REFERENCES

THAI

ไพศาล กิตติศุภกร. เอกสารคำสอนวิชา 2105-619 การควบคุมกระบวนการอัตโนมัติขั้นสูง. ภาควิชาวิศวกรรมเคมี จุฬาลงกรณ์มหาวิทยาลัย, 2543.

มนัส สังวรศิลป์ และ วรรัตน์ ภัทรอมรกุล. คู่มือการใช้งาน MATLAB ฉบับสมบูรณ์. กรุงเทพมหานคร: ศูนย์การพิมพ์พลชัย,2543.

ENGLISH

Arkljn, Y.; and Downs, J. A general method to calculate input-output gains and the relative gain array for integrating processes. Comp. Chem. Eng. 14 (1990):1101-1110.

Bequette, B.W. Process Control Modeling, Design, and Simulation. USA.: Prentice Hall, 2003.

Cao, Y; Rossiter, D.; and Owens, D. Input selection for disturbance rejection under manipulated variable constraints. Comp. Chem. Eng. 21(1997): Suppl., pp. $403-$408.

Cao, Y; and Rossiter, D. An input pre-screening technique for control structure selection. Comp. Chem. Eng. 21 (1997): 563-569.

Cao, Y.; Rossiter, D.; Edwards, D. W.; Knechtel, J.; and Owens, D. Modelling issues for control structure selection in a chemical process. Comp. Chem. Eng. 22 (1998): Suppl., pp. s411-s418.

Chen, D.; and Seborg, D. E. Relative gain array analysis for uncertain process models. AIChE 48 (2002): 302-310.

Chodavarapu, S. K.; and Zheng, A. Definition of Steady-State Plantwide Controllability. Ind. Eng. Chem. Res. 41 (2002): 4338-4345.

Dimain, A. C.; Groenendijk, A. J.; and Iedema, P. Systems analysis in handling impurities in complex plants. Comp. Chem. Eng. 20 (1996): Suppl., pp. $805-$810.

Dimian, A. C. ; Groenendijk, A. J.; and Iedema, P. D. Recycle interaction effects on the control of impurities in a complex plant. Ind. Eng. Chem. Res. 40 (2001): 5784-5794.

Douglas, J. M. Conceptual Design of Chemical Processes. Singapore: McGraw Hill., 1988.

Driels, M. Linear Control Systems Engineering. Singapore: McGraw Hill., 1996.

Fisher, W. R.; Doherty, M. F.; and Douglas, J. M. The interface between design and control. 1. process controllability. Ind. Eng. Chem. Res. 27 (1988): 597-605.

Grosdldler, P.;and Morarl, M. Closed-loop properties from steady-state gain information. Ind. Eng. Chem. Fundam. 24 (1985): 221-235.

Groenendijk, A. J.; Dimain, A. C.; Iedema, P.; and Kersten, S. R. A. Effect of recycle interactions on dynamics and control of complex plants. Comp. Chem. Eng. 21 (1997): Suppl., pp. S291-s296.

Groenendijk, A. J.; Dimain, A. C.; and Iedema, P. Systems approach for evaluating dynamics and plantwide control of complex plants. AIChE 46 (2000):133-145.

Haggblom, K. E. Analytical approach to evaluation of distillation control structures by frequency-dependent relative gains. Comp. Chem. Eng. 21 (1997): 1441-1449.

Herrmann, G.; Spurgeon, S. K.;and Edwards, C. A model-based sliding mode control methodology applied to the HDA-plant. Journal of Process Control 13 (2003) :129–138.

Hovd, M.; and Skogestad, S. Simple frequency-dependent tools for control system analysis, structure selection and design. Automatica 28 (1992): 989-996.

Kariwala, V.; Forbes, J. F.; and Meadows, E. S. Block relative gain: properties and pairing rules. Ind. Eng. Chem. Res. 42 (2003): 4564-4574

Kookos, I. K.; and Perkins, J. D. Heuristic-based mathematical programming framework for control structure selection. Ind. Eng. Chem. Res. 40 (2001): 2079-2088.

