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BASE CONTROL FOR THE TENNESSEE EASTMAN PROBLEM

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Abstract—This paper presents an approach to configure a basic PID control system for the recently published Tennessee Eastman testbed process control problem. A multiloop single-input–single-output control architecture is used. The control design approach involves using a combination of steady-state screening tools, followed by dynamic simulation of the most promising candidates. The steady-state tools employed are the relative gain, Niederlinski index, and disturbance analysis. The resulting control system satisfies all of the specifications required for the design. The final PID system is appropriate for adding on top of it an advanced strategy for online optimization and it can be used as a basis for assessing the benefits of advanced control.

INTRODUCTION

Recently, several companies have published testbed problems for use in evaluating advanced process control approaches. The first such problem was published by Shell in 1986 (Prett and Morari, 1986). Since then Amoco (McFarlane *et al.*, 1993), Johnson Wax (Chylla and Haase, 1993) and Tennessee Eastman (Downs and Vogel, 1993) have published problems. This paper focuses on the Tennessee Eastman problem which involves a process with 41 measurements and 12 manipulated variables. A detailed description of this process, including typical disturbances and baseline operating conditions, is given in Downs and Vogel (1993). The process involves the production of two products, G and H, from four reactants: A, C, D, and E. In addition there are two side reactions that occur and an inert B essentially all of which enters with one of the feed streams.

The authors of the Tennessee Eastman problem point out that it is an appropriate testbed for a number of topics. These include: plant-wide control strategy design, multivariable control, optimization, predictive control, estimation/adaptive control, nonlinear control, process diagnostics and education. The purpose of this paper is to present a systematic approach to developing a plant-wide decentralized control system design. This design is based on multiple single-input–single-output (SISO) control loops. The resulting design can form the basis upon which an advanced control scheme, such as predictive control, can be built. In addition it can also be used to compare the advantages of employing other more advanced control approaches.

The systematic approach presented consists of four broad stages, based upon loop speed. In Stage 1 inner cascade loops are closed. In Stage 2 the basic decentralized PID system is designed. Stage 2 design involves all loops except those associated with the process analyzer and product rate. Stage 3 design involves closing the analyzer and product rate loops. Lastly, at Stage 4 higher level controls, such as model predictive control and/or optimization can be added. As one proceeds from Stages 1–3, the speed of the loops involved decreases. The flow loops are the fastest, followed by the level, temperature and pressure loops. The product composition and product flow loops are the slowest. Thus, the plant-wide strategy decomposes the problem into stages based upon relative loop speed. The majority of the paper is concerned with the Stage 2 design. Before discussing Stage 2 design, an overview of the various design stages is given.

OVERVIEW OF CONTROL SYSTEM DESIGN APPROACH

The plant control system can be designed in several stages. An overview of these stages is given below followed by a detailed discussion for Stage 2.

Stage 1

At stage 1 inner cascade loops are closed, based upon experience. As can be seen in Fig. 1 there are eight flow and two temperature cascade loops that can be closed. The flow loops involve the four feed streams, the purge stream, the stripper bottoms, the separator bottoms and the stripper steam flow. The two temperature cascades involve the condenser and reactor cooling streams. Once these loops are closed

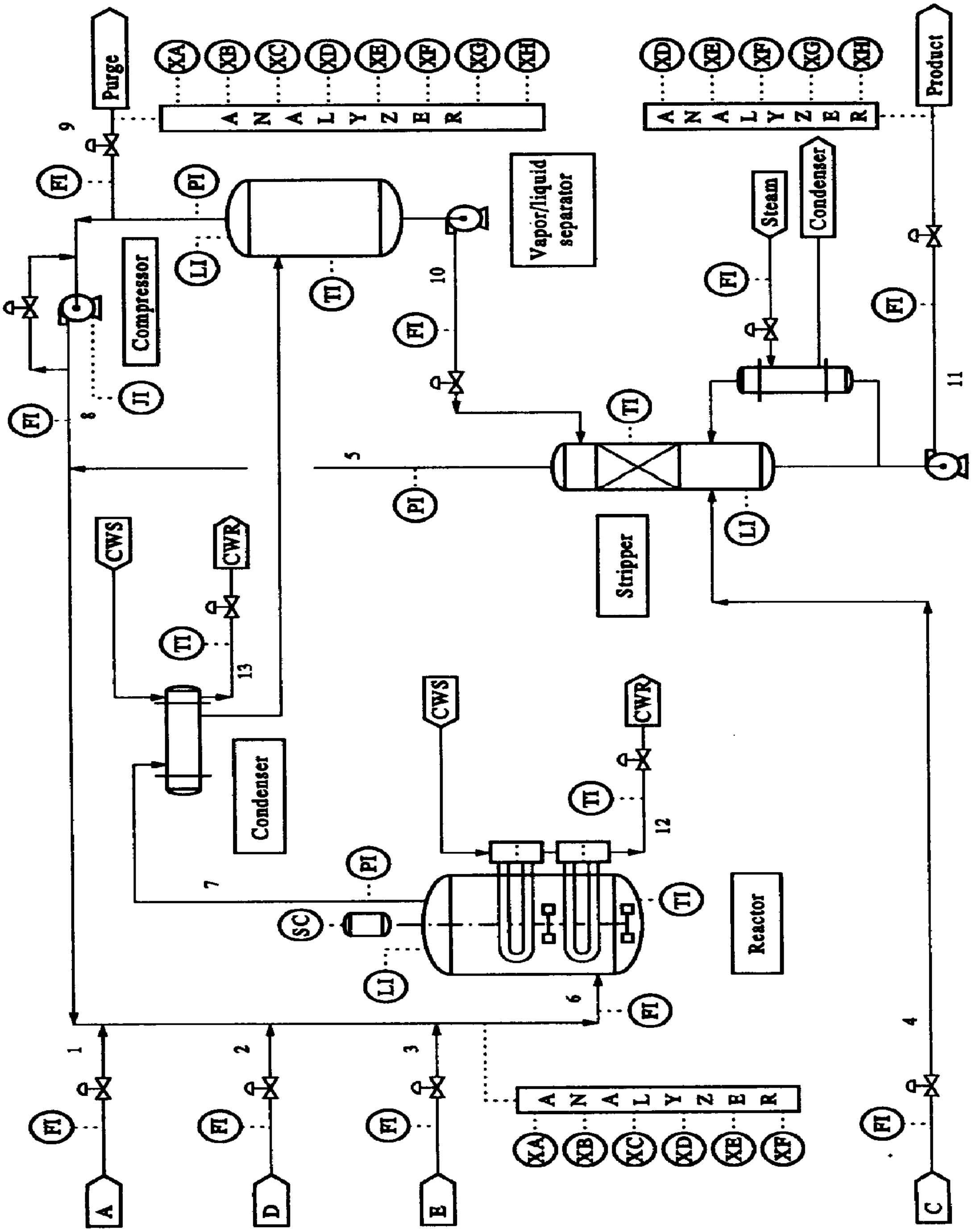
Purge
Product
Compressor
Vapor/liquid separator
Stripper
Condenser
Reactor
Steam
CWS
CWR
ANALYZER
A
D
E
C

Fig. 1

Table 1. Process disturbances

Variable number	Process variable	Type
IDV(1)	A/C feed ratio, B composition constant (Stream 4)	Step
IDV(2)	B composition, A/C ratio constant (Stream 4)	Step
IDV(3)	D feed temperature (Stream 2)	Step
IDV(4)	Reactor cooling water inlet temperature	Step
IDV(5)	Condenser cooling water inlet temperature	Step
IDV(6)	A feed loss (Stream 1)	Step
IDV(7)	C header pressure loss—reduced availability (Stream 4)	Step
IDV(8)	A, B, C feed composition (Stream 4)	Random variation
IDV(9)	D feed temperature (Stream 2)	Random variation
IDV(10)	C feed temperature (Stream 4)	Random variation
IDV(11)	Reactor cooling water inlet	Random variation
IDV(12)	Condenser cooling water inlet temperature	Random variation
IDV(13)	Reaction kinetics	Slow drift
IDV(14)	Reactor cooling water valve	Sticking
IDV(15)	Condenser cooling water valve	Sticking

and the controllers tuned, the flow and temperature indicators, FI and TI, in Fig. 1 can be replaced with controllers, FC and TC. The manipulated variables then become the setpoints of the flow and temperature loops. The speed controller on the reactor agitator, labeled SC, is in effect identical to the setpoints of the inner flow and temperature loops of the cascades. The closure of the 10 cascades eliminates 10 of the 41 process measurements.

One result of closing the cascade loops is that the impact of several of the process disturbances, shown in Table 1, are decreased significantly since they enter the inner loop of the cascades. These disturbances involve inlet cooling water temperature, IDV(4), IDV(5), IDV(11) and IDV(12), the pressure in the C feed line, IDV(7), and the sticking valves, IDV(14) and IDV(15). Treatment of disturbances is discussed later in the paper.

Stage 2

To carry out the next level of control design, it is assumed that the plant must be operational even if the analyzer is not functioning. This assumption is important since analyzers are typically less reliable than the more common temperature, pressure, flow and level sensors. In addition the analyzer loops are typically slower. Stage 3 design, discussed below, involves closing the analyzer loops. If the 19 analyzer measurements are eliminated, then there are $41 - 10 - 19 = 12$ potential variables to be controlled at Stage 2. These variables are listed in Table 2. There are also 12 manipulated variables, which include the 10 cascade setpoints, the agitator speed and the recycle valve around the compressor. These manipulated variables are also listed in Table 2. Figure 2 illustrates the 12×12 problem that must be addressed at Stage 2. It should be emphasized that it may not be possible or desirable to close all 12 loops. The tools that are used here to address this problem are: the relative gain (Bristol, 1966), the Niederlinski Index (1971), linear saturation analysis, nonlinear disturbance and saturation analysis (Vogel and Downs, 1991) and finally dynamic simulation (Vogel and Downs, 1991). Singular value decomposition (Smith *et al.*, 1981) also yields useful information on this problem, but it is not considered here due to space limitations. The approach used is discussed in detail in the following section.

Stage 3

At Stage 3 it is assumed that process levels, flows, temperatures and pressures are controlled as the result of the Stage 2 design. Next, one needs to configure the analyzer loops. To do so the process chemistry and the specifications on production rate and product mix need to be considered. Product mixes of 10/90, 50/50 and 90/10 for the G/H ratio need to be produced and the product flow needs to be adjusted. To develop a control strategy for how to make these changes it is convenient to examine simplified overall material balances for the plant. Although a relative gain analysis could be applied to the simplified material balance results, an approach based on material balance arguments is taken below. Both approaches lead to the same conclusions. After stage 2 the plant can be viewed from an overall perspective as shown in Fig. 3. Although the

Table 2. Manipulated and controlled variables

Manipulated	Controlled
A-feed setpoint	Reactor level
D-feed setpoint	Separator level
E-feed setpoint	Stripper bottom level
C-feed setpoint	Reactor pressure
Purge setpoint	Reactor feed flow
Product setpoint	Reactor temperature
Stripper steam flow setpoint	Compressor power
Separator bottom flow setpoint	Compressor exit flow
Reactor cooling water setpoint	Separator pressure
Condenser cooling water setpoint	Separator temperature
Compressor recycle valve	Stripper pressure
Stirrer speed	Stripper temperature

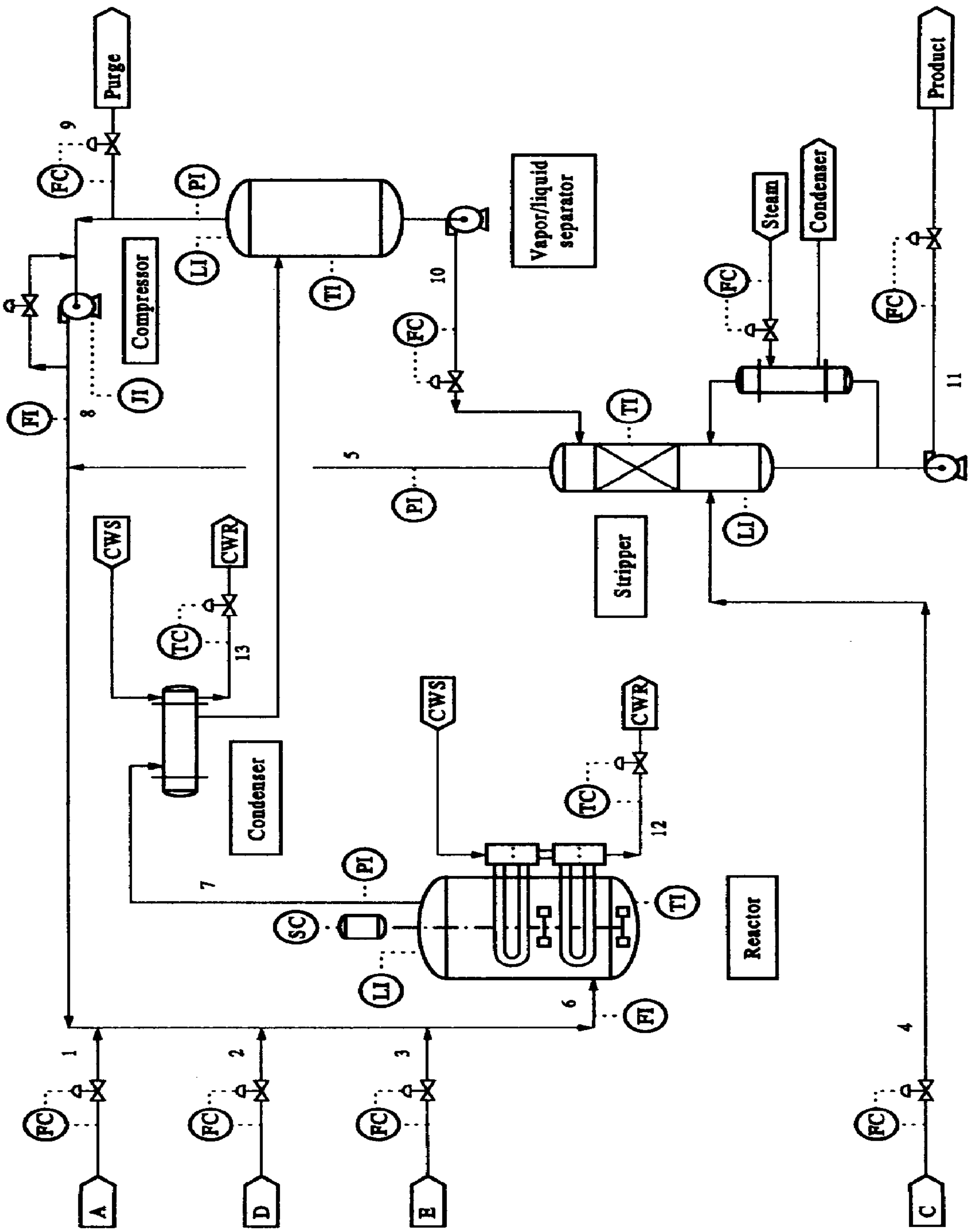
Purge
Product
Vapor/liquid separator
Compressor
Condenser
Stripper
Reactor
Steam
Condenser
CWS
CWR
CWS
CWR
A
D
E
C
FC
PI
LI
TI
TC
SC
JI
FI
1
2
3
4
5
6
7
8
9
10
11
12
13

