Process dynamics and control 4th edition pdf

The new 4th edition of Seborg's Process Dynamics Control provides full local coverage for process control courses in the chemical engineering curriculum, highlighting how process control and related process modeling and optimization areas are essential for the development of high-value products. The main objective of this new version is to describe modern techniques for control procedures, with an emphasis on complex systems necessary for the development, design and operation of modern processing units. Control process trainers can cover the basic material and also have the flexibility to include advanced themes. Chapter 2 2.1 (a) Total mass balance: d (rV) = w1 + w2 − w3 dt (1) Energy balance: C d ⎡ev (T3 − Tref) o o o = w1C (T1 − Tref) + w2C (T2 − Tref) dt − w3C (T3 − Tref) (2) Because r = constant and V = constant, constant and V = constant, Eq. 1 becomes: w3 = w1 + w2 b) From Eq. 2, substitution Eq. 3 rCV d (T3 − Tref) dT = rCV 3= w1C (T1 − Tref) + w2C (T2 − Tref) dt − (w1 + w2) C (T3 − Tref) (4) C and Tref constants can be cancelled: rV dT3 = w1T1 + w2T2 − (w1 + w2) T3 dt The simplified model now consists only of Eq. 5. Solution Manual for Process Dynamics and Control, 4th Edition Copyright © 2016 by Dale E. Seborg, Thomas F. Edgar, Duncan A. Mellichamp and Francis J. Doyle III 2-1 (5) Degrees of freedom for the simplified model: Parameters : r, V Variables : w1, w2, T1, T2, T3 NE = 1 NV = 5 So, NF = 5 − 1 = 4 Because w1, w2, T1 and T2 are determined by upstream units , we assume that time functions are known: w1 = w1(t) w2 = w2 (t) T1 = T1(t) T2 = T2(t) Thus, NF is reduced to 0. 2.2 Energy balance: Cp d o ⎡a (T − Tref) o o = wC p (Ti − Tref) − wC p (T − Tref) − UAs (T − Ta) + Q dt Simplification dT = wC p Ti − wC p T − UAs (T − Ta) + Q dt dT pVC p = wC p (Ti − T) − UAs (T − Ta) + Q dt pVC p b) T increases if Ti increases and vice versa. T decreases if w increases and vice versa if (Ti − T) < 0. In other words, if q > UAs(T-Ta), the content is heated, and T >Ti. 2-2 2.3 a) Mass balances: rA1 dh1 = w1 − w2 − w3 dt (1) dh2 = w2 dt (2) Let P1 be the pressure at the bottom of tank 1. Allow P2 to be the pressure at the bottom of tank 2. Let pa be the pressure of the environment. w2 = Then P1 − P2 pg (h1 − h2) = R2 g c R2 (3) P1 − Pa pg h1 = R3 g c R3 (4) w3 = b) Seven parameters: r, A1, A2, g, gc, R2, R3 Five variables : h1, h2, w1, w2, w3 Four equations So NF = 5 − 4 = 1 1 input = w1 (defined function of time) 4 outputs = h1, h2, w2, w3 2-3 2.4 Suppose constant liquid density, r . The mass balance for the tank is d (rAh + m g) dt = r(qi − q) Because r, A and mg are stable, this becomes A dh = qi − q dt (1) The square root ratio for flow through the control valve is ⎛ − rgh − Pa Pa q = C v ⎜⎜ Pg + gc ⎟ 1/ 2 (2) From the ideal gas law, Pg = (m g/ M) RT (3) A(H − h) where T is the absolute temperature of the gas. Equation 1 gives the model of unstable state when replacing q with Eq. 2 and Pg from Eq. 3: 1/ 2 o (mg / M) RT r gh = qi − Cv ⎜ − Pa o dh dt ⎝ A(H − h) (4) Because the model contains Pa, the function of the system is not independent of Pa. For an open system Pg = Pa and Eq. 2 indicates that the system is independent of Pa. 2-4 2.5 a) For the linear flow characteristics of the valve, Pd − P1 P − P2 , wb = 1 , Ra R b The mass balances for wave tanks wa = dm1 = wa − wb , but wC = P2 − Pf Rc dm2 = wb − wc dt (1) dt (2) where m1 and m2 are the masses gas tanks 1 and 2 Respectively. If the ideal gas law applies, then P1V1 = m1 RT1, M P2V2 = m2 RT2 M (3) where M is the molecular weight of the gas T1 and T2 are the temperatures in the wave tanks. Substituting m1 and m2 from Eq. 3 to