Lyman, P. R.; and Georgakis, C. Plantwide control of the Tennessee Eastman problem. Comp. Chem. Eng. 19 (1995): 321-331.

Luyben, W. L. Process Modeling, Simulation and Control for Chemical Engineers. Singapore: McGraw Hill., 1990.

Luyben, M. L.; and Tyreus, B. D. Plantwide control design procedure. AIChE 43 (1997): 3161-3174.

Luyben, W. L.; Tyreus, B.D.; and Luyben, M. L. Plantwide Process Control. Singapore: McGraw Hill., 1999.

Luyben, W. L. Control of outlet temperature in adiabatic tubular reactors. Ind. Eng. Chem. Res. 39 (2000): 1271-1278.

Luyben, W. L. Impact of reaction activation energy on plantwide control structures in adiabatic tubular reactor systems. Ind. Eng. Chem. Res. 39 (2000): 2345-2354.

Luyben, W. L. Plantwide Dynamic Simulators in Chemical Processing and Control. USA.: Marcel Dekker Inc., 2002.

McAvoy, T.; and Miller, R. Incorporating integrating variables into steady-state models for plantwide control analysis and design. Ind. Eng. Chem. Res. 38 (1999): 412-420.

McAvoy, T. Synthesis of plantwide control systems using optimization. Ind. Eng. Chem. Res. 38 (1999): 2984-2994.

McAvoy, T.; and Braatz, R. D. Controllability of processes with large singular values. Ind. Eng. Chem. Res. 42 (2003): 6155-6165.

Ogunnaike, B. A.; and Ray, H. W. Process Dynamics, Modeling, and Control. USA.: Oxford University Press Inc., 1994.

Price, R. M.; Lyman, P. R.; and Georgakis, C. Throughput manipulation in plantwide control structures. Ind. Eng. Chem. Res. 33 (1994): 1197-1207.

Qiu, Q.F.; Rangaiah, G.P.;and Krishnaswamy, P.R. Application of a plant-wide control design to the HDA process. Comp. Chem. Eng. 27 (2003): 73- 94.

Robinson, D.; Chen, R.; McAvoy, T.; and Schnelle, P. D. An optimal control based approach to designing plantwide control system architectures. Journal of Process Control 11 (2001): 223-236.

Samyudia, Y.; LEE, P. L.; and Cameron, T. A new approach to decentralised control design. Comp. Chem. Eng. 50 (1995): 1695-1706.

Schmidt, H.; and Jacobsen, E. W. Selecting control configurations for performance with independent design. Comp. Chem. Eng. 27 (2003): 101-109.

Seborg, D. E.; Edgar, T. F.; and Mellichamp, D. A. Process Dynamic and Control. Canada: John Wiley & Sons Ltd ,1989.

Semino, D.; and Giuhani, G. Control configuration selection in recycle systems by steady state analysis. Comp. Chem. Eng. 21 (1997): Suppl., pp. $273-$278.

Skogestad, S.; and Morari, M. Effect of disturbance directions on closed-loop performance. Ind. Eng. Chem. Res. 26 (1987): 2029-2035.

Skogestad, S.; and Morari, M. Implications of large RGA elements on control performance. Ind. Eng. Chem. Res. 26 (1987): 2323-2330.

Skogestad, S. Control Structure Design. AIChE Annual Meeting, Chicago, 14 Nov. 1996 paper 149a.

Skogestad, S. A procedure for SISO controllability analysis – with application to design of pH neutralization process. Comp. Chem. Eng. 20 (1996): 373-386.

Skogestad, S.; and Havre, K. The use of RGA and condition number as robustness measures. Comp. Chem. Eng. 20 (1996): Suppl., pp. SI005-SI010.

Skogestad, S.; and Postlethwaite, I. Multivariable Feedback Control Analysis and Design. England: John Wiley & Sons Ltd ,1996.

Skogestad, S. Control structure design for complete chemical plants. Comp. Chem. Eng. 28 (2004): 219-234.