Fig. 2

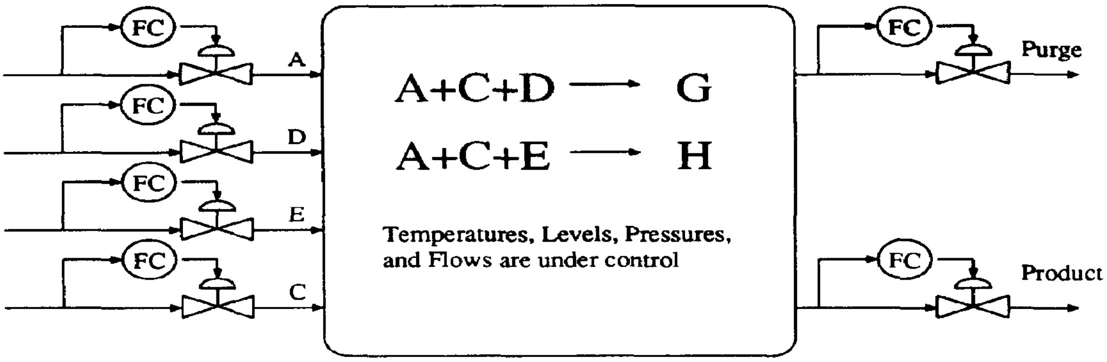

Fig. 3

four feed streams, the product stream and the purge stream are shown, some of these streams may not be available for manipulation if they are assigned to loops during Stage 2 design. This point will be addressed later on.

Since the purge stream is small compared to the product flow, it will be neglected in the simplified material balance. Further, only the two reactions producing the G and H products will be considered. The extent of reaction 1 is taken as e_1 and reaction 2 as e_2. Then, from reaction 1 the amount of G produced is e_1 and the amount of D reacted is also e_1. From reaction 2 the amount of H produced is e_2 and the amount of E reacted is e_2. Since C is required in both reactions 1 and 2, e_1 plus e_2 moles of C react and e_1 plus e_2 moles of product are produced. The amount of A that enters with the C feed is calculated from the base case compositions given in (Downs and Vogel, 1993). The moles of A in the C feed are equal to $(0.485/0.510)\times(e_1+e_2)$. The A feed is assumed to provide the additional moles of A so that a total of e_1+e_2 moles of A enter the system.

Table 3. Simplified overall material balances

Stream	Component	Moles	Mass
G/H = [1408 (kg/h)]/[12,669 (kg/h)] 10/90			
Product	G	22.7 (e1)	1408
	H	166.7 (e2)	12,669
	Total		14,077
C-feed	A	180.1	360.3
	C	189.4	5303
	Total		5663.3
A-feed	A	9.3	18.6
D-feed	D	22.7	736.7
E-feed	E	166.7	7668.2
G/H = (7038 kg/h)/(7038 kg/h) 50/50			
Product	G	113.5 (e1)	7038
	H	92.6 (e2)	7038
	Total		14,076
C-feed	A	196	392
	C	206.1	5770.8
	Total		6162.8
A-feed	A	10.1	20.2
D-feed	D	113.5	3632
E-feed	E	92.6	4259.6
G/H = (12,669 kg/h)/(1408 kg/h) 90/10			
Product	G	204.3 (e1)	12,669
	H	18.5 (e2)	1408
	Total		14,077
C-feed	A	211.9	423.8
	C	222.8	6238.4
	Total		6662.2
A-feed	A	10.9	21.8
D-feed	D	204.3	6537.6
E-feed	E	18.5	851

Table 3 shows the results of this simplified material balance for the three product mix conditions. The results in Table 3 indicate that to control the product mix, the relative amounts of D and E need to be changed substantially. After Stage 2 design at least one of these two manipulated variables must be available to control the product mix. If both the D and E flows are available, then changing the product mix is straightforward. If only one flow is available, then the other must be used to control an inventory variable, i.e. level or pressure, so that it can respond to changes in the free input flow. When the free variable is changed, then the inventory would be affected in such a way that the manipulated variable tied to it changes to achieve the desired product mix. Further, the simplified analysis indicates that the G/H ratio varies directly with the D/E ratio, and thus manipulation of the D/E ratio to control the G/H ratio in the product is suggested. The results in Table 3 also show that for a fixed production rate, the C feed flow does not change appreciably as the product mix changes. This result can be expected since C is required for both products. In order to vary the production rate, the best variable to use is the product rate itself. If product rate cannot be used because it is required for Stage 2 control, then

Table 4. Constraints results from Stage 3; analysis based on simplified material balance

1. Both the C feed and product flow cannot be used in Stage 2 design
2. Both the D feed and E feed cannot be used in Stage 2 design.
3. If the accumulation of B is a problem, the purge setpoint needs to be used to control the composition of B.

the C feed can be used. The D and E feeds are not appropriate for production rate control, since they change appreciably with product mix at a fixed production rate. The A feed is too small for production rate control, and it does not change enough for product mix control. Thus, after Stage 2 design either the product flow or the C flow should be available for production rate control. Also, it may be desirable to ratio feed flow(s), internal flow(s) and compressor power to the setpoint of product flow.

The last point to consider is the elimination of the inert component B. This component enters with the C feed and it essentially goes out in the purge stream. Since B is a gas, its accumulation could cause problems, including a rise in pressure. If B purge must be used to control the amount of B in the system. The constraints on the manipulated vari- accumulates, then the purge stream needs to be manipulated to control the amount of B in the plant. A disturbance analysis can be used to decide if the ables resulting from the Stage 3 analysis are summarized in Table 4.

Stage 4

At stage 4 higher level controls are added on top of the basic plant control system. These higher level controls include: steady state control (Piovoso, 1992), steady state optimization (Forbes *et al.*, 1992) and model predictive control (Cutler and Ramaker, 1979). Stage 4 controls are beyond the scope of this paper.

DETAILED DESCRIPTION OF STAGE 2 DESIGN

There are a number of steps that need to be carried out to complete the Stage 2 design. These include: Step 1 close the level loops; Step 2 assess interaction, stability, and saturation problems; Step 3 carry out a steady-state disturbance analysis; and Step 4 tune and test candidate control systems via dynamic simulation. Each step is discussed separately below.

Step 1

Of all the controlled variables in a plant, levels are probably the most important. One cannot afford to

	A Feed Setpt	D Feed Setpt	C Feed Setpt	Purge Setpt	Steam Setpt	Rea Cl Setpt	Sepa Cl Setpt	Recy Valve	Agit Speed
Feed Rea y(1)	-8.2062	-0.0030	3.2054	-7.6820	0.0009	-0.6652	0.2346	-0.0360	0.1291
Rea Temp y(2)	6.4830	0.0022	1.6619	0.8968	0.0005	1.1106	0.0281	0.0141	-0.2149
Rea Pres y(3)	-3103.2488	-0.2796	287.0500	-965.7145	0.0477	-31.4794	11.9680	7.7756	6.1135
Sepa Temp y(4)	67.8248	0.0109	-7.4203	21.0053	-0.0020	2.2502	0.2853	-0.1954	-0.4359
Stri Temp y(5)	46.0956	0.0095	-6.4259	20.1120	0.0368	1.8663	0.4130	-0.0380	-0.3615
Recy Flow y(6)	-11.2025	-0.0028	1.6500	-7.0837	0.0004	-0.6510	0.2640	-0.0187	0.1256
Comp Power y(7)	126.7784	-0.0159	13.2484	-27.4335	0.0090	-4.5126	2.5222	2.9658	0.8732
Sepa Pres y(8)	-3053.8689	-0.2744	281.1390	-948.6965	0.0465	-30.9414	11.8017	8.5438	6.0093
Stri Pres y(9)	-3377.3005	-0.3080	319.9876	-1059.6489	0.0561	-34.4602	12.8932	3.4986	6.6962

Fig. 4

have a vessel overflow or run dry. Further, level loops must be closed in order to calculate steady-state gains. Otherwise, step changes in manipulated variables produce ramp-like responses which result in valve saturation or constraint violation. At Stage 2 there are 3 levels that need to be controlled: the separator level, the stripper bottoms level and the reactor level. The logical choice for the separator level is its bottoms flow setpoint. For the stripper bottoms level, either the product flow setpoint, or the steam flow setpoint can be used. Since there are constraints on how fast the product flow can be manipulated, if it is used then a loosley tuned averaging level loop should be employed. For the reactor level, tight control is required and the cooling water setpoint or the E feed setpoint are simple possibilities. Ricker *et al.* (1993) discuss a more complicated level control strategy in which recycle rate and condenser cooling are used. This more complex strategy may have an advantage for plant operation over the complete 10/90, 50/50 and 90/10 product mix. Using the E feed for level control means that the E feed can only be set at some percentage of its maximum, e.g. 90%, otherwise level control will be lost due to valve saturation. For the 10/90 G/H case limiting the E feed to 90% will also limit the maximum production rate.

This paper addresses control around the 50/50 setpoint and the various control objectives given in Downs and Vogel (1993) as tests for a control system design. As discussed in Downs and Vogel (1993), feed streams A and D have constraints on their rate of change and thus they can be ruled out, since fast level control cannot be achieved using them. Once the level loops are assigned, steady-state gains for the resulting 9×9 process can be calculated using the procedure given in McAvoy (1983). Small positive and negative changes are made in the manipulated variables and the resulting changes in the controlled variables are averaged. Since there are four possible level configurations, there are four 9×9 systems that need to be analyzed. Detailed results for one of these systems are presented below along with a summary of results for the other three cases. The specific case considered involves using the E feed to control reactor level and the product flow to control the stripper level. The gain matrix for this system is shown in Fig. 4.

Table 5

Scheme 1

	A-feed flow setpoint	Steam flow setpoint	Reactor cooling setpoint	Comp. recycle valve
Reactor temperature	−0.036	−0.019	**1.030**	0.025
Reactor pressure	**0.921**	0.015	−0.045	0.108
Strip temperature	0.012	**1.007**	−0.023	0.004
Comp power	0.102	−0.003	0.037	**0.863**

Scheme 2

	A-feed flow setpoint	Steam flow setpoint	Reactor cooling setpoint	Condenser cooling setpoint	Comp recycle valve
Reactor temperature	−0.009	−0.037	**0.981**	0.074	−0.009
Reactor pressure	**0.946**	0.018	−0.047	−0.060	0.143
Strip temperature	0.002	**1.074**	−0.019	−0.057	−0.000
Comp power	0.112	−0.008	0.041	0.118	**0.738**
Feed reactor	−0.051	−0.047	0.045	**0.925**	0.128

Scheme 3

	A-feed flow setpoint	Steam flow setpoint	Reactor cooling setpoint	Condenser cooling setpoint	Comp recycle valve
Reactor temperature	−0.062	−0.002	**1.079**	−0.073	0.058
Reactor pressure	**0.744**	−0.008	−0.025	0.437	−0.148
Strip temperature	0.018	**0.969**	−0.026	0.032	0.007
Comp power	0.110	−0.007	0.040	0.091	**0.766**
Separator temperature	0.191	0.048	−0.069	**0.513**	0.317

Scheme 4

	A-feed flow setpoint	Steam flow setpoint	Reactor cooling setpoint	Condenser cooling setpoint	Comp recycle valve
Reactor temperature	−0.014	−0.034	**0.989**	0.062	−0.003
Reactor pressure	**0.962**	0.020	−0.049	−0.101	0.167
Strip temperature	0.007	**1.039**	−0.021	−0.028	0.002
Comp power	0.109	−0.007	0.040	0.086	**0.771**
Recycle flow	−0.065	−0.020	0.041	**0.981**	0.062

must be controlled results in a smaller number of RGA cases to be examined, but it does not change the basic methodology. Also, it is possible that if too many variables are specified as definitely having to be controlled, one may not get to a solution. In this case the specification on variables that definitely have to be controlled has to be relaxed. By specifying that 4 variables must be controlled, the number of RGAs that must be considered is relatively small. There are 3 6×6 cases, 18 5×5 cases and 15 4×4 cases, giving a total of 36 cases. In addition to using the Niederlinski Index to rule out unstable pairings, physical arguments can be used as well. For example, one would not pair the D-feed flow with the stripper temperature due to how far apart physically these variables are. Similarly, the use of the very small purge flow to control a much larger flow, for example the feed to the reactor, can be ruled out since valve saturation is likely during transients. In the results given below, only RGA pairings between 0.5 and 4.0 are considered acceptable. Lastly, a linear valve saturation analysis (Skogestad and Wolff, 1992) can be carried out based on the process steady-state gains. Schemes in which valves saturate are ruled out.