Eq. 2 and noting that V1, T1, V2 and T2 are stable, the V2 M dP2 V1M dP1 = wa − wb and = wb − wc = wb − wc RT2 dt RT1 dt (4) The dynamic model consists of Eqs. 1 and 4. b) For adiabatic operation, Eq. 3 is replaced by γ ⎛V ⎛V ⎞ P1 ⎜⎜ 1 ⎟⎟ = P2 ⎜⎜ 2 ⎛ m2 ⎞ m1 ⎞ or ⎛ P1V1 γ m1 = ⎜⎜ ⎛ C ⎞ ⎟ ⎟ ⎟ ⎞⎟ = C , a constant ⎞ 1/ γ and ⎛ P2V2 γ m2 = ⎜⎜ ⎛ C Substituting Eq. 6 into Eq. 2 gives, 1 γ ⎛ V1 γ ⎞ ⎜ C ⎟ ⎟ ⎟ ⎞ 1/ γ P1 (1− γ) / γ 2-5 dP1 = wa − wb dt ⎟ ⎟ ⎞ (5) 1/ γ (6) 1 γ ⎛ V2 γ ⎞ ⎜ C ⎟ ⎟ ⎟ ⎞ 1/ γ P2 (1− γ) / γ dP2 = wb − wc dt as the new dynamic model. If the ideal gas law were not valid, one would use an appropriate equation of the state instead of Eq. 3. 2.6 a) Cases: 1. Each apartment is mixed perfectly. 2. r and C are fixed. 3. No heat loss in the environment. Compartment 1: Overall balance (No accumulation of mass): 0 = ρq − pq1 thus q1 = q (1) Energy balance (No change in volume): dT1 pqC (Ti − T1) − UA(T1 − T2) V1pC = dt (2) Compartment 2: Overall balance: 0 = pq1 − pq2 thus q2 = q1= q (3) Energy balance: dT2 V2pC = pqC (T1 − T2) + UA(T1 − T2) − U c Ac (T2 − Tc) dt b) Eight parameters: ρ, V1, V2, C, U, A, Uc, Ac Five variables: Ti, T1, T2, q, Tc Two equations: (2) and (4) 2-6 (4) Thus NF = 5 − 2 = 3 3 outputs = T1, T2 3 inputs = Ti, Tc, q (specify as functions of t) c) Three new variables: ci, c1, c2 (concentration of species A). Two new equations: The remaining material components in each apartment. c1 and c2 are new outputs. ci must be a known function of time. 2.7 As in section 2.4.2, there are two equations for this system: dV 1 ⎛ V = ⎜⎜ wi − Cv ⎜ dt r ⎝ A ⎟⎟ wi dT Q = (Ti − T) + dt rV rVC This leaves us with the following: V , T , wi , Ti , Q 4 parameters: C , r , Cv , A 2 equations The degrees of freedom are 5 − 2= 3 . To make sure the system is fixed, we have: 2 output variables: T , V 2-7 2 manipulated variables: Q, wi 1 variable disturbance: Ti (b) In this part, two controllers have been added to the system. Each controller provides an additional equation. Also, the flow out of the tank is now a manipulated variable regulated by the controller. So we have 4 parameters: C , r , Tsp , Vsp 6 variables: V , T , wi , Ti , Q, w 4 equations 2 . To determine the two degrees of freedom, we Freedom degrees are 6 − 4 = set the variables as follows: 2 output variables: T , V 2 manipulated variables: Q, w 2 disturbances: Ti , wi 2.8 Additional hypotheses: (i) Density of liquid, r, and density of refrigerant, rJ, are constant. (ii) The specific heat of the liquid, C, and refrigerant, CJ, is constant. Because V is stable, the mass balance for the tank is: r dV = q F − q = 0 ; so q = qF dt Energy balance for tank: rVC dT 0.8 = q F rC (TF − T) − Kq J A(T − TJ) dt (1) Energy balance for jacket: r J VJ C J d TJ dt = q J r J C J (Ti − TJ) + Kq J 0.8 A(T − TJ) (2) where A is the heat transfer area (in ft2) between the liquid process and the refrigerant. 2-8 Eqs.1 and 2 are the dynamic model for the system. 2-9 (3) 2.9 Suppose the food contains only A and B, and no C. The remaining ingredients for A, B, C beyond the reactor give. dc A = qi c Ai − qc A − Vk1e − E1 / RT c A dt (1) dcB = qi cBi − qcB + V (k1e − E1 / RT c A − k2e − E2 / RT cB) dt (2) dcC = −qcC + Vk2e − E2 / RT cb dt (3) V A total mass balance above the jacket indicates that qc = qci Because of refrigerant in the jacket and the density of the refrigerant are constant. Energy balance for the reactor: d ⎡(Vc A M A S A + VcB M B S B + VcC M C SC) T ⎤ = (qi c Ai M A S A + qi cBi M B S B) (Ti − T) dt (4) −UA(T − Tc) + (−ΔH1)Vk1e− E1 / RT c A + (−ΔH 2)Vk2e− E2 / RT cB where MA, MB, MC are molecular weights of A, B, and C, respectively SA, SB, SC are specific heats of A, B, and C. U is the overall heat transfer coefficient A is the surface area of heat transfer Energy balance for the jacket: dTc ρ j S j Vj= ρ j S j qci (Tci − Tc) + UA(T − Tc) j dt where : rj, Sj is density and special heat of the refrigerant. Vj is the volume of refrigerant in the jacket. Eqs. 1 - 5 represent the dynamic model for the system. 