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APPENDICES

APPENDIX A

TUNING OF CONTROL STRUCTURES

A.1 Tuning Controllers

Notice throughout this work uses several types of controllers such as P, PI, and PID controllers. They depend on the control loop. In theory, control performance can be improved by the use of derivative action but in practice the use of derivative has some significant drawbacks:

1. Three tuning constants must be specified.
2. Signal noise is amplified.
3. Several types of PID control algorithms are used, so important to careful that the right algorithm is used with its matching tuning method.
4. The simulation is an approximation of the real plant. If high performance controllers are required to get good dynamics from the simulation, the real plant may not work well.

A.2 Tuning Flow, Level and Pressure Loops

The dynamics of flow measurement are fast. The time constants for moving control valves are small. Therefore, the controller can be turned with a small integral or reset time constant. A value of τ_I = 0.3 minutes work in most controllers. The value of controller gain should be kept modest because flow measurement signal are sometime noisy due to the turbulent flow through the orifice plate. A value of controller gain of K_c= 0.5 is often used. Derivative action should not be used.

Most level controllers should use proportional-only action with a gain of 1 to 2. This provides the maximum amount of flow smoothing. Proportional control means there will be steady state offset (the level will not be returned to its setpoint value). However, maintaining a liquid level at a certain value is often not necessary when the liquid capacity is simply being used as surge volume. So the recommended tuning of a level controller is $K_c = 2$. Most pressure controllers can be fairly easily tuned. The process time constant is estimated by dividing the gas volume of the system by the volumetric flowrate of gas flowing through the system. Setting the integral time equal to about 2 to 4 times the process time constant and using a reasonable controller gain usually gives satisfactory pressure control. Typical pressure controller tuning constants for columns and tanks are $K_c = 2$ and $\tau_I = 10$ minutes.

A.3 Relay- Feedback Testing

The relay-feedback test is a tool that serves a quick and simple method for identifying the dynamic parameters that are important for to design a feedback controller. The results of the test are the ultimate gain and the ultimate frequency. This information is usually sufficient to permit us to calculate some reasonable controller tuning constants.

The method consists of merely inserting an on-off relay in the feedback loop. The only parameter that must be specified is the height of the relay, h. This height is typically 5 to 10 percent of the controller output scale. The loop starts to oscillate around the setpoint with the controller output switching every time the process variable (PV) signal crosses the setpoint. Figure B.1 shows the PV and OP signals from a typical relay-feedback test. The maximum amplitude (a) of the PV signal is used to calculate the ultimate gain, K_U from the equation

$$K_U = \frac{4h}{a\pi} \qquad (1)$$

The period of the output PV curve is the ultimate period, P_U from these two parameters controller tuning constants can be calculated for PI and PID controllers, using a variety of tuning methods proposed in the literature that require only the ultimate gain and the ultimate frequency, e.g. Ziegler-Nichols, Tyreus-Luyben.

The test has many positive features that have led to its widespread use in real plants as well in simulation studies:

1. Only one parameter has to be specified (relay height).

2. The time it takes to run the test is short, particularly compared to the extended periods required for methods like PRBS.
3. The test is closedloop, so the process is not driven away from the setpoint.
4. The information obtained is very accurate in the frequency range that is important for the design of a feedback controller.
5. The impact of load changes that occur during the test can be detected by a change to asymmetric pulses in the manipulated variable.

These entire features make relay-feedback testing a useful identification tool. Knowing the ultimate gain, K_U and the ultimate period, P_U permits us to calculate controller settings. There are several methods that require only these two parameters. The Ziegler-Nichols tuning equations for a PI controller are:

$$K_C = K_U / 2.2 \qquad (2)$$

$$\tau_I = P_U / 1.2 \qquad (3)$$

These tuning constants are frequently too aggressive for many chemical engineering applications. The Tyreus-Luyben tuning method provides more conservative settings with increased robustness. The TL equations for a PI controller are:

$$K_C = K_U / 3.2 \qquad (4)$$

$$\tau_I = 2.2 P_U \qquad (5)$$

A.4 Inclusion of Lags

Any real physical system has many lags. Measurement and actuator lags always exist. In simulations, however, these lags are not part of the unit models. Much more aggressive tuning is often possible on the simulation than is possible in the real plant. Thus the predictions of dynamic performance can be overly optimistic. This is poor engineering. A conservative design is needed. Realistic dynamic simulations require that we explicitly include lags and/or dead times in all the important loops. Usually this means controllers that affect Product quality or process constraint. Table

B.1 summarizes some recommended lags to include in several different types of control loops.

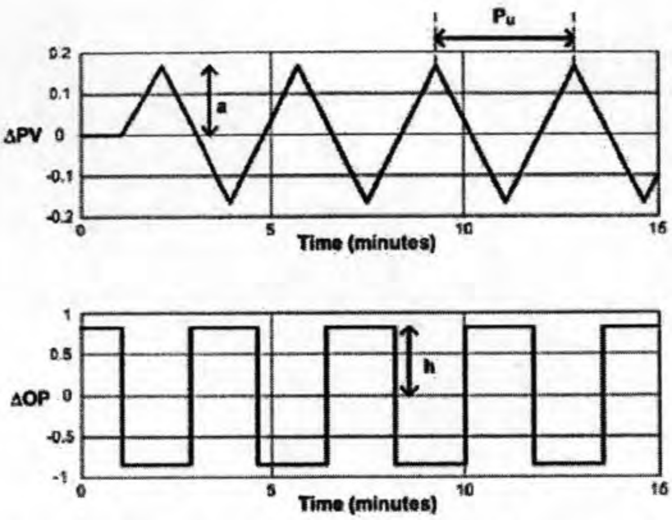

Figure A.1 Input and Output from Relay-Feedback Test

Table A.1 Typical measurement lags

		Number	Time constant (minutes)	Type
Temperature	Liquid	2	0.5	First-order lags
	Gas	3	1	First-order lags
Composition	Chromatograph	1	3 to 10	Deadtime

Any real physical system has many lags. Measurement and actuator lags always exist. In simulations, however, these lags are not part of the unit models. Much more aggressive tuning is often possible on the simulation than is possible in the real plant. Thus the predictions of dynamic performance can be overly optimistic. This is poor engineering. A conservative design is needed.

Realistic dynamic simulations require that we explicitly include lags and/or dead times in all the important loops. Usually this means controllers that affect Product quality or process constraint. Table B.1 summarizes some recommended lags to include in several different types of control loops.

APPENDIX B

EQUIPMENT AND DATA SPECIFICATION OF HDA PROCESS

Table B.1 Equipment data and Specifications of HDA Process

Equipments	*Specifications*	*Alternative*	
		5	6
Reactor	Diameter (m)	17.374	17.374
	Length (m)	2.905	2.905
	Number of tube	1	1
Furnace	Tube volume (m^3)	8.5	8.5
Cooler	Tube volume (m^3)	8.5	8.5
Separator	Liquid volume (m^3)	1.13	1.13
FEHE1*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	3.38×10^5	3.594×10^5
FEHE2*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	3.543×10^5	6.988×10^4
FEHE3*	Shell volume (m^3)	-	14.16
	Tube volume (m^3)	-	14.16
	UA (kJ/C-h)	-	1.601×10^5
Reboiler1 (R1)*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	5.817×10^5	5.502×10^4
Reboiler2 (R2)*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	1.379×10^5	1.159×10^5
Reboiler3 (R3)*	Shell volume (m^3)		14.16
	Tube volume (m^3)		14.16
	UA (kJ/C-h)		1.118×10^5
Condensor/Reboiler (CR)*	Shell volume (m^3)	14.16	14.16
	Tube volume (m^3)	14.16	14.16
	UA (kJ/C-h)	4.11×10^5	4.595×10^5
Tank Bottom C1(TB1)**	Vesel volume (m^3)	7.08	7.08
Tank Bottom C2(TB2)**	Vesel volume (m^3)	8.50	8.50
Tank Bottom C3(TB3)**	Vesel volume (m^3)	-	1.42
Tank Top C3 (TT3)* **	Vesel volume (m^3)	2.83	2.83

Table B.3 Column Specifications of HDA process alternative 6

Stream name	*Stabilizer Column*	*Product Column*	*Recycle Column*
Model	Refluxed Absorber	Refluxed Absorber	Absorber
Number of theoretical tray	6	27	7
Feed tray	3	15	5
Pressure (kPa)	1034.25	206.85	540.00
Diameter (m)	1.067	1.981	0.6096
Weir length (m)	0.8842	1.405	0.4544
Weir height (m)	0.0508	0.0508	0.0508
Tray spacing (m)	0.6096	0.6096	0.6096
Tray type	Sieve	Sieve	Sieve
Reboiler vol. (m3)	-	-	-
Condenser vol. (m3)	0.510	10.37	-
Specification 1	Benzene mole fraction in overhead = 0.042	Toluene mole fraction in overhead = 0.0003	-
Specification 2	-	-	-

VITA

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