Of the 36 cases, only 4 passed all the screening tests. In all 4 schemes reactor pressure is paired with A-feed flow, reactor temperature with reactor cooling temperature, stripper temperature with steam flow and compressor power with the recycle valve around the compressor. Table 5 shows the RGAs for the four candidate control systems. The next step in the analysis is to compare the steady-state ability of these schemes to reject disturbances.

Step 3

The ultimate goal of the final control system is to keep both the product flow and composition as close to the setpoints as possible in spite of upsets. In Steps 1 and 2 above, product compositions and flows are not considered explicitly. Downs and Vogel (1991) have presented an approach, based upon a paper by Luyben (1975), through which the ability of a plant's basic PID control system to reject disturbances on the more important product variables can be assessed. This approach is used here to screen the 4 schemes which result from Step 2 and then select candidate schemes for dynamic simulation.

To carry out Downs and Vogel's approach, one considers each significant upset one at a time. Table 1 lists these upsets. As mentioned earlier, closing the cascade loops effectively compensates for upsets IDV(4), IDV(5), IDV(7), IDV(11), IDV(12), IDV(14) and IDV(15). Further, it was found that upset IDV(3) was very easy to control and it causes no problems. Thus, at this step only IDV(1), IDV(2) and IDV(6) need to be examined. Upset IDV(6) is discussed separately below. To analyze for IDV(1), a plot of the steady-state product flow and composition, shown in Fig. 5, is made as a function of the size of the disturbance. To make this plot one has to solve the nonlinear steady state process model. What one desires in the basic PID control system is a scheme that inherently has the ability to reject disturbances without the use of the analyzer. If such performance can be achieved then the task of the analyzer control loops will be that much easier. Figure 5 shows that in the face of the IDV(1) upset, all four schemes perform about the same. A perfect control scheme would keep all product variables exactly at their setpoints. A similar plot can be made for IDV(2) and it is shown in Fig. 6. The fact that the plots for the four schemes end at IDV(2) = 0.30 is indicative of the fact that if the purge flow is held constant then the B material balance cannot be met, and the steady-state equations have no solution. Figure 7 shows the same plot as Fig. 6, but with the purge used to control the composition of B in the purge stream. Now, the full effect of upset IDV(2) can be handled. It can be concluded that to handle IDV(2), the purge should be used to control the composition of B in the purge stream. As Fig. 7 shows, there is little difference between the four candidate schemes. In carrying out this disturbance analysis, one can also assess potential valve saturation problems using the complete nonlinear model, as compared to using linear approaches (Skogestad and Wolff, 1992). For all four schemes all valves are safely within their saturation limits.

Next IDV(6) is considered. IDV (6) involves the loss of the A feed stream which is manipulated to control pressure. This upset is similar to IDV(2) in that it results in an imbalance of gaseous components entering and leaving the plant. The excess gas can only be eliminated through the purge stream, or by cutting back on the feed to the plant. In the case of IDV(2) additional B has to be removed. For normal plant operation the inputs of A and C are roughly equal. When IDV(6) occurs, the loss of A means that excess C must be purged from the plant, otherwise pressure will continue to rise. Purging the excess C can be accomplished by switching the pressure controller to the purge stream when the A feed is lost. An examination of Fig. 4 shows that after the A feed, the purge has the most important effect on pressure. Using the purge to control pressure gives rise to the RGAs shown in Fig. 8, and these are acceptable. Next, a linear saturation analysis (Skogestad and Wolff, 1992) is carried out and it shows that the purge valve will saturate when the A feed is lost. The purge stream simply cannot handle all of the excess gas and inerts that need to be eliminated. One possible solution is to lower the C feed to the plant since it is this stream that brings in the excess gas as well as the inert B. However, in the 4 schemes under consideration, the C feed is used for production rate control. Thus, this approach to IDV(6) requires that production be cut back.

To verify these conclusions, Fig. 9 was developed for steady-state analysis of IDV(6). For each of the four schemes, the purge was used for pressure control. When the purge valve reached 90% of its full open value, then the production rate was lowered. In calculating steady state conditions, it was found useful to ratio the compressor power, reactor feed (scheme 4), and compressor exit flow (scheme 2) to the product flow set point. These same ratios are used in the dynamic simulations discussed below. Before the product flow set point is ratioed, it is sent through a 2 h time lag to avoid sudden step changes from affecting the ratioed variables. The value 90% for the purge valve is arbitrary, but is chosen so that even after the A feed loss the purge can still have some rangeability for control. Figure 9 shows that when IDV(6) exceeds 0.7, a steady-state solution cannot be found for scheme 3. Similarly, for scheme 1 a steady-state solution cannot be found when IDV(6) exceeds 0.87. In both cases with the purge fixed at 90%, too much excess C remains in the system for steady-state to be achieved. Thus, schemes 1 and 3 are eliminated. The other two schemes are very close in their steady state ability in so far as IDV(6) is concerned. For both schemes 2 and 4, the condenser cooling temperature is lowered in the face of IDV(6). This lowering of temperature allows more liquid to flow out with the product, and therefore less gas builds up. Clearly, for scheme 1 one could consider lowering the condenser exit cooling water temperature setpoint when IDV(6) occurs. Alternatively, for scheme 3 one could consider lowering the separator temperature setpoint when IDV(6) occurs. Neither of these two alternatives is considered here.

The next step in the analysis involves tuning the various control loops and carrying out dynamic simulation. Before discussing this step, the results of carrying out Steps 1–3 for the other level control configurations will be summarized. First, when steam flow is used to control the stripper level, then product flow is available for other uses. However, because of the restrictions on the rate of change of product flow, it would have to be manipulated very slowly. Thus, it would not be effective as a manipulated variable. Simply leaving product flow out of the basic PID system resulted in configurations that were inferior in terms of their ability to reject disturbances to those when product flow controlled stripper level. Similarly, the use of reactor coolant to control reactor level gave very poor results. Not only did large RGAs result, but control valve saturation problems resulted as well. No viable pairings were found when such a level scheme was examined.

Step 4

The last step in the analysis involves tuning the various control loops and dynamic simulation to assess the system's response to disturbances. In tuning loops, the same order that is used in Steps 1–3 is used. First, the inner loops of the cascades are tuned. Then the level loops are tuned. Next, the remaining, noncomposition/production rate loops are tuned. Finally the composition and production rate loops are tuned. Initial loop tuning was carried out with no noise in the simulation. Then, noise was added and only flow loops and the two temperature coolant loops were detuned. For both the stripper level—product flow and reactor pressure—A feed loops the controllers are tuned to give an averaging type control response (McDonald *et al.*, 1986) to meet the constraints on how fast the two manipulated variables can move. Also an averaging pressure control approach is used for the purge flow—pressure loop for the IDV(6) upset. The production rate—C feed loop and the product mix—D/E ratio loop are also tuned to respond slowly enough that the constraints on the rate of change in the various flows are satisfied. Finally, the temperature setpoint for the stripper control is used to control the E mole fraction in the product in a double cascade arrangement. After tuning and simulation, it was found that the two remaining schemes gave almost equivalent performance. In the results given below, scheme 4 is used. In all cases PI controllers are used and the resulting controller parameters are given in Table 6. The final plant control scheme is shown in Fig. 10.

One last point can be noted. When IDV(6) occurs, the production rate setpoint is stepped down by 23.8%, as indicated by the steady-state analysis shown in Fig. 9. During the transient produced by IDV(6), the purge valve saturated for a period of time. However, at steady-state the valve came back to 90% open. For the purge flow pressure loop a controller gain of −0.00352 kscfm/kPa was used with a reset time of 100 min.

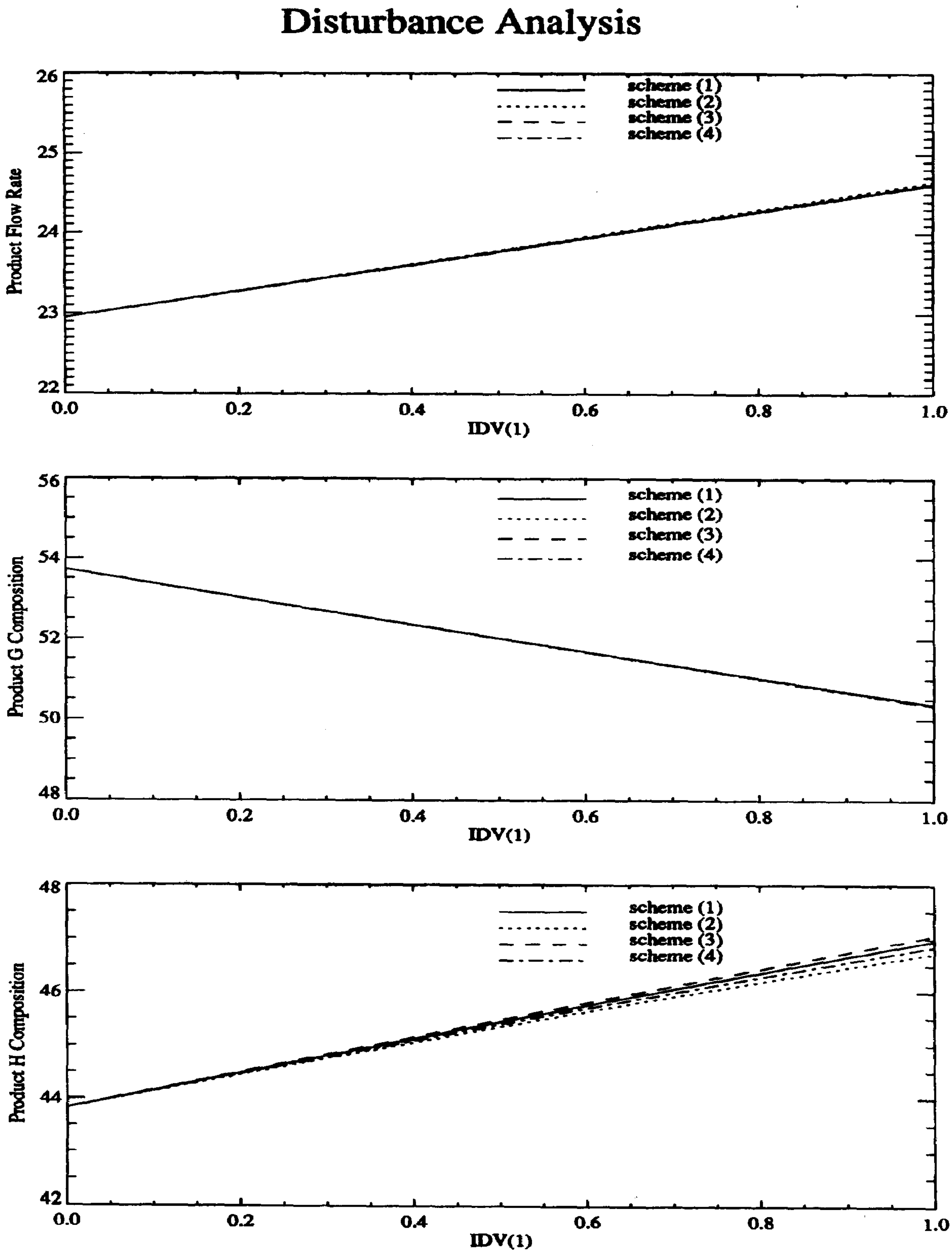
Disturbance Analysis
scheme (1)
scheme (2)
scheme (3)
scheme (4)
Product Flow Rate
26
25
24
23
22
0.0
0.2
0.4
0.6
0.8
1.0
IDV(1)
Product G Composition
56
54
52
50
48
IDV(1)
Product H Composition
48
46
44
42
IDV(1)