2-9 (5) 2.10 The plots should appear as shown below: Notice that the functions are good only for t = 0 to t = 18, which is fully drained. The function is blown up because the volume function is negative. Negative. 2.11 a) Note that the only retention equation required to find h is a total mass balance: dm d (rAh) dh w1 w2 − w = = rA =+ dt dt dt Dt Valve equation: w = C v′ pg h = Cv h gc where C v v′ pg gc (1) (2) (3) Replacing the valve equation in the mass balance, dh 1 = (w1 + − C v h) dt rA (4) Fixed state model: 0 = w1 + w2 − C v h w1 + w2 = 2.0 + 1.2 3.2 kg/s = = 2.13 1/2 1.5 2.25 m h b) C = v c) Feeder control 2-11 (5) Rearrangement 2,5 2,5 to get the feedforward controller ratio (FF), w2 = C v hR − w1 where hR = 2.25 m w2 = (2.13) (1.5) − w1 = 3.2 − w1 (6) Note that Eq. 6, 2014 , for a value w1 = 2.0, gives w2 = 3.2 − 1.2 = 2.0 kg/s which is the desired value. If the actual FF controller is the relationship, w2 = 3.2 − 1.1w1 (flow transmitter 10% higher), w2 will change as soon as the FF controller is activated, w2 = 3.2 −1.1 (2.0) = 3.2 − 2.2 = 1.0 kg/s (instead of the correct value, 1.2 kg/s) Then C v h = 2.13 h = 2.0 + 1.0 or h= 3 = 1.408 and h = 1.983 m (instead of 2.25 m) 2.13 Error in desired level = 2.25 − 1.983 × 100% = 11.9% 2.25 2-12 Sensitivity does not look very bad in the sense that a 10% error in the flow measurement gives ~ 12% error at the desired level. Before this conclusion, however, you should check how well the operating FF controller works for a change to w1 (e.g., D w1 = 0.4 kg/s). 2.12 (a) Tank model (normal operation): dh = w1 + w2 − w3 dt p 2 A= p= 3.14 m 2 rA (1800)(3.14) (Below the leakage point) dh = 120 + 100 − 200 = 0.007962 m /min dt (800)(3.14) Leak point reach time (h = 1 m) = 125.6 min. b) Leak tank model and w1 , w2 , w3 constant: rA dh =20 − d q4 =20 − r(0.025) h − 1 = 20 − 20 h − 1 , h ≥ 1 dt To check for overflow, someone can just find the hm level at which dh/ dt = 0. This is the maximum value of the layer when no overflow occurs. 0 = 20 − 20 hm − 1 or hm = 2 m So there is no overflow for leakage that occurs between hm < 2.25 m. 2-13 2.13 Process model Total material balance: rAT dh = w1 + w2 − w3 dt (1) Component: rAT d (hx3) = w1 x1 + w2 x 2 − w3 x3 dt rAT h dx3 dh + rAT x3 = w1 x1 + w2 x 2 − w3 x3 dt dt Replacement for dh/dt (Eq. 1) rAT h dx3 + x3 (w1 + w2 − w3) = w1 x1 + w2 x 2 − w3 x3 dt or a) dx3 1 ⎡w1 (x1 − x3) + w2 (x 2 − x3)⎤ = dt rAT h (2) (3) In initial stable condition , w3 = w1 + w2 = 120 + 100 = 220 Kg/min 220 = 166.3 Cv = 1.75 b) If x1 suddenly changes from 0.5 to 0.6 without changing the flow rates, then the level remains constant and Eq.3 can be solved in detail or numerically to find the time to achieve 99% of the x3 response. From the material balance, the final value is x3 = 0.555. Then dx3 1 = [120(0.6 − x3) + 100(0.5 − x3)] dt (800)(1.75)p 2-14 = 1 [(72 + − 220 x3)])] = 0,027738 − 0,050020 x3 dx3 = dt 0,027738 − 0,050020 x3 ∫0 όπου x3o=0,5 και x3f =0,555 − (0,555)(0,01) = 0,549 Επίλυση, t = 47,42 λεπτά y) Εάν το w1 αλλάξει σε 100 kg/min χωρίς να αλλοιώνεται καμία άλλη μεταβλητή εισόδου, τότε x3 δεν θα αλλάξει και Eq. 1 μπορεί να λυθεί για να βρει το χρόνο για να επιτευχθεί το 99% της απάντησης h. Από το ισοζύγιο υλικών, η τελική τιμή της στάθμης της δεξαμενής είναι h=1.446 m. 800π dh = 100 + 100 − Cv h dt 1 ⎰ dh 200 − 166,3 h = dt 800π ⎰ ⎰ = 0,079577 − 0,066169 h όπου ho =1,75 και hf =1,446 + (1,446)(0,01) = 1,460 Με τη χρήση της εντολής MATLAB ode45 , t = 122.79 minutes The solution of the ode is shown in the Figure. S2.13 2-15 1.8 1.7 h(m) 1.6 1.5 1.4 0 50 100 150 200 time (min) 250 300 Figure S2.13. Numerical solution of the ode for part c) d) In this case, both h and x3 will change the functions of time. Therefore, both Eqs. 1 and 3 should be resolved at the same time. Δεδομένου ότι η συγκέντρωση δεν εμφανίζεται στο Eq. 1, δεν θα περιμέναμε καμία επίδραση στην απάντηση h. d) Το δυναμικό μοντέλο για τη χημειοστάτη δίνεται από: 2,14 dX = Vrg − FX dt ή dX ⎛F⎞ = rg − ⎜ ⎟ X dt ⎝V ⎠ (1) Προϊόν: V dP = Vrp − FP dt ή dP ⎛F⎞ = rp − ⎜ ⎟ P dt ⎝V ⎠ (2) Υποστρώματος: V κύτταρα: V dS 1 = F (S f − S) − Vrg dt YX / S ή 1 1 dS ⎛ F ⎞ ⎟(S f − S) − YX / S YP / S dt ⎝ V ⎠ β) Σε σταθερή κατάσταση, 2-16 (3) τότε, dX =0 dt ∴ rg = DX μ X = DX ∴ D= μ (4) Μια απλή στρατηγική ανάδρασης μπορεί να εφαρμοστεί όπου ο ρυθμός ανάπτυξης ελέγχεται με το χειρισμό του ρυθμού ροής μάζας , F, so that the F/V remains stable. c) Flushing occurs if dX/dt is negative for a long time. i.e., rg − DX < 0 or D>m So if the D > m cells will be flushed. d) In a stable state, the dynamic model given by Eqs. 1, 2 and 3 becomes: 0 = rg - DX DX = rg (5) 0 = rp - DP DP = rp (6) 0 = DD SSff − SS − YY 1 XX/SS rrgg (7) from Eq. 5, DX = rg (8) From Eq. 7 rg = Y X / S (S f − S) D (9) Replacement Eq. 9 to Eq. 8, DX = Y X / S (S f − S) D (10) From Eq. 4 S = DK S mmax − D 2-17 Replacing these two equations in Eq. 10, ⎛ DK S = DX YX / S ⎜ S f − ⎜D m D − max ⎝ ⎠ (11) For Yx/s = 0.5, Sf = 10, Ks = 1, X = 2.75, μmax = 0.2, the following plot can be produced based on Eq. 11. Figure S2.14. Steady-state DX cell production rate as a function of dilution rate D. From figure S2.14, leaching occurs in D = 0.18 h-1, while maximum output is displayed in D = 0.14 h-1. Notice that maximum and rinsing points are dangerously close to each other, so special attention should be taken when increasing cell productivity by increasing the dilution rate. 2-18 2.15 a) We can assume that r and h are about constant. The dynamic model is given by: rd = − dM = kAc s dt (1) ότι: M = ρV ∴ dM dV dV dt dt (2) dr d DV = (2p) =A dt dt dt Substitution (3) in (2) and then in (1), V = pr 2 h − rA ∴ dr = kAc s dt −r ∴ (3) dr = kc s dt Integration, r ∫r dr = − o kcs t dt r ∫0 r (t) = ro − ∴ kc s t r (4) Finally, M = rV = rxr 2 and then kc ⎛ M (t) = r⎜⎜ ro − s r ⎝ b) t ⎟⎟ ⎠ 2 The time required to reduce the radius of the pill r by 90% is given by Eq. 4: 0.1ro = ro − kc s t r ∴ Therefore, t = 54 minutes . 2-19 t= 0.9f r (0.9)(0.4)(1.2) = = 54 minutes kc s (0.016)(0.5) 2.16 For V = constant and F = 0, the simplified dynamic model is: S dX = rg = μ max X Ks + S dt S dP X = rp = YP / X m max Ks + S dt 1 1 dS rP =− rg − YP / X dt YX / S Substituting numeric values: dX SX = 0.2 dt 1+ S dP SX = (0.2)(0,2) 1+ S dt dS SX = 0.2 dt 1+ S • 1 0.2 o ⎟⎢− 0.5 − 0.1 o Using MATLAB, this system of differential equations can be solved. The time to achieve a conversion of 90% of S is t = 22.15 h. Figure S2.16. Dynamic fed-batch bioreat reaction behavior. 2-20 2.17 (a) Using a simple volume balance, for the system when the drain is closed (q = 0) AA ddd ddd = qq1 (1) Resolving this ODEON with the given initial state gives a height that increases at a rate of 0.25 ft/min. suppose a time constant in a linear 3-minute transfer mode, so that a constant state is essentially achieved. (3 < t < 18). Suppose the process returns to its previous stable state exponentially, reaching 63.2% of the response in three minutes. 