Smith, C. A.; and Corripio, A. B. Principles and Practice of Automatic Process Control. Canada: John Wiley & Sons Ltd ,1997.

Stephanopoulos, G. Chemical Process Control an Introduction to Theory and Practice. New Jersey: Prentice Hall, 1984.

Stephanopoulos, G.; and Ng, C. Perspectives on the synthesis of plant-wide control structures. Journal of Process Control 10 (2000): 97-111.

Wang, P.; and McAvoy, T. Synthesis of plantwide control systems using a dynamic model and optimization. Ind. Eng. Chem. Res. 40 (2001); 5732-5742.

Yi, C. K.; and Luyben, W. L. Evaluation of plant-wide control structures by steady-state disturbance sensitivity analysis. Ind. Eng. Chem. Res. 34 (1995): 2393-2405.

Zhu, Z. X. Variable pairing selection based on individual and overall interaction measures. Ind. Eng. Chem. Res. 35 (1996): 4091-4099.

APPENDICES

APPENDIX A

RELATIVE GAIN ARRAY

There are two basic methods used for calculating the RGA for a square, linear system, whose transfer function matrix $\mathbf{G}(s)$ is available: (1) the "first principles" method, and (2) the matrix method.

Calculating RGA's from First Principles

We will illustrate this procedure with the 2 x 2 system of Eqs. (3.1) and (3.2).

First, we observe that, by definition, the RGA is concerned with steady-state conditions, thus we only need the steady-state form of this model, which is:

$$y_1 = K_{11}m_1 + K_{12}m_2 \tag{A.1}$$

$$y_2 = K_{21}m_1 + K_{22}m_2 \tag{A.2}$$

To obtain λ_{11} from the definition in Eq. (3.6), we need to evaluate both of the indicated partial derivatives from Eqs. (A.1) and (A.2) as follows: Eqs. (A.1) and (A.2) represent steady-state, open-loop conditions; therefore, the numerator partial derivative in Eq. (3.6) is obtained from Eq. (A.1) by straightforward differentiation:

$$\left(\frac{\partial y_1}{\partial m_1}\right)_{\text{all loops open}} = K_{11} \tag{A.3}$$

The second partial derivative calls for Loop 2 to be closed, so that in response to changes in m_1, the second control variable m_2 can be used ultimately to restore y_2 to its initial value of 0. Thus, to obtain the second partial derivative, we first find, from Eq. (A.2), the value m_2 must take to keep $y_2 = 0$ in the face of changes in m_1; what effect this will have on y_1 is deduced by substituting this value for m_2 in Eq. (A.1).

Setting $y_2 = 0$ and solving for m_2 in Eq. (A.2) gives:

$$m_2 = -\frac{K_{21}}{K_{22}}m_1 \tag{A.4}$$

and substituting this into Eq. (A.1) gives:

$$y_1 = K_{11}m_1 - \frac{K_{12}K_{21}}{K_{22}}m_1 \tag{A.5}$$

Having thus "eliminated" m_2 we may then differentiate, so that now:

$$\left(\frac{\partial y_1}{\partial m_1}\right)_{\text{loop 2 closed}} = K_{11}\left(1 - \frac{K_{12}K_{21}}{K_{11}K_{22}}\right) \tag{A.6}$$

and we obtain, finally, that:

$$\lambda_{11} = \frac{1}{1-\zeta} \tag{A.7}$$

where

$$\zeta = \frac{K_{12}K_{21}}{K_{11}K_{22}} \tag{A.8}$$

As an exercise, the reader is encouraged to follow the illustrated procedure and confirm that the other relative gains, λ_{12}, λ_{21}, and λ_{22}, are given by:

$$\lambda_{12} = \lambda_{21} = \frac{-\zeta}{1-\zeta} \tag{A.9}$$

and

$$\lambda_{22} = \lambda_{11} = \frac{1}{1-\zeta} \tag{A.10}$$

Thus the RGA for the 2 x 2 system is given by:

$$\Lambda = \begin{bmatrix} \frac{1}{1-\zeta} & \frac{-\zeta}{1-\zeta} \\ \frac{-\zeta}{1-\zeta} & \frac{1}{1-\zeta} \end{bmatrix} \tag{A.11}$$

Note carefully that the elements add up to unity across any row as well as down any column.