Fig. 5

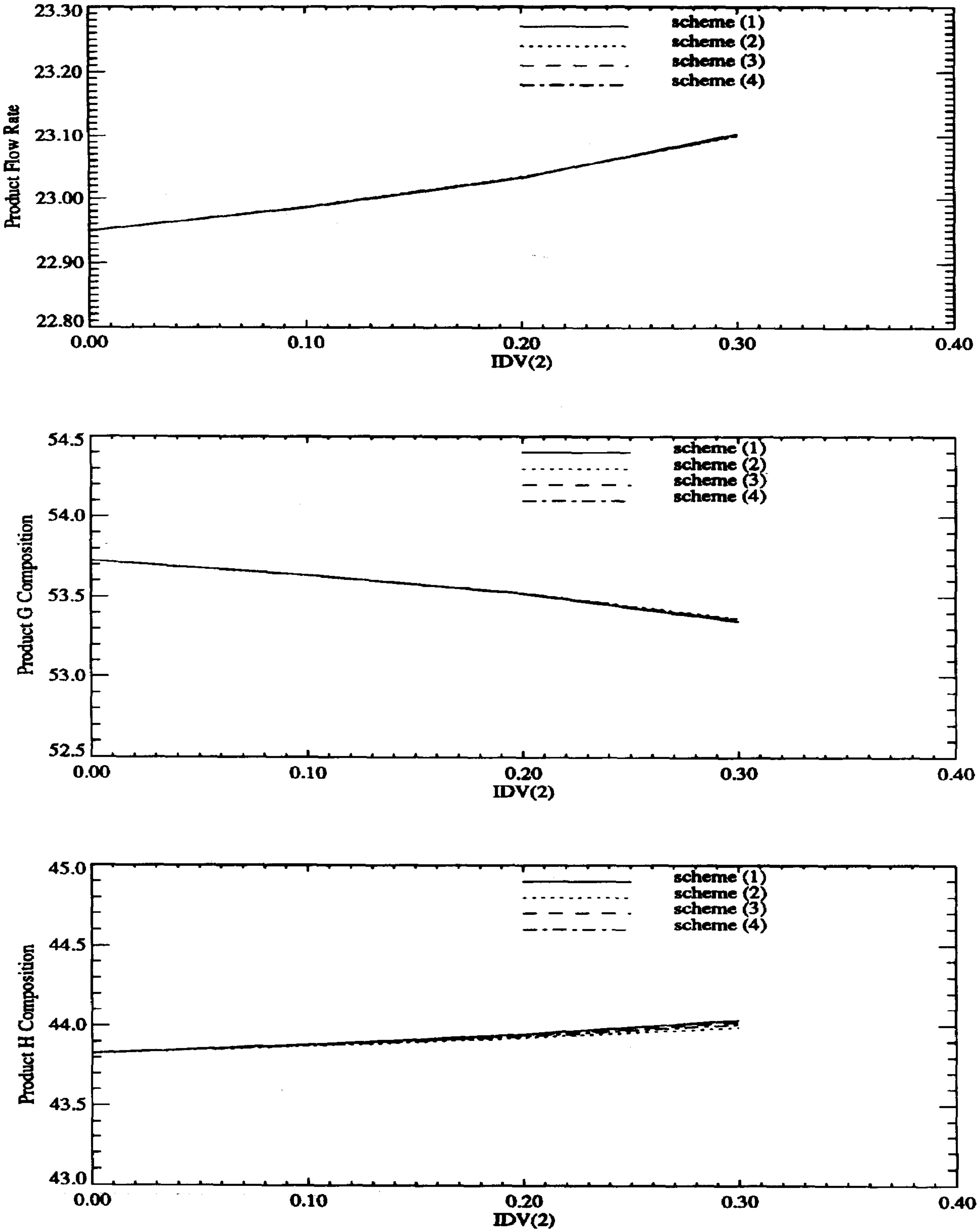

Fig. 6. Disturbance analysis (purge B composition not controlled).

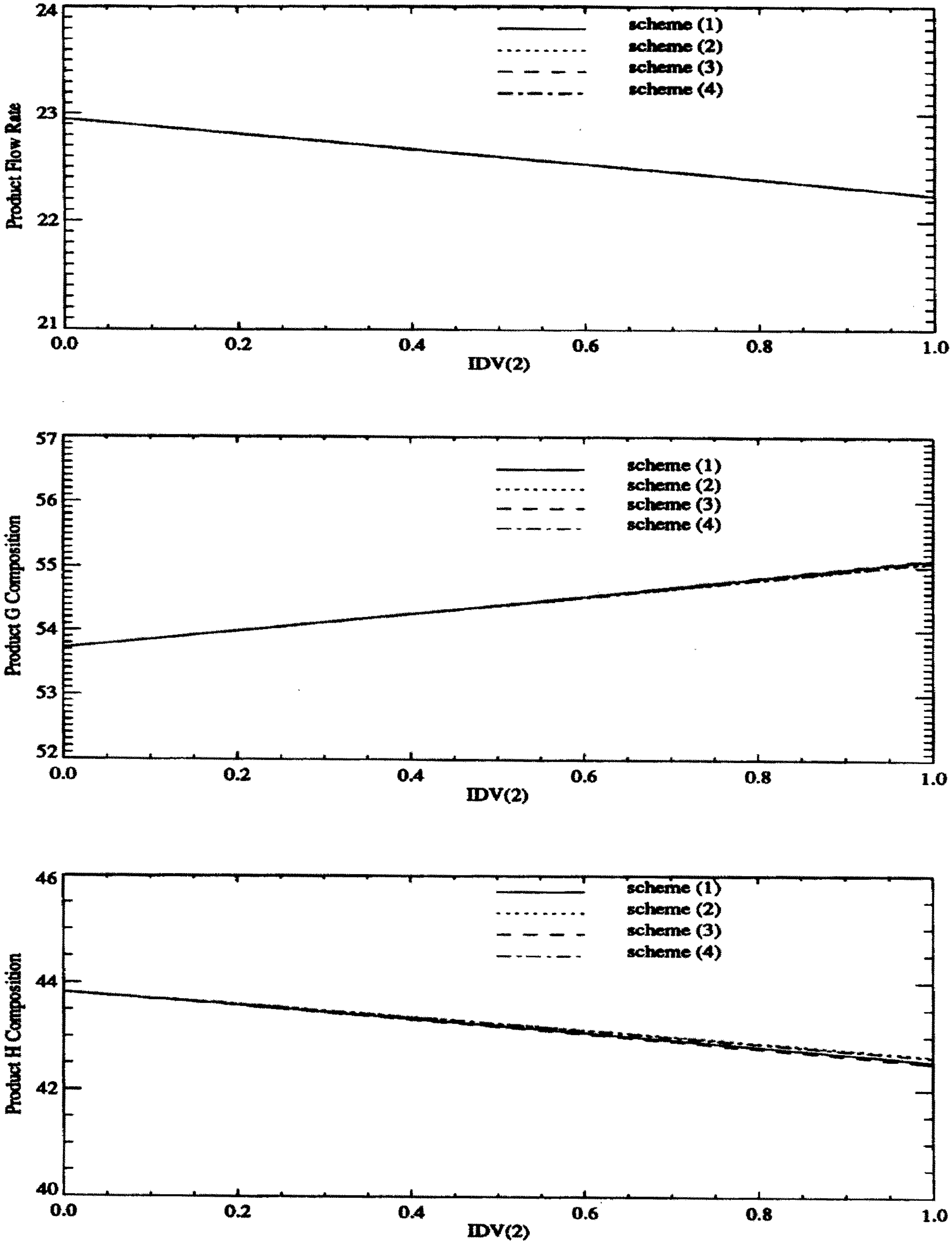

Fig. 7. Disturbance analysis (purge B composition not controlled).

SCHEME(1)					
	purge flow setpoint	steam flow setpoint	rea. cooling setpoint	comp recycle valve	
rea. temp.	-0.019	-0.017	1.017	0.018	
rea. presu.	1.083	0.027	-0.025	-0.085	
strip temp	0.020	0.999	-0.022	0.003	
comp pow	-0.084	-0.009	0.030	1.063	

SCHEME(2)					
	purge flow setpoint	steam flow setpoint	rea. cooling setpoint	cond. cooling setpoint	comp recycle valve
rea. temp.	-0.006	-0.035	0.980	0.069	-0.008
rea. presu.	1.323	0.056	-0.043	-0.504	0.169
strip temp	0.004	1.069	-0.019	-0.054	-0.000
comp pow	-0.109	-0.029	0.039	0.389	0.710
feed react	-0.213	-0.061	0.043	1.101	0.131

Fig. 8

SCHEME(3)					
	purge flow setpoint	steam flow setpoint	rea. cooling setpoint	cond. cooling setpoint	comp recycle valve
rea. temp.	-0.032	0.002	1.056	-0.072	0.046
rea. presu.	0.871	0.001	-0.009	0.445	-0.308
strip temp	0.029	0.958	-0.024	0.031	0.006
comp pow	-0.089	-0.013	0.032	0.085	0.986
sepa temp	0.222	0.052	-0.055	0.511	0.270

SCHEME(4)					
	purge flow setpoint	steam flow setpoint	rea. cooling setpoint	cond. cooling setpoint	comp recycle valve
rea. temp.	-0.008	-0.032	0.987	0.057	-0.004
rea. presu.	1.269	0.050	-0.039	-0.392	0.112
strip temp	0.013	1.028	-0.021	-0.023	0.002
comp pow	-0.100	-0.022	0.036	0.257	0.829
recyc flow	-0.174	-0.024	0.037	1.101	0.060

Fig. 8—*Continued*

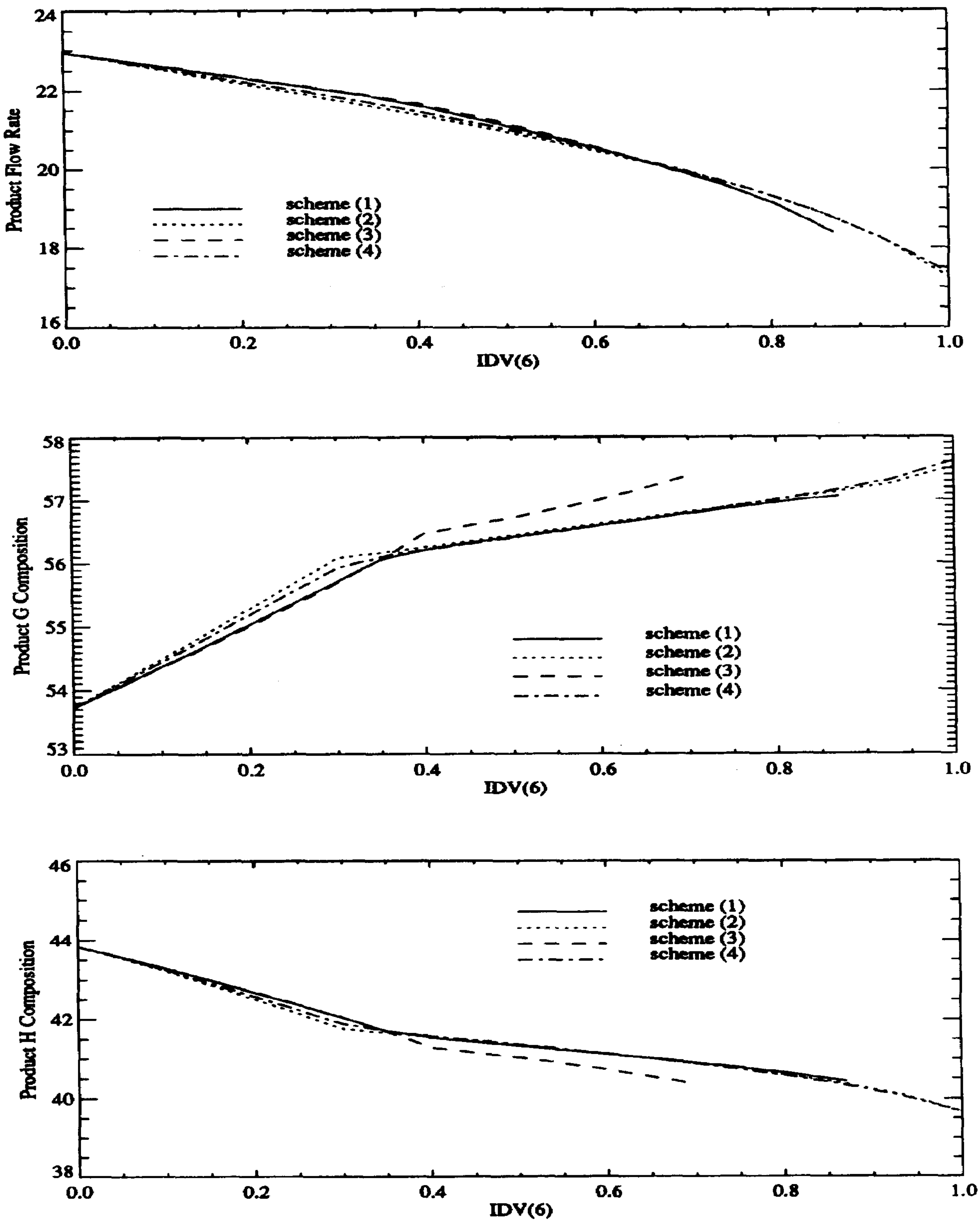

Fig. 9. Disturbance analysis (compressor power, recycle and feed to reactor ratio to product setpoint).

Table 6a

PI parameters (cascade inner loops)				
	A-feed flow	D-feed flow	E-feed flow	C-feed flow
P	200 (%/kscmh)	0.002 (%/kg/h)	0.002 (%/kg/h)	0.1 (%/kscmh)
T_i (min)	0.1	0.3	0.3	0.3
	Purge flow	Separator under flow	Strip under flow	Strip steam flow
P	100 (%/kscmh)	0.3 (%/m³/h)	0.5 (%/m³/h)	0.03 (%/kg/h)
T_i (min)	0.3	0.3	0.3	0.3
	Reactor cooling temperature	Separator cooling temperature		
P	−10 (%/°C)	−10 (%/°C)		
T_i (min)	1	1		

Table 6b

PI parameters (control loops)				
	Reactor temperature	Reactor pressure	Stripper temperature	Compressor power
P	1.0	−0.0032 (kscmh/kPa)	10.0 (kg/h/°C)	0.08 (%/kW)
T_i (min)	50	300	10	20
	Reactor level	Separator level	Stripper level	Purge B composition
P	500 (kg/h/%)	−2.5 (m³/h/%)	−0.5 (m³/h/%)	−0.03 (kscmh/%)
T_i (min)	200	200	300	100
	Product flow	Product G/H ratio	Recycle flow	Product E composition
P	0.08 (kscmh/m³/h)	0.05	1.5 (°C/kscmh)	−0.5 (°C/%)
T_i (min)	45	40	50	100

Step 2

At Step 2 manipulated and/or controlled variables are eliminated based upon operating considerations and examination of the process gain matrix. For the gain matrix given in Fig. 4, product flow and E feed are used to control levels. Following the discussion given under Stage 3 above, D/E should be used for control of product mix. Also, since the product flow is used for level control, the C feed must be used for production rate control. Thus, these two manipulated variables can be eliminated. An examination of the gain matrix in Fig. 4 shows very strong correlation between the agitator speed and the reactor cooling temperature setpoint. Column 9 is almost a constant multiple of column 6. Further all of the pressure measurements are strongly correlated. Rows 3, 8 and 9 are almost constant multiples of one another. Thus, it will be extremely difficult to manipulate agitator speed and reactor coding independently and therefore agitator speed is dropped. It will also be very difficult to control all three pressures and therefore only the reactor pressure is retained. Clearly, a singular value decomposition analysis (Smith *et al.*, 1981) could be used to get the same insights. Dropping agitator speed and the separator and stripper pressures results in a 7×6 problem.