2-21 (c) the input rate is doubled for 6 minutes (18 < t < 24) The height should be increased exponentially towards a new constant state value twice that of the constant state value in Part b), but it should be obvious that the height does not reach this new constant state value in t = 24 minutes.. The new steady state will be 1 ft. (d) the input rate is returned to its original value for 16 minutes (24 < t < 40) 2-22 The graph should show an exponential decrease in the previous constant state of 0.5 ft. The starting value should coincide with the final value from part c). Putting all the charts together will look like this: 2-23 2.18 Parameters (determined by the design process): m, C, me, Ce, he, Ae. CVs: TV and Te. Input variables (disorder): w, Ti. Input variables (manipulated): Q. Degrees of freedom = (11-6) (number of variables) – 2 (number of equations) = 3 The three input variables (w, Ti, Q) are assigned and the resulting system has zero degrees of freedom. 2-24 2.19 (a) First we simulate a step change in the steam flow rate from 0.033 to 0.045 m3/s. The resulting plots of xD and xB are presented below. Plot xD, xB and V versus time to change step to V from 0.033 to 0.045 m3/s. Looking at the resulting data, we can find the constant state values of xD xD xB before and after the step change in V. Start End Change xD 0.85 0.73 -0.12 xB 0.15 0.0050 -0.145 (b) Next we simulate a step change in feed composition (zF) from 0.5 to 0.55. Note that the steam flow rate, V, is still set to 0.045 m3/s. 2-25 Figure: Plot of xD, xB, and zF versus time for a step change in zF from 0.5 to 0.55 Looking at the resulting data, we can find the constant state values of xD and xB before and after the step change in zF. Start End Change xD 0.73 0.80 +0.066 xB 0.0050 0.0068 +0.0018 (c) Increase V causes xD and xB to decrease, while zF growth causes both xD and xB to increase. The size of the result is larger for the V change than for the zF change. When you change V, xB changes faster than xD. 2-26 2.20 (a) First we simulate a step change in fuel gas purity (FG_pur) from 1 to 0.95. The resulting oxygen output concentration (C_O2) and hydrocarbon output temperature (T_HC) are shown below. Picture: Plot C_O2, T_HC and FG_pur for a step change in FG_pur from 1 to 0.95. By looking at the resulting data, we can find the constant state values of C_O2 and T_HC before and after the step change in the FG_pur. Start Final Change C_O2 0.92 1.06 0.14 T_HC 609 595 -14 (b) Then simulate a gradual change in the hydrocarbon flow rate (F_HC_sp) from 0.035 to 0.0385. Note that the fuel gas purity, FG_pur, is still set at 0.95. 2-27 Figure: Plot of C_O2, T_HC, and F_HC_sp versus time for a step change in F_HC_sp from .035 to 0.0385. Looking at the resulting data, we can find the constant state values of C_O2 and T_HC before and after the step change in F_HC_sp. Start Final Change C_O2 1.06 1.06 0 T_HC 595 572 -23 (c) Decrease FG_pur causes C_O2 to increase while T_HC decreases. The increase in F_HC_sp causes T_HC to decrease while C_O2 remains the same. The change in T_HC occurs faster when you change F_HC_sp than when you change FG_pur. 2-28 2.21 The key to this problem is to solve the mass balance of the tank in each part. Mass balance: d (r Ah =) r qi − r qo dt - r (density) and A (tank cross section) are constant, therefore: dh A = qi − qo dt - The problem determines qo is linearly related to the height of the tank qo = 1 h R dh 1 A = qi − h dt R - Next, we can receive R (valve constant) from the steady state information in the problem dh = 0 at constant state dt 0= qi − 0= 2 − ∴ - 1 h R 1 (1) R 1 = 2 R = 0.5 R ft 2 minutes In addition, we can find that 2 -29 = = t AR (4) ⎛⎜ =⎜ 1 2 minutes ⎝2⎠ Part A dh = qi − qo dt (Mass