freedom controller). Furthermore, for distillation columns the effect of nonlinearity caused by deviations from the set-point may be almost eliminated by using logarithmic compositions as inputs to the controller (Skogestad and Morari, 1988).

IV. Use of Temperatures to Infer Compositions. If only one temperature is used to infer composition, it may be correct that a relative volatility of 1.5 represents a lower limit. We propose, however, that all available measurements along the column should be used when estimating compositions. In this case, it is possible to get a good estimate of the compositions even when the temperature difference across the column is relatively small.

V. Intuitive Arguments for Ratio Configurations. The McCabe-Thiele arguments presented by Riggs are intuitively appealing, but it is incorrect to believe that one can deduct control properties using these arguments. First, the arguments apply to steady state only; and as the DB configuration shows, these may be totally misleading for evaluating control performance. Second, even at steady state, the argument may be misleading (Skogestad and Morari, 1987). For example, it may be shown that the control properties of the LV and $(L/V)V$ configurations are essentially the same, even though the latter has a ratio as one of the inputs. As for Riggs' final comments regarding a boilup rate disturbance, this is exactly the information contained in Table 3 of our paper.

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To the Editor:

In the R&D note titled "Use of High-Gradient Magnetic Fields for the Capture Ferritin" (December 1989, p. 2058), Scott described an experimental investigation of the capture of ferritin particles by high-gradient magnetic separation (HGMS). It was concluded that the efficiency of capture was a function of molecular charge and the standard trajectory theory of the process is inadequate.

A perusal of the text quickly reveals that the author is not familiar with recent and not-so-recent developments in HGMS. This is regrettable, since this interesting problem deserves an expert approach.

It was shown about ten years ago (Collan et al., 1979; Svoboda, 1981) that surface forces may have a substantial effect on the performance of a magnetic separator for particles as large as 20 μm. Further experimental and theoretical research into the role of surface interactions in magnetic separation and flocculation (Svoboda, 1982; Svoboda and Zofka, 1983; Parker et al., 1982; Parsonage, 1984; Svoboda et al., 1986; Svoboda et al., 1987) showed that optimum selectivity and efficiency of magnetic separation of colloidal particles are achieved by control of the surface properties of various components of a suspension. The results imply that the best performance of mag-

netic-separation process can be expected at the pH value at which the surface potentials of the components to be separated have the same sign. However, this pH value should not be too far from the point-of-zero charge of the component to be recovered magnetically.

Moreover, it has been shown (Svoboda and Corrans, 1985) that the desorption of quasicolloidal particles from the matrix of a magnetic separator is strongly dependent on the pH value of the flushing fluid. A total wash-off of particles cannot be achieved without affecting the surface interactions between the adsorbed particles and the matrix.

It is obvious that such an important