Next, a relative gain analysis (Bristol, 1966) is carried out to determine loop pairings. The stability of the resulting loops is checked using the Niederlinski Index (Niederlinski, 1971). To carry out an RGA analysis on the 7×6 problem, one of the controlled variables has to be eliminated. If all 7 controlled variables were eliminated one at a time, then there would be 7 (6×6) RGAs to consider. However, in any realistic control system, some of the process variables must be under control. In the present case these variables would include reactor temperature and pressure. In addition, since stripper temperature reflects product composition, it will also have to be controlled. Finally, it is decided to control compressor work. Thus, the controlled variables that will be eliminated one at a time are: reactor feed flow, compressor exit flow and separator temperature. Deciding that certain variables

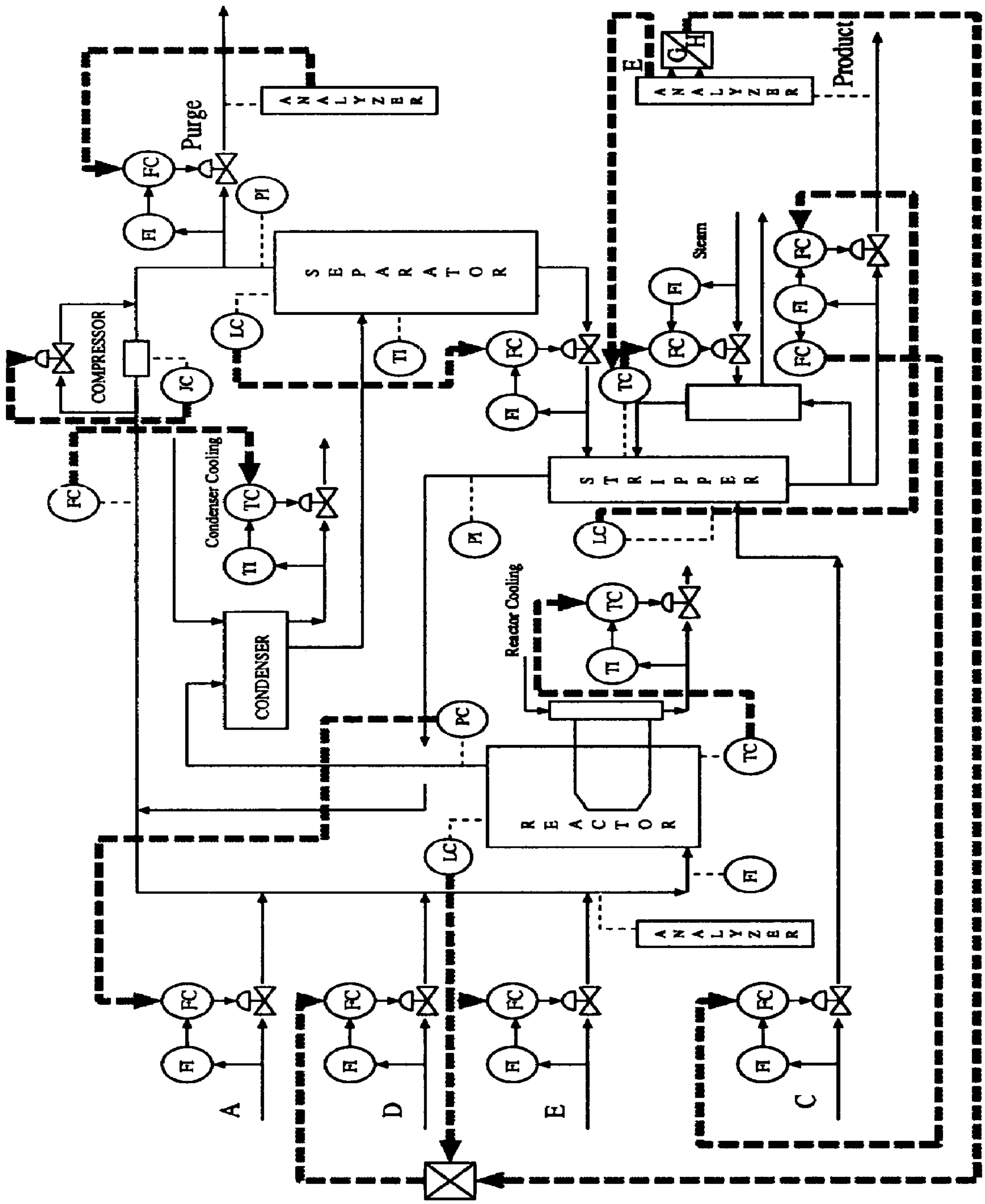

Fig. 10. Base control system of the Tennessee Eastman process.

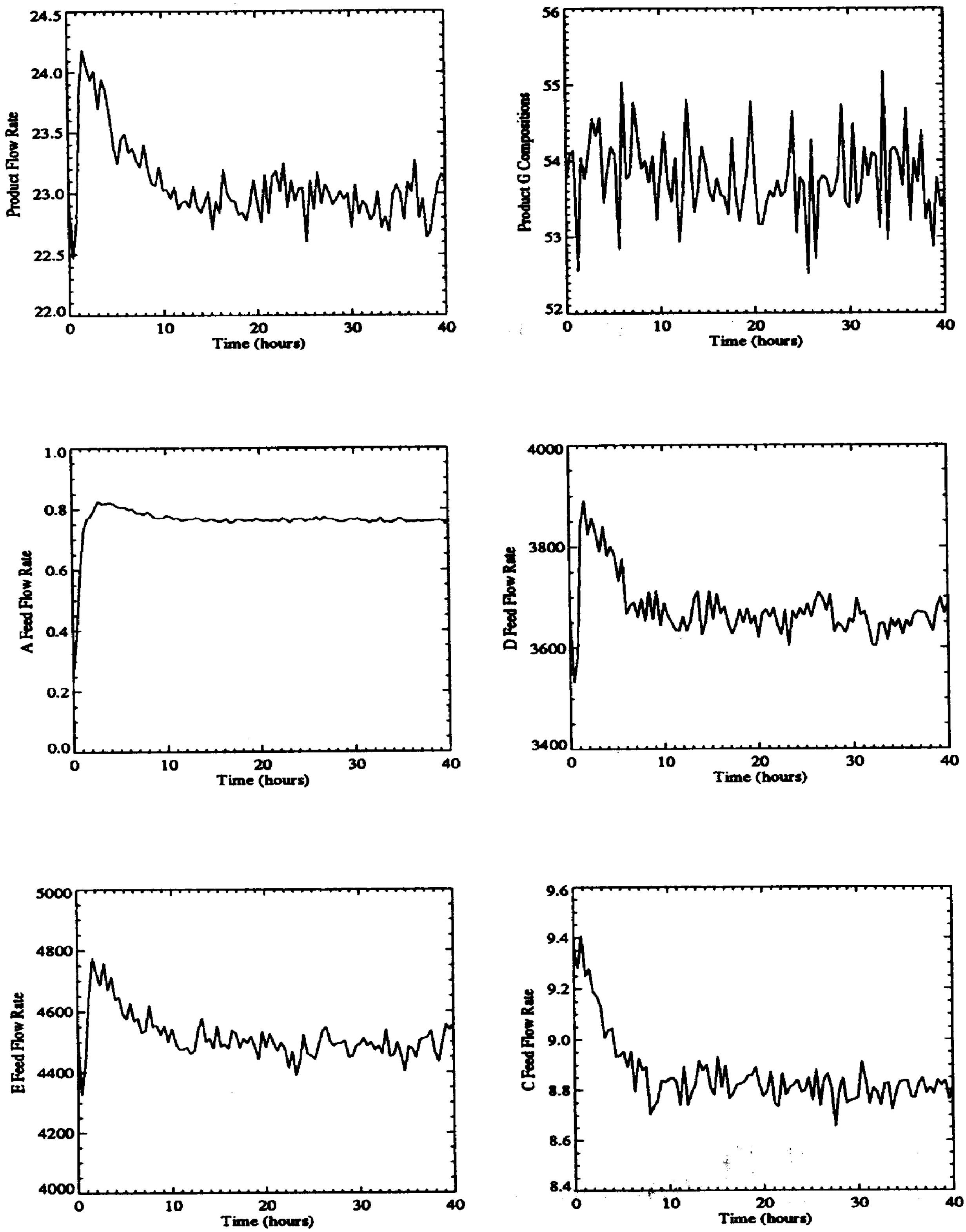
Product Flow Rate
24.5
24.0
23.5
23.0
22.5
22.0
Product G Compositions
56
55
54
53
52
A Feed Flow Rate
1.0
0.8
0.6
0.4
0.2
0.0
D Feed Flow Rate
4000
3800
3600
3400
E Feed Flow Rate
5000
4800
4600
4400
4200
4000
C Feed Flow Rate
9.6
9.4
9.2
9.0
8.8
8.6
8.4
0
10
20
30
40
Time (hours)

Fig. 11a. IDV(1).

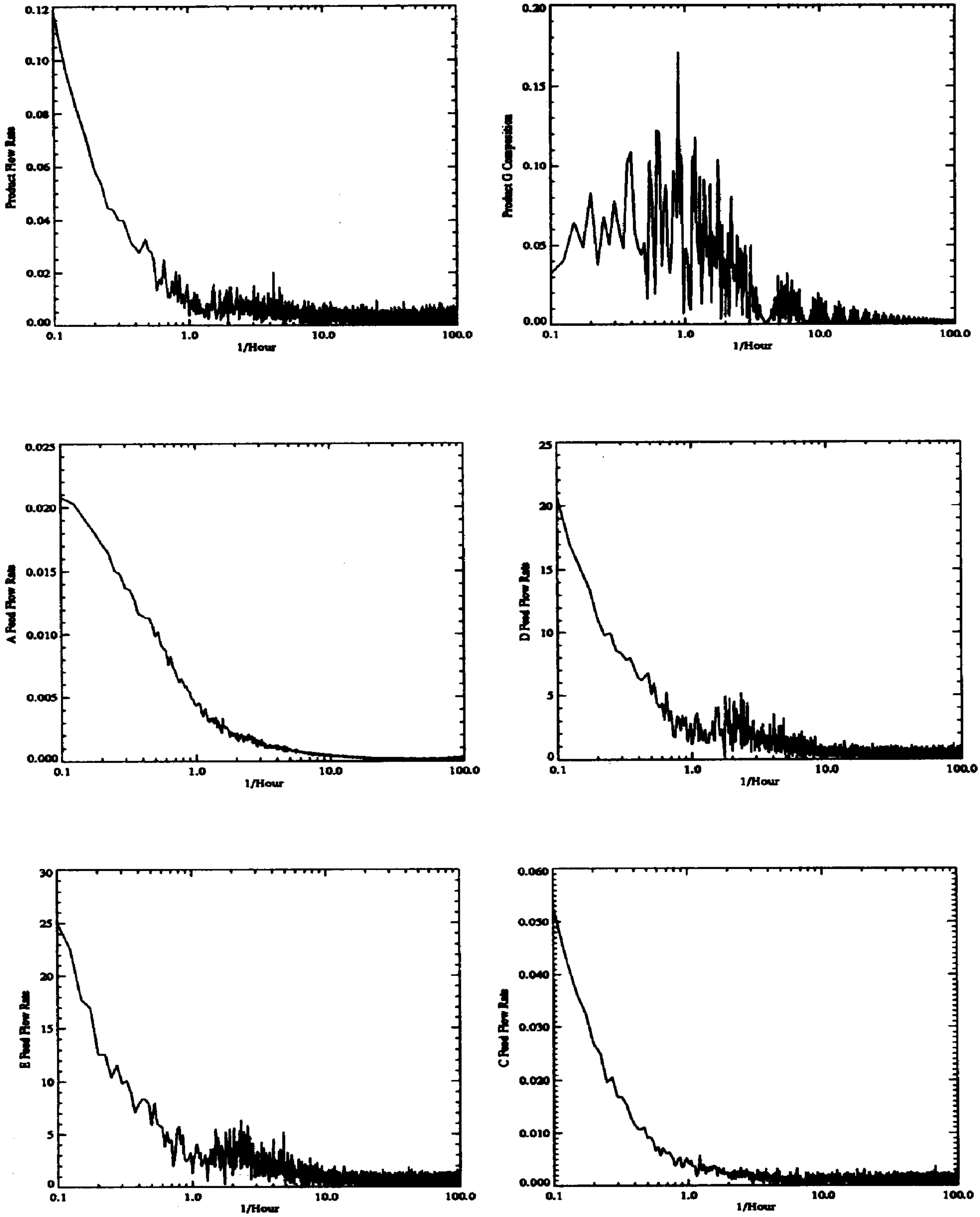

Fig. 11b. IDV(1) (Fourier coefficients).