Balance) 4 dh =2 dt (Separate ODEON) 1 ∫ dh = ∫ 2 dt 1 h(t) = t +C 2 h(t) = 1 t +1 2 h(0) = 1 0≤t 0 Taking transformation of the input function, a constant, gives Ci (s) • ci s so that sVC (s) + qC (s) = q ci s or C (s) = Split numerator and denominator with q (s) = ci V os o1 s • q use transformation pair #3 #3 =V/q) 3-21 qci (sv) q)s q − t c(t) = c i • 1 − e V • Using matlab, the concentration response is shown in Fig. S3.18. (See point V = 2 m3, ci=50 Kg/m3 and q = 0.4 m3/min) 50 45 40 35 c(t) 30 25 20 15 10 5 0 0 5 10 15 20 25 30 Time scheme S3.18. Reactor sewage current concentration response. 3.19 a) Take the Laplace conversion of each term, bearing in mind that all initial conditions are zero: s 2Y − sy (0) − y '(0) o 5SY − 5 y (0) o Y • 8sU − 8u (0) u s 2Y • 5sY o Y o U (8s o 1) U (s) • 1 s Y (s 2 o 5s o 1) • Y • 8s o 1 s 8s o 1 s (s 5s 1) 2 Now they use symbolic mathematical software (ex. Mathematica) to resolve for y(t). ReverseLaplaceTransform[(8*s+1)/(s*(s^2+5*s+1)), s,t] 3-22 5- 21 5- 21 − t − t 1 2 g[t_]:= • −21e − 11 21e 2 − 21e 41 o o 42 Plot[g[t],{t ,0,100},AxesLabel,{time,Y},PlotRange,{0,100},{0 ,2}]] 21 −5 t 2 11 21e 21 −5 t 2 y 4 3 2 1 t 0 0 2 4 6 8 10 Figure S3.19a: Tank level response to change in unit flow rate. b) Specify the time at which y(t) reaches the maximum amount as tmax. This moment occurs when y′(t)=0. Solve this time using Mathematica and find that tmax = 0.877 and y (tmax)=1.558. Therefore, the tank will not overflow. (c) Now find the general solution for any input step size, M (the solution is stated in this case as YM(s) and yM(t) for clarity). The input U is M/s. U(s) • M s YM (s 2 o 5s 1) • YM • M (8s 1) s M (8s 1) • MY s (s 2 o 5s 1) YM is the previous Y, multiplied by the size of the step, M. M. M. Since M is a constant, taking the inverse Transformation Laplace gives: yM (t) • My(t) Now solve the equation: 3-23 yM (tmax) • 2.5 • me (tmax) • M (1,558) M • 2,5 • 1,605 1,558 The maximum step change in the flow rate in the tank that will not overflow the tank is 1,605. y 4 3 2 1 t 0 0 2 4 6 8 10 Figure S3.19b: Tank level response to a change in flow rate of 1,605 steps. 3.20 (a) Given the fixed volumes, the total balances in the three tanks indicate that the flow rate from each tank is equal to the remaining q For the tracer in each tank; V1 dc1 • q (ci − c1) dt V2 dc2 • q (c1 − c 2) dt V3 dc3 • q (c2 − c3) dt 3-24 b) The laplace conversion of the above equations and the elimination of C1 (s) and C2 (s) gives q q q q • q / V • • 1 o q / V2 o o V3 C (s) o s q / V1 os o q / V2 os o q / V3 • i From the ci (t) • • (t) 1. Ci(s) • 1 V1 • V2 • V3 • V • 3 • 1 • 2 (q / V) 3 C3 (s) • o 3 2 (s o q / V) (s) q / V) (s (q / V) (s) q / V)3 c3 (t) • • 1e](q /V)t o • 2te](q /V)t o o 3t 2e−(q /V)t 2. V1 μ V2 μ V3 μ V1 c3 (t) • 4e (c) − (q /V1)t o o 5e −(q /V2)t o o 6e −(q /V3)t Yes, tracer can be calculated by measuring c3 (t), tracer quantity = qc3 (t)dt , which can be evaluated numerically 0 3.21 Start with the Laplace version of equations from exercise 13.20: q q q • • V1 • V2 • V3 C3 (s) • C (s) o q / V1 os o q / V2 os o q / V3 o i From V1=V2=V3, this equation is reduced to: 3 q • • • V C3 (s) • • 3 Ci (s) os o q /V o (1) 3-25, where ci(t) is a pulse of size A and width tw. A pulse can be described by the sum of the two step functions. The first will be a step function of size A at the time 0. The second will be a step function of -A at t = tw. ci i i t) • AS (t) − AS (t − t w) 1 − e − tw s A − tw s − e • A • (2) s • s Now substitute Equation (2) in equation (1). For simplicity's sake, set a new variable f=q/V. Ci (s) • A o f o (1 − e−tw s) 3 C3 (s) • s os o f o Now use a symbolic math software to find the reverse Laplace, giving c3(t). The solution is formulated as a function of t, f, A and tw. Then, for example, we design c3(t) for f=1/20, A=10, and tw=1. 