It is useful to observe that if we define:

$$\lambda = \lambda_{11} = \frac{1}{1-\zeta} \tag{A.12}$$

for this 2 x 2 system, then the RGA may be written:

$$\Lambda = \begin{bmatrix} \lambda & 1-\lambda \\ 1-\lambda & \lambda \end{bmatrix} \qquad \text{(A.13)}$$

The immediate consequence is that the single element λ is sufficient to determine the RGA for a 2 x 2 system. Thus for 2 x 2 system, λ is often called the relative gain parameter.

If we realistically assess this first-principles method of obtaining the RGA, it will not be difficult to come to the conclusion that it will be too tedious to apply in cases where the system is of higher dimension than 2 x 2. The problem is not with the numerator partial derivatives – these are always easy to obtain. The problem is with the denominator partial derivatives: the procedure for their evaluation in higher dimensional systems will involve solving several systems of linear algebraic equations.

The fact that linear algebraic equations are quite conveniently solved by matrix methods should then alert us to the possibility of using matrix methods to simplify the tedium involved in calculating RGA's, especially for higher dimensional systems.

The Matrix Method for Calculating RGA's

Let **K** be the matrix of steady-state gains of the transfer function matrix **G**(s), i.e.:

$$\lim_{s \to 0} G(s) = K \qquad \text{(A.14)}$$

whose elements are K_{ij}; further, let **R** be the transpose of the inverse of this steady-state gain matrix, i.e.:

$$R = (K^{-1})^T$$

with elements r_{ij}. Then, it is possible to show that the elements of the RGA can be obtained from the elements of these two matrices according to:

$$\lambda_{ij} = K_{ij} r_{ij} \qquad \text{(A.15)}$$

It is important to note that the above equation indicates an element-by-element multiplication of the corresponding elements of the two matrices **K** and **R**; it has nothing to do with the standard matrix product.

Computing The RGA with MATLAB

If *G* is a constant matrix then the RGA can be computed using

```
RGA = G.*pinv(G.');
```

If *G(jw)* is a frequency-dependent matrix generated using the μ toolbox, e.g.

```
G = pck(A,B,C,D); omega=logspace(-2,2,41); Gw=frsp(G,omega);
```

Then the RGA as a function of frequency can be computed using

```
RGAw = veval('.*',Gw,vpinv(vtp(Gw)));
```

Appendix B

Data of HDA Process for Simulation

Name	Fresh Hydrogen	Fresh Toluene	purge	stabilizer gas	benzene product	diphenyl product
Vapour Fraction	1.00	0.00	1.00	1.00	0.00	0.00
Temperature [F]	86.00	86.00	113.00	123.87	221.99	558.83
Pressure [psia]	605.00	635.00	476.80	150.00	30.00	31.00
Molar Flow [lbmole/hr]	490.38	289.00	483.40	19.73	270.81	6.28
Mass Flow [lb/hr]	1194.96	26628.69	5379.81	343.48	21154.42	968.31
Heat Flow [Btu/hr]	-4.45E+05	1.75E+06	-8.87E+06	-5.15E+05	7.09E+06	6.50E+05
Comp Mole Frac (Hydrogen	0.9700	0.0000	0.4012	0.0902	0.0000	0.0000
Comp Mole Frac (Methane)	0.0300	0.0000	0.5875	0.8674	0.0000	0.0000
Comp Mole Frac (Benzene)	0.0000	0.0000	0.0101	0.0420	0.9997	0.0000
Comp Mole Frac (Toluene)	0.0000	1.0000	0.0012	0.0003	0.0003	0.0003
Comp Mole Frac (BiPhenyl)	0.0000	0.0000	0.0000	0.0000	0.0000	0.9997

Data of HDA Process for Simulation (Continue)