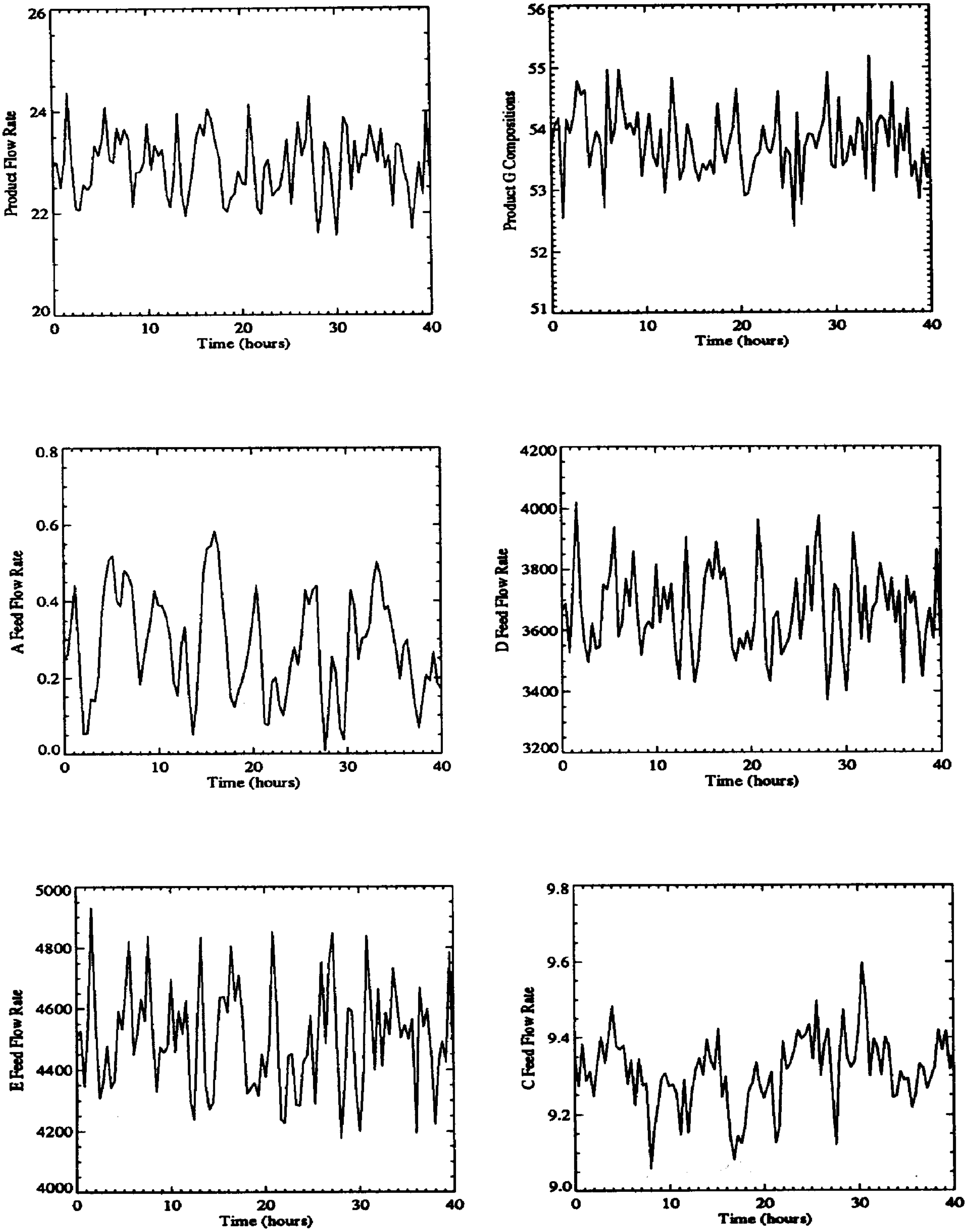
Product Flow Rate
Time (hours)
Product G Compositions
Time (hours)
A Feed Flow Rate
Time (hours)
D Feed Flow Rate
Time (hours)
E Feed Flow Rate
Time (hours)
C Feed Flow Rate
Time (hours)

Fig. 12a. IDV(8).

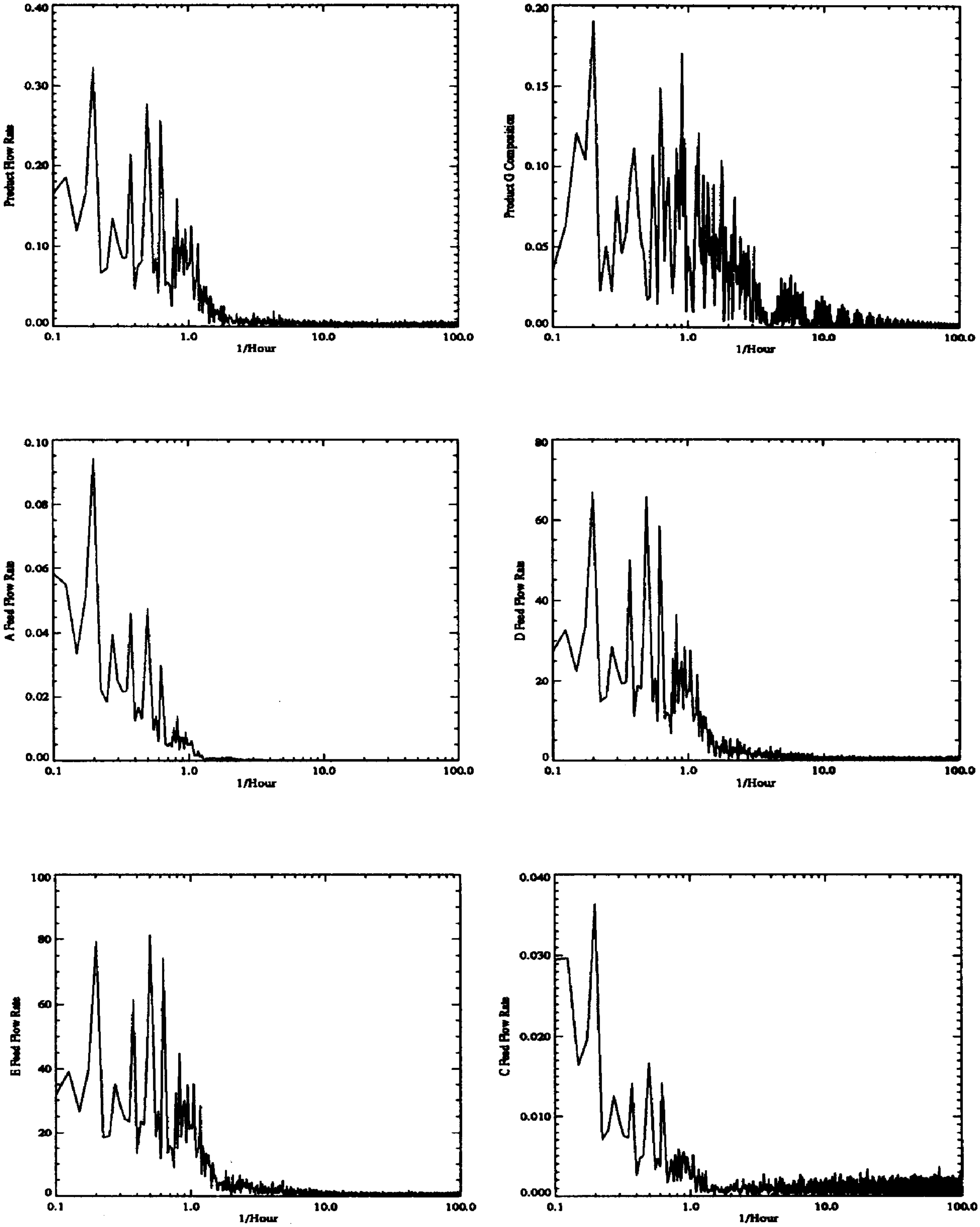

Fig. 12b. IDV(8) (Fourier coefficients).

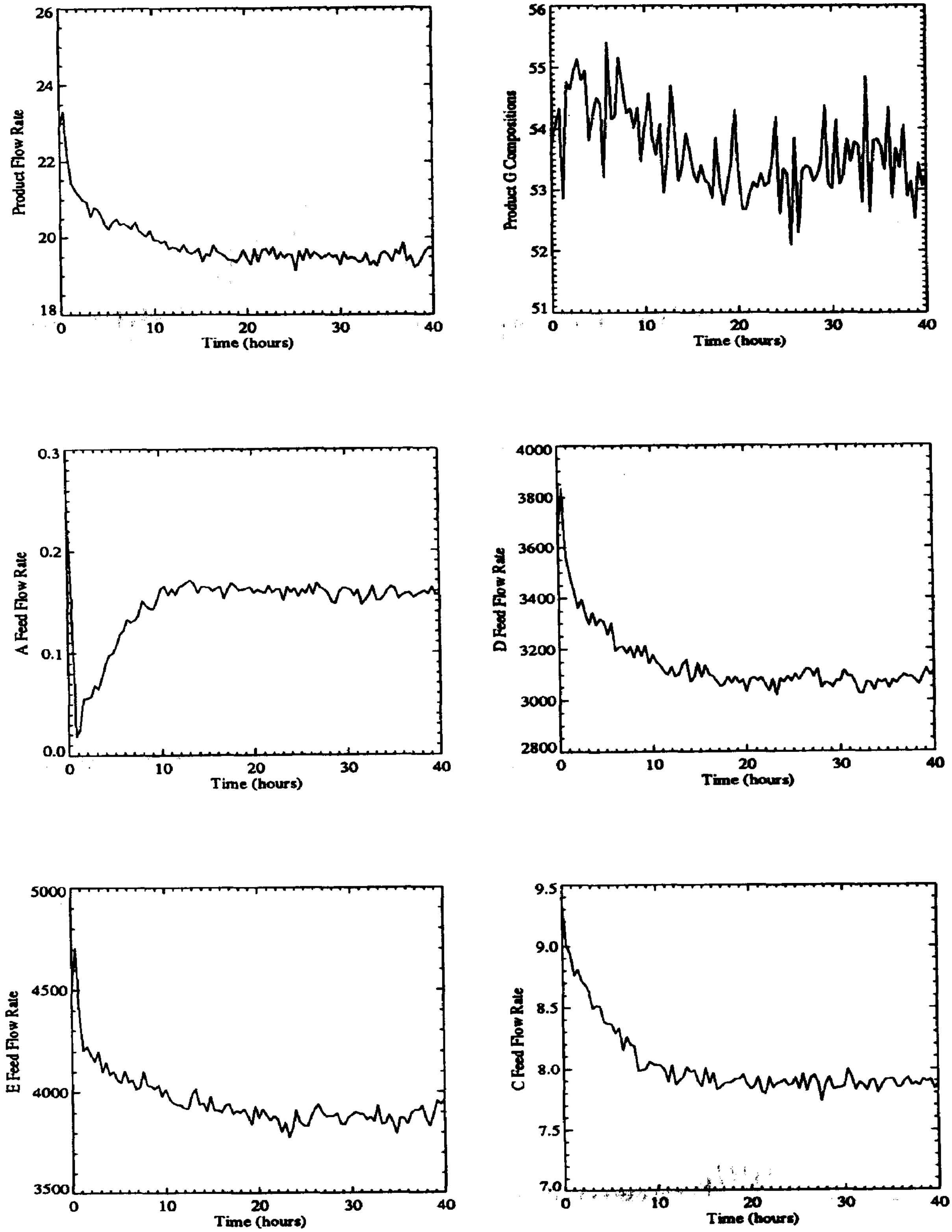

Fig. 13a. Product flowrate setpoint change—15%.

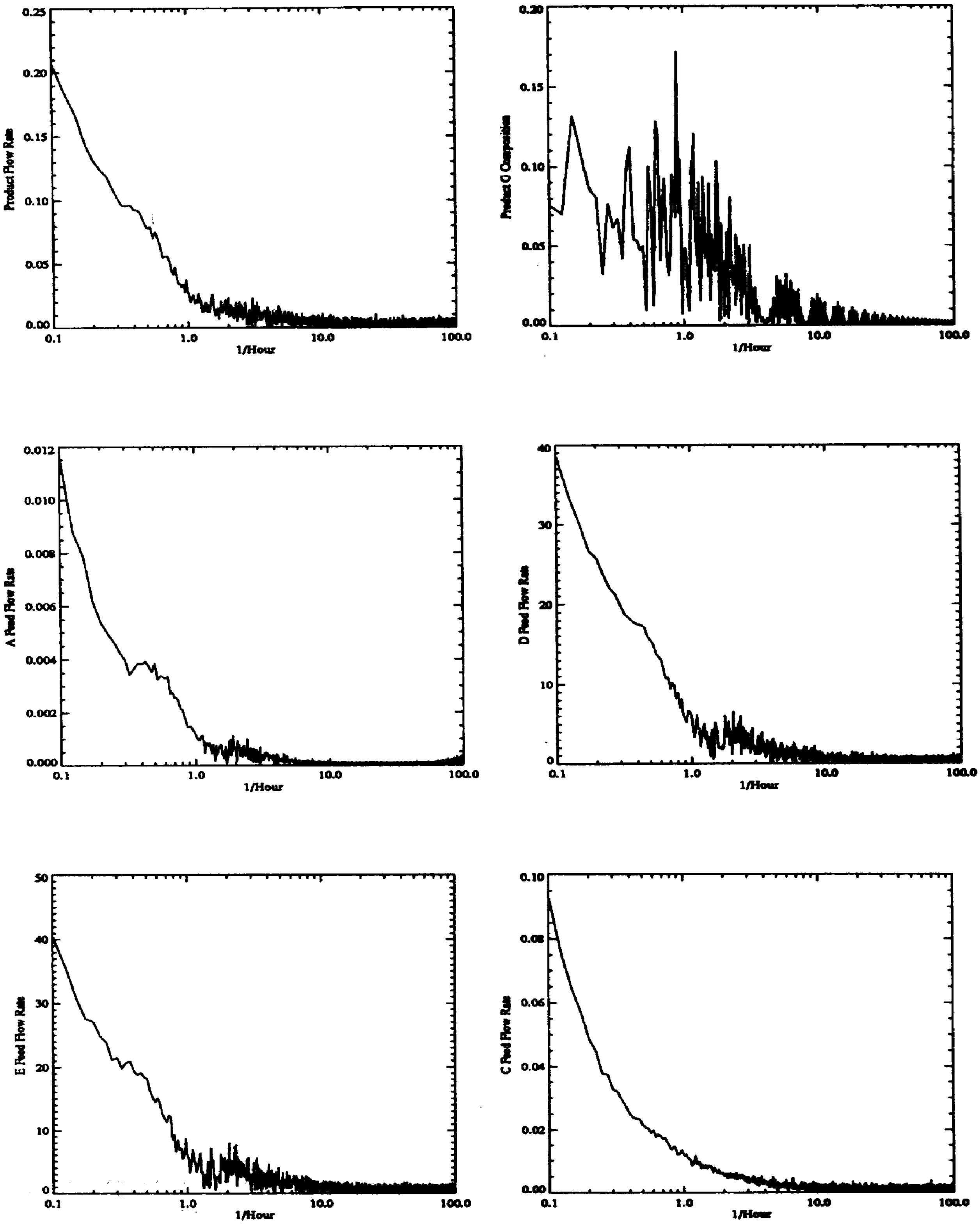

Fig. 13b. Product flowrate setpoint change—15% (Fourier coefficients).

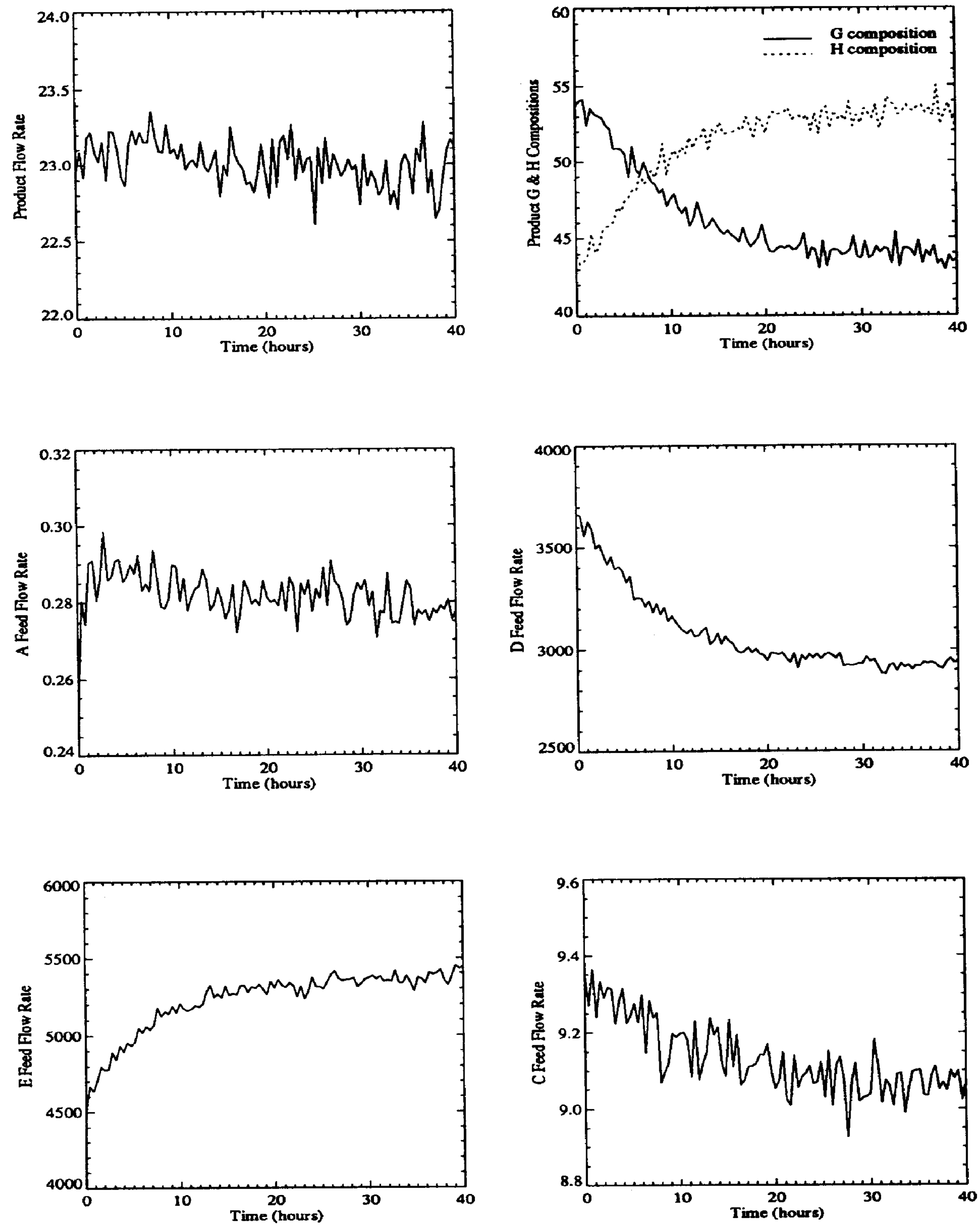

Fig. 14a. Product G/H ratio setpoint change from 50/50 to 40/60.

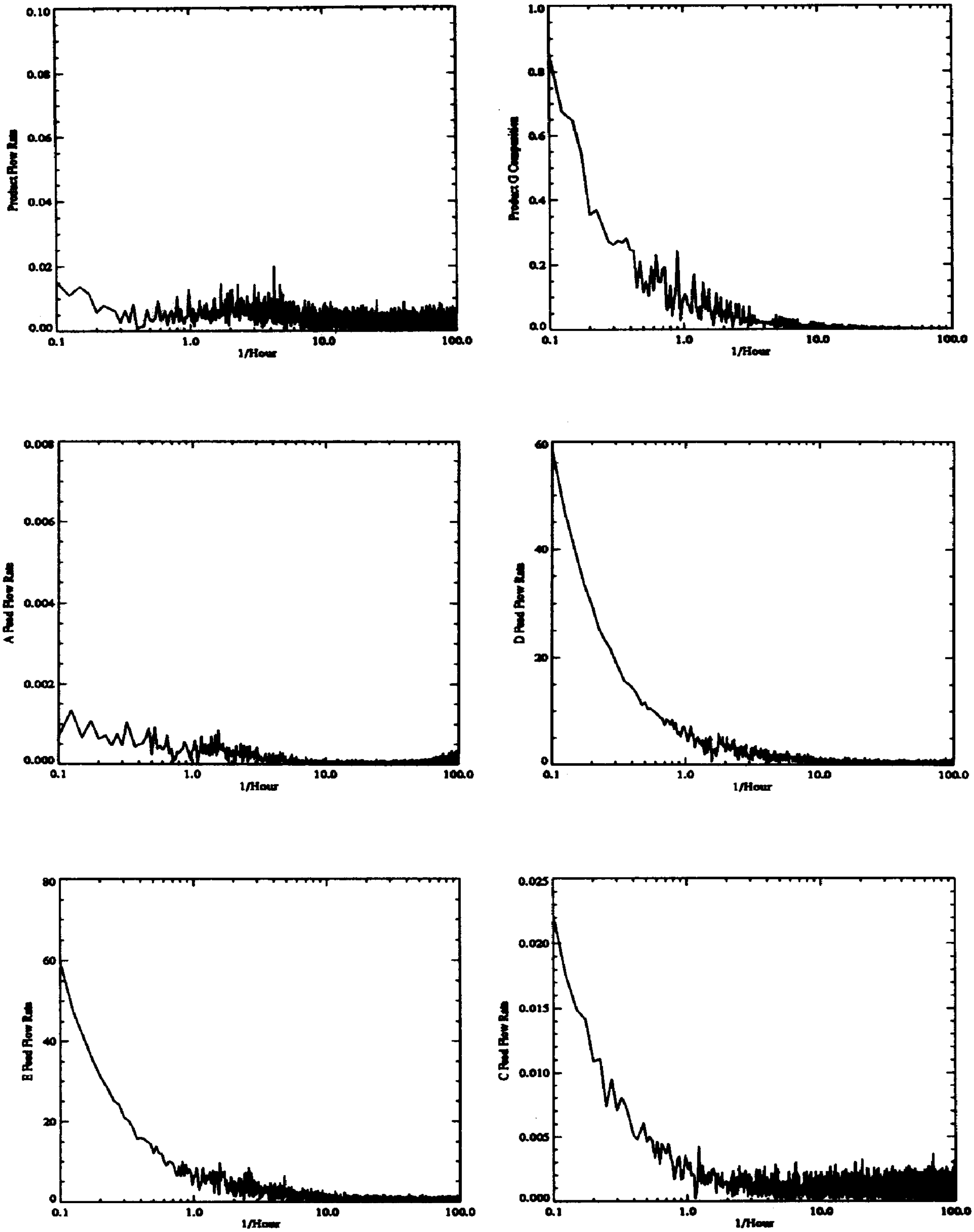

Fig. 14b. Product G/H ratio setpoint change from 50/50 to 40/60 (Fourier coefficients).

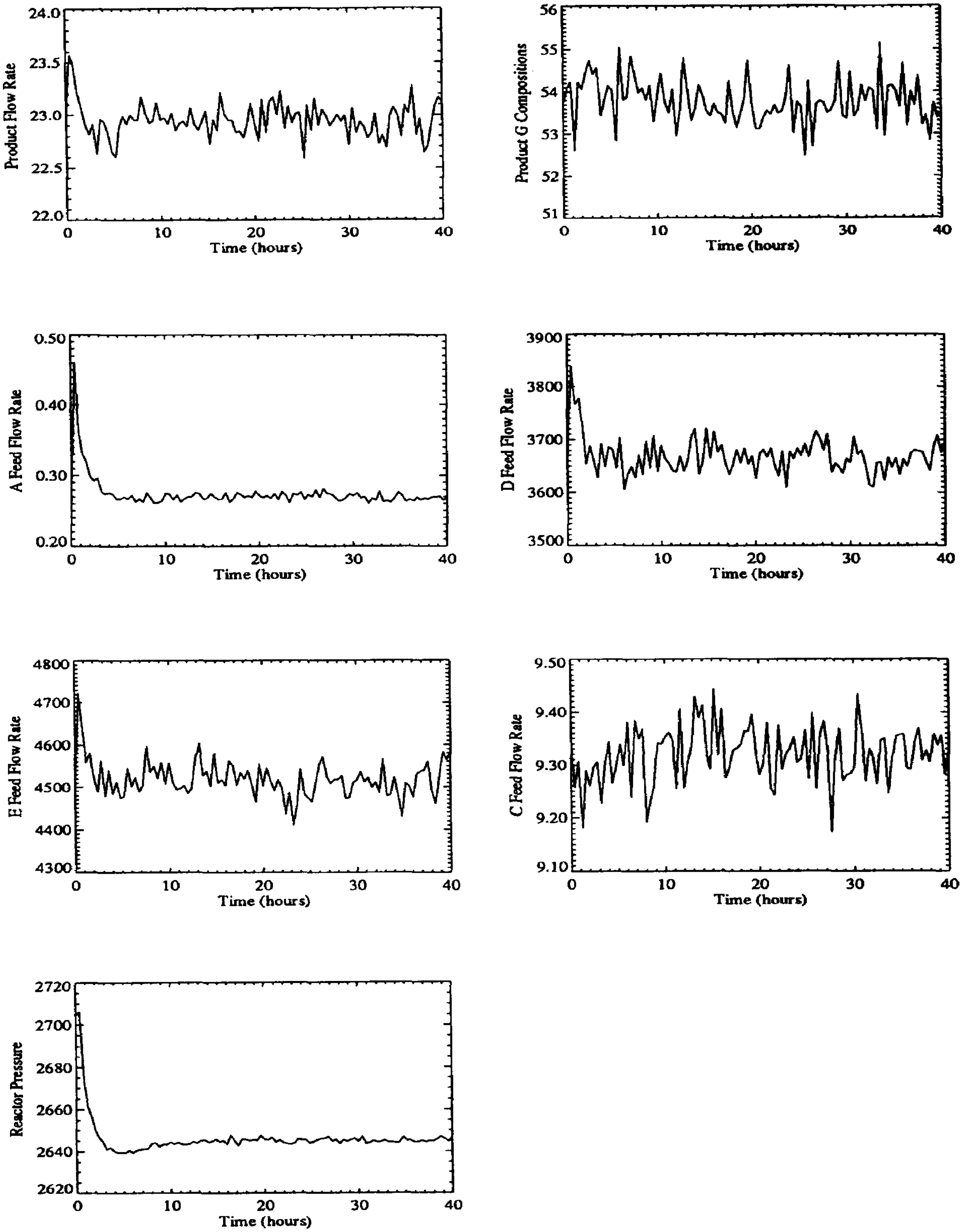

Fig. 15a. Reactor pressure setpoint change—60 kPa.