3 In Mathematics, take the inverse Laplace transformation: ReverseLaplaceTransform[A ∗ f^3 ∗ (1 − Exp[−tw ∗ s])/(s ∗ (s + f)^3) , t] The solution: c3(t) = 1/2 A(e −ft (−f 2 + 2e ft − 2ft − f 2 t 2) − e(−t+tw) (−2 + 2e f(t−tw) − 2f(t − tw) − f 2 (t − tw)2)HeavisideThes[t − tw]) Set the function by parameters: g[t_, f_, A_, tw_] 1 ≔ A(e −ft (−2 + 2e ft − 2ft − f 2 t 2) − e f(−t+tw) (−2 2 + 2e f(t−tw) − 2f(t − tw) − f 2 (t − tw)2)HeavisideTheta[t − tw]) , draw the concentration over time, assuming f=1/20, A=10, and tw=1. Design[g[t, 1/20, 10.1], {t, 0.200}, AxesLabel → {hour, C3}] 3-26 C3 0.12 0.10 0.08 0.06 0.04 0.02 time 50 100 150 200 Figure S3.21: Diagram c3 over time in response to pulse in ci width 10 and width 1, with f=1/20. 3.22 Solve this problem by using a symbolic software program such as Mathematica. The following scenario will solve the problem (note that only 4 of the 5 possible initial conditions for y and its derivatives are included, otherwise the problem is overspecified). DSolve[{y''[x] + 16 ∗ y'''[x] + 86 ∗ y''[x] + 176 ∗ y′[x] + 105 ∗ y[x] = = 1, y[0] == 0, y′[0] == 0, y''[0] == 0, y'''[0] == 0}, y[x], x] Running this script will give the result: {{y[x] → e −7x (−1 + e x)4 (5 + 20e x + 29e 2x + 16e 3x) }} 1680 Use the Expand[ ] command to extend this solution to its individual terms. 1 e e −5x e −3x e −x {{y[x] → + − − }} 105 336 80 48 48 If desired, fractions can be accessed as decimal places: y(t) • 0.003e −7t − 0.0125 e −5t 0.021e −3t − 0.021e −t 0.01 3-27 Chapter 4 © 4.1 Y(s) d • U (s) bs c a) K gain can be achieved by setting s = 0 d K • c Alternatively, the transport function can be placed in the standard fixed form of profit/time by dividing the numerator and denominator by c: d b Y (s) K , where K • and o . c u(s) os o 1 b) To determine the delimitation of the output response, consider a gradual input of size M. Then use U (S) = M/s and Y (s) • K M • s From Table 3.1, the step response is: y(t) • KM (1 − e--t /)) From inspection, this response will be delimited only if > 0, or equivalently, only if b/c > 0. 4.2 (a) (b) (c) K=3 o=10 We use the Final Value Theorem to find the value of y(t) when t-. 12e − s Y (s) s (10 s 1) sY (s) • 12e − s (10 s 1) 12e − s • 12 s 0 (10 s 1) lim From the Final Value Theorem, y(t) = 12 when t- d) [Type here] y(t) = 12(1--e](t-1)/10), then y(10) = 12(1-e-9/10) = 7.12 4-1 [Type here] 7.12/12=0.593. e) Reuse the final value theorem. 3e s (1 e s) Y (s) (10 s 1) s 3e s (1 e s) sY (s) (10 s 1) 3e s (1 e s) 3(1 1) lim 0 s 0 (10 s 1) 1 From the Final Value Theorem, y(t) = 0 when t f) 3e s Y (s) 1 (10 s 1) 3se s sY (s) (10 s 1) 3se s lim 0 s 0 (10 s 1) From the Final Value Theorem, y(t)= 0 when t g) Y (s) 3e s 10 30e s (10 s 1) (s 2 4) (10 s 1)(s 2 4) then 10 (10t 1) 1 y (t) 30 S (t 1) e (sin(2(t 1)) 20 cos(2(t − 1))) 802 • 401 The sinusoidal input produces a sinusoidal output and y(t) has no limit when t-. These solutions can be verified using mathematical software such as Mathematica or Simulink. 4-2 y 12 y 10 0.25 8 0.20 6 0.15 4 0.10 2 0.05 10 20 30 40 50 time 10 Fig. S4.2a. Exit for Parts (c) and (d). 20 30 40 50 times Fig. S4.2b. Production for part e). y 0.30 y 0.25 2 0.20 0.15 1 0.10 0.05 10 10 20 30 40 50 20 30 40 50 50 time 1 Fig. S4.2c. Production for part f). Image. S4.2d. Production for part g). 