Name	gas recycle	toluene recycle	furnance in	reactor in	reactor out	quench
Vapour Fraction	1.00	0.00	1.00	1.00	1.00	0.00
Temperature [F]	148.71	284.17	1106.00	1150.00	1230.97	113.83
Pressure [psia]	575.00	615.00	513.00	503.00	486.00	486.00
Molar Flow [lbmole/hr]	3519.20	84.67	4383.25	4383.25	4383.25	109.10
Mass Flow [lb/hr]	39119.40	7800.72	74743.78	74743.78	74742.70	8654.94
Heat Flow [Btu/hr]	-6.35E+07	1.18E+06	6.33E+05	3.76E+06	3.76E+06	1.88E+06
Comp Mole Frac (Hydrogen	0.4020	0.0000	0.4313	0.4313	0.3668	0.0047
Comp Mole Frac (Methane)	0.5868	0.0000	0.4745	0.4745	0.5404	0.0449
Comp Mole Frac (Benzene)	0.0100	0.0006	0.0080	0.0080	0.0711	0.7124
Comp Mole Frac (Toluene)	0.0012	0.9993	0.0862	0.0862	0.0203	0.2215
Comp Mole Frac (BiPhenyl)	0.0000	0.0000	0.0000	0.0000	0.0014	0.0165

Data of HDA Process for Simulation (Continue)

Name	fehe hot in	fehe hot out	sep gas out	stabilizer feed	stabilizer bottom	product bottom
Vapour Fraction	1.00	0.95	1.00	0.03	0	0
Temperature [F]	1149.90	228.56	113.00	114.75	374.7792337	292.5600107
Pressure [psia]	486.00	480.00	476.80	152.00	153	33
Molar Flow [lbmole/hr]	4492.35	4492.35	4002.60	381.25	361.5212442	90.70702121
Mass Flow [lb/hr]	83397.64	83397.64	44545.55	30244.74	29901.25883	8746.842286
Heat Flow [Btu/hr]	5.64E+06	-5.60E+07	-7.34E+07	6.56E+06	10784879.86	1660476.999
Comp Mole Frac (Hydrogen	0.3580	0.3580	0.4012	0.0047	0.0000	0.0000
Comp Mole Frac (Methane)	0.5284	0.5284	0.5875	0.0449	0.0000	0.0000
Comp Mole Frac (Benzene)	0.0866	0.0866	0.0101	0.7124	0.7490	0.0006
Comp Mole Frac (Toluene)	0.0252	0.0252	0.0012	0.2215	0.2336	0.9302
Comp Mole Frac (BiPhenyl)	0.0018	0.0018	0.0000	0.0165	0.0174	0.0692

Appendix C

Parameter Tuning of Three Control Structures

Controller	Reference Control Structure		Control Structure 1		Control Structure 2	
	K_c	T_i	K_c	T_i	K_c	T_i
PCR	2	10	2	10	2	10
FC	0.225	0.0419	0.1	0.0125	0.1	0.0125
TCR	0.2	0.5	0.2	0.5	0.2	0.5
TCQ	0.105	0.383	0.105	0.383	0.0726	0.252
TCC	0.5	5	0.5	5	0.5	5
LC	2	-	2	-	2	-
CC	0.449	14.1	0.449	14.1	0.449	14.1
PC1	1	10	1	10	1	10
LC_C1	1	-	1	-	1	-
LC_R1	2	-	2	-	2	-
TC1	1	12	1	12	1	12
PC2	1	20	1	20	1	20
LC_C2	2	-	2	-	2	-
LC_R2	2	-	2	-	2	-
TC2	2	10	2	10	2	10
PC3	1	15	1	15	1	15
LC_C3	1	-	1	-	1	-
LC_R3	2.9	-	2.9	-	2.9	-
TC3	1	30	1	30	1	30

VITA

Miss Pattra Tangsombutvisit was born in Bangkok on October 23, 1979. She graduated Bachelor Degree in Science of department of Chemical Technology from Chulalongkorn University in 1999. After that she studied for Master Degree in Chemical Engineering, Chulalongkorn University in 2000.