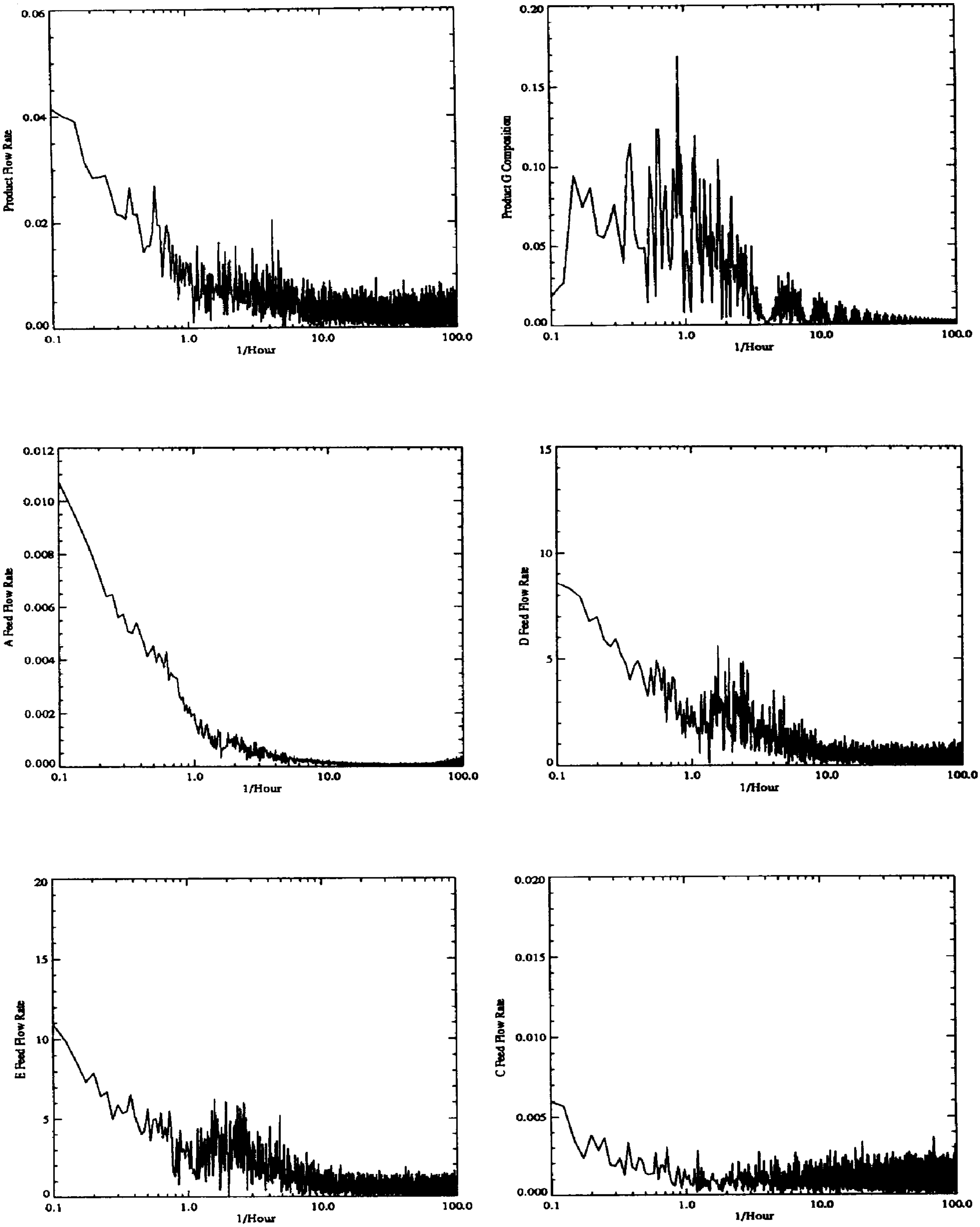

Fig. 15b. Reactor pressure setpoint change—60 kPa (Fourier coefficients).

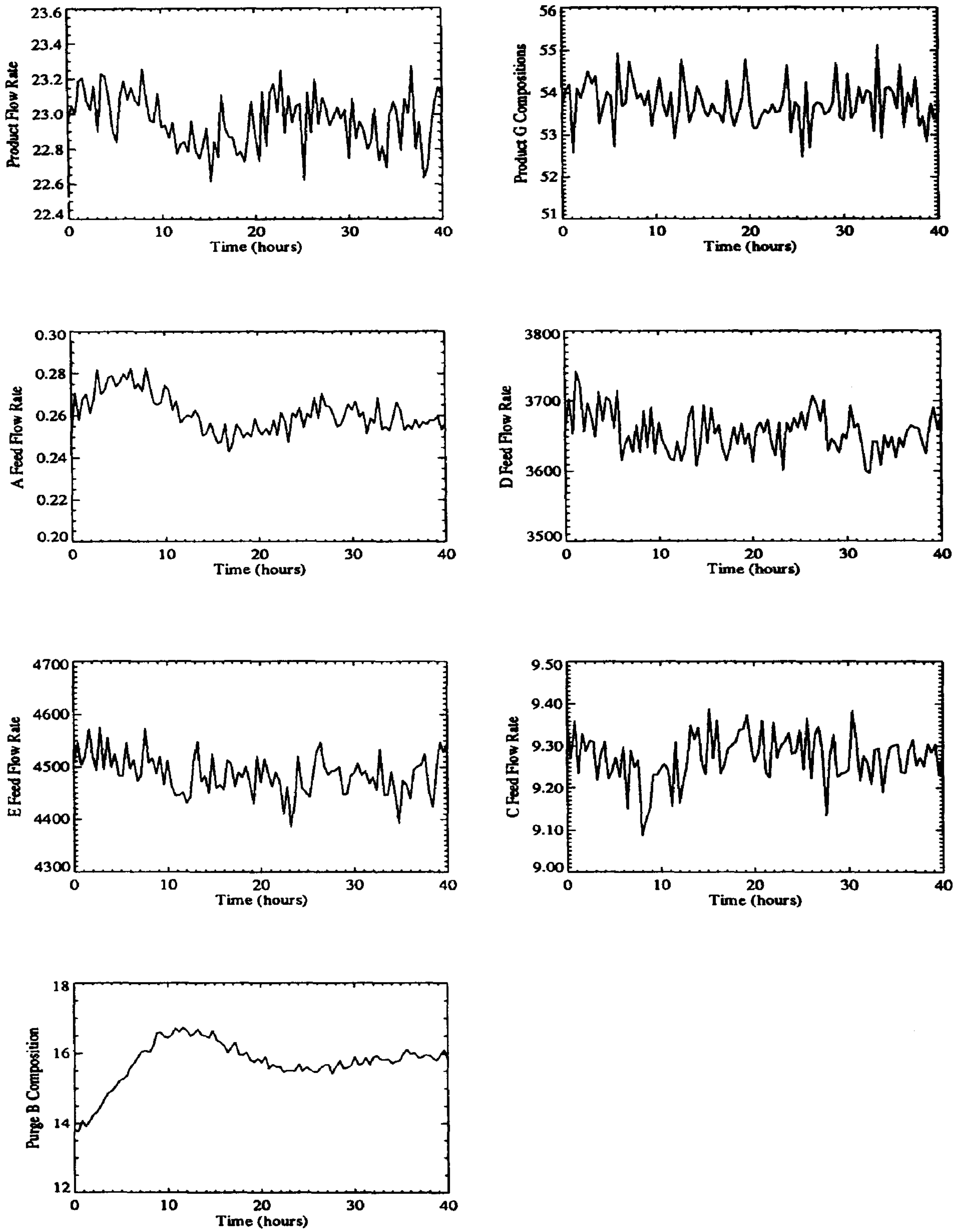

Fig. 16a. Purge B composition setpoint change 2%.

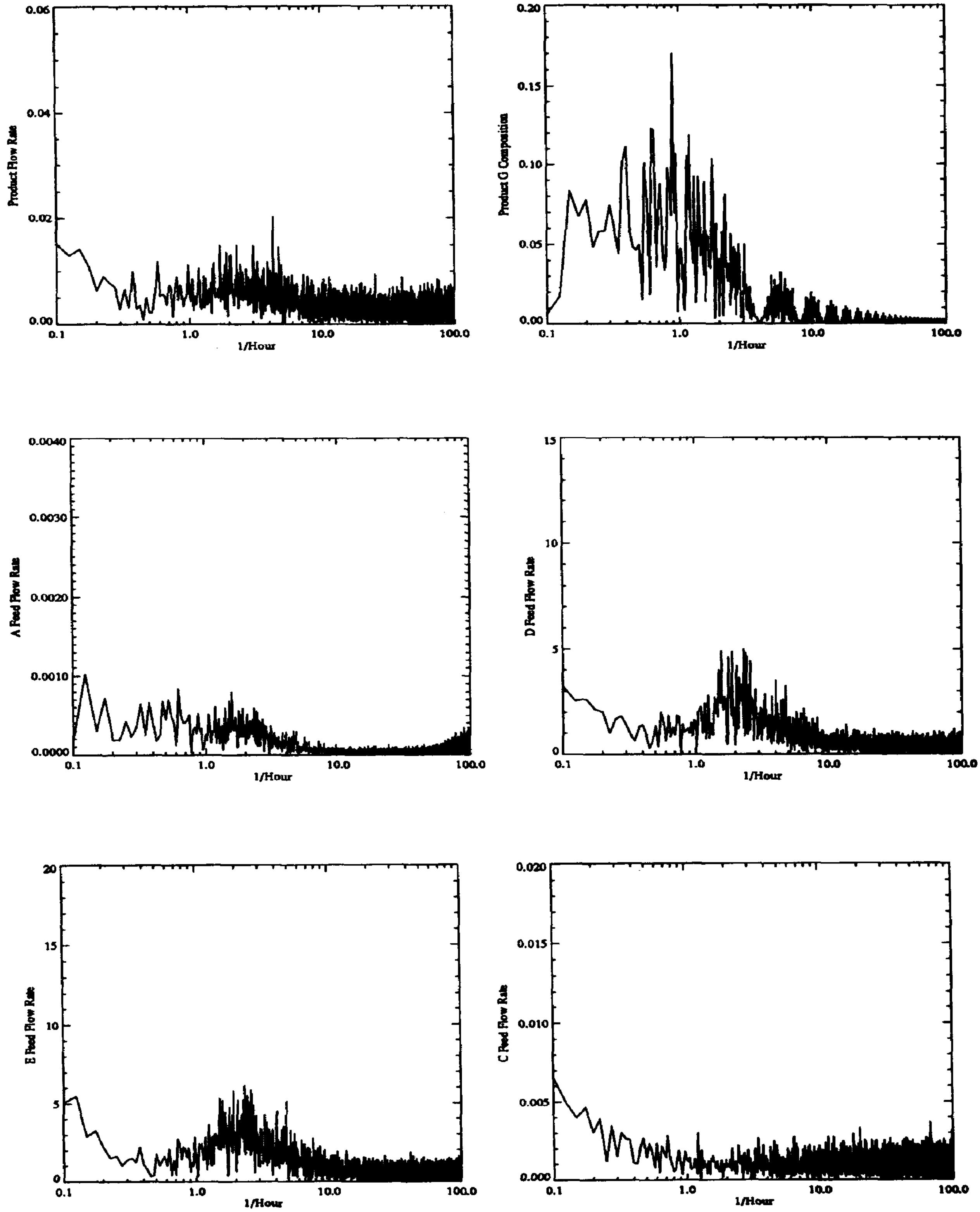

Fig. 16b. Purge B composition setpoint change 2% (Fourier coefficients).

CONTROL SYSTEM RESULTS

In their paper, Downs and Vogel suggested that the following setpoint changes and upsets be considered in evaluating potential control schemes:

IDV(1)	Step change
IDV(4)	Step change
IDV(8)	Random variation
IDV(12), IDV(15)	Simultaneous random variation and sticking valve
Production rate	Step change −15%
Product mix	Step change 50/50–40/60 G/H
Pressure change	Step change −60 kPa
Composition of B	Step change 2%

For comparison purposes they also suggested presenting the frequency content of process flowrates to these upsets. The subroutine FFTRF in the IMSL Math/Library was used to calculate the frequency spectra of the flowrates. FFTRF computes the discrete Fourier transform of a real vector of size N. In FFTRF, it is assumed that the real vector repeats itself periodically. In our calculations, $N = 8000$ and the corresponding time is 40 h. The Fourier coefficients calculated by FFTRF are divided by N/2. Using this approach for a unity amplitude cosine function of frequency ω, gives Fourier coefficients which are all zero except at the frequency ω where the Fourier coefficient equals 1.

Our base control system gave almost perfect results for IDV(4) and the IDV(12) + IDV(15) combination. Thus, responses to these upsets are not shown. Figures 11–16 give the results for the remaining disturbances, together with the frequency content of the process flows. As can be seen, some of the responses can take as long as 20–40 h to die out. This long transient period is due to the recycle nature of the plant. In all cases tested, all control valves remained within their saturation limits. Thus, the scheme presented provides an acceptable solution to the plant wide control problem that was posed. Our results can be used as a basis to judge the benefits and improvements that can be achieved from more advanced control approaches. In another paper (Ye and McAvoy, 1993) we discuss the benefits that can be gotten from the use of optimal averaging level control (McDonald *et al.*, 1986) on the Tennessee Eastman problem.

CONCLUSIONS

This paper has presented a methodology for designing a base, decentralized PID control system for the Tennessee Eastman Control Problem. The methodology involves screening various alternative designs using steady-state techniques such as the RGA, Niederlinski Index, and disturbance analysis. Engineering judgement is also employed. After, reducing the number of alternatives, dynamic simulation is used to tune loops and compare alternatives to arrive at a final scheme.

The approach used produces a final design that meets all of the requirements posed in the problem. It is shown that for one upset where the A feed is lost, a selector coupled with a production cutback is required to keep pressure under control. The control scheme presented can be used both to compare the improvements attained with an advanced control approach and as base system upon which an advanced scheme can be placed.

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