4.3 The transfer function for the pressure transmitter is given by, Pm(s) 1 • P (s) 10s 1 (1) and P(s) • 15 / s to change step from 35 to 50 psi. The replacement (1) and rearrangement gives: 1 15 m (s) • 10s o 1 s From point #13 in table 3.1, the step answer is given by: Pm (t) • 15 (1 − e --t / 10) (2) Let ta be the moment when the sound alarm. Then pm (ta) • 45 − 35 • 10 psi (3) Replacement (3) and t = ta in (2) and resolution gives ta = 11s. So the alarm will sound 11 after 1:30 p.m. 4-3 4.4 From Exercise 4.2, Y(s) 3e − s • U(s) 10s 1 Rearrangement, Y(s) [10s 1] • 3e − sU(s) (2) Take his L-1 (2), 10 dy o y • 3 u (t − 1) dt (3) Take his L (3) for y (0)=4, 10[sY (s) − 4] o Y(s) • 3e − sU (s) Substitute U(s) • 2 / s and rearrange to give, 10sY- 40 +Y= Y (10 s 1) • 6 − s 6e − s • 40 s Partial fraction extension: Y(s) • ea-- s 6 40 s (10 s 1) (10 s 1) a2 6 • 1 s (10 s) s 10 s 1 Find 1 : Multiplies by s and set • 0 • • 1 • 6 Find • 2 : Multiplies by 10s o 1 and sets • −0.1 • 2 • 60 6 4 6 Y (s) • e o o . 0.1 (s) 0.1) Take L-1 , y (t) • 6 S (t − 1)(1) e − (t −1)/10) o 4e − t /10 Control: T =0, y (0)=4. 4-4 4.5 a) dy1 = -2y1 – 3y2 + 2u1 dt dy 2 = 4y1 – 6y2 + 2u1 + 4u2 dt 2 (1) (2) Taking Laplace convert the above equations and rearrange, (2s+2)Y1(s) + 3Y2(s) = 2U1(s) (3) -4 Y1(s) + (s+6)Y2(s)=2U1(s) + 4U2(s) (4) Resolve Eqs. (3) and (4) simultaneously for Y1(s) and Y2(s), Y1(s) = (2s) 6)U1(s) − 12U 2(s) 2(s) 2(s) 3)U1 (s) (s) (s) − 12U 2(s) 2(s) 2(s) 2(s) 2(s) 2(s) 2(s) 3)U1(s) 2(s) 2(s) 2(s) 2 2(s) 2(s) 2(s) 2(s) 2(s) 2(s) 3)U1(s) − 12U 2(s) 2(s) 2(s) 3)(s) 4) 2s 2 o 14s o 24 Y2(s) = (4s 12)U1 (s) − (8s 8)U 2 (s) 4(s) 3)U1(s) 8(s o 1)U 2 (s) • 2(s) 3)(s 4) 2s 2s 2 os 14s o 24 Therefore, Y1 (s) 1 o U1 (s) s 4 , Y1 (s) −6 • U 2 (s) n. 3)(s 4) Y2 (s) 2 • U1 (s) s 4, Y2 (s) 4(s) 1) • U 2 (s) (s) 3)(s) 4.6 a) Taking the L-1 gives, x (t) • 0.09e− t /10 and x(t) • x x x (t) • 0.3 o 0.09e − t /10 The int values x (0) • 0.09 and x(0) • x (0) o x • 0.09 • 0.3 • 0.39. The plot of the concentration response appears in the Shape. S4.6. 4-5 x t Fig. S4.6. Transient response. The transfer function is given by: X (s) 0.6 • X i(s) 10s 1 For push input, xi(t) • 1.5(t) and table 3.1, X i(s)=1.5 . Thus, 0.9 x • 10 s 1 b) Home Value Theorem: 0.9 • 0.09 10 So, x(0) • x(0) • x • 0.09 o 0.3 • 0.39 c) For the steady state, x(0) • lim sX • s o x(0) • x • 0.3 d) As stated in the plot, the push response is discontinuous in t=0. The results for parts (a) and (b) give the values of x(0) for t=0+ while the result for point (c) gives the value for t=0-. 4.7 The simplified stage concentration model becomes dx1 • L(x 0 − x1) v (y 2 − y1) dt y1 = a0 + a1x1 + a2x12 +a3x13 H 4-6 (1) (2) a) Allow the right side of Eq. 1 to be marked as f(L, x0, x1, V, y1, y2) H dx1 f f • f (L, x0 , x1, V, y1 , y 2) • L o o o dt • • x0 f x0 o o o x1 • x1 f f f 2 V y y V 1 2 s Αντικαθιστώντας τα μερικά παράγωγα και σημειώνοντας ότι H dx1 dx1 : dt dt dx1 (x0 x1) L L x0 L x1 (y 2 y1)V V y 2 V y1 dt (3) Ομοίως, g x1 (a1 2a 2a 2a 2 x1 3a3 x1 2) x1 y1 g (x1) x1 s b) (4) Για σταθερούς ρυθμούς ροής υγρών και ατμών, L V 0 Λαμβάνοντας Laplace μετατρέπει eqs. 3 και 4, HsX 1 (ες) L X 0 (ες) L X 1 (ες) V Y2 (ες) V Y1 (ες) (ες) (5) Y1(-ες) (a1 2a2 x1 3a3 x1) X 1(-α) (6) 2 Από Eqs. 5 and 6, the desired transfer functions are: L X 1 (s) H X 0 (s) s 1 Y1(s) X 0 (s) Y1 (s) Y2 (s) V X 1 (s) H Y2 (s) s 1 , (a1 2a 2 x1 3a3 x1) 2 s 1 (a1 2a 2 x1 3a3 x1) 2 s 1 L H V H where: H L V (a1 2a 2 x1 3a3 x1) 2 4-7 4.8 The material balance is, d (Ah) wi Rh 1.5 dt or dh 1 R 1.5 wi h dt A A Use a Taylor series expansion to linearize dh 1 R 1.5 1 1.5Rh 0.5 wi h (wi wi) (h h) dt A A A A Since the bracketed term is identically zero at steady state , dh 1 1.5 Rh 0.5 wi h dt A A Rearranging Thus, A dh 1 h wi 0.5 dt 1.5Rh 1.5Rh 0.5 H (s) K Wi(s) s 1 where, K height 1 h h 0.5 1.5 1.5w flowrate 1.5Rh 1.5Rh mass time A Ah V 0.5 1.5 1.5w mass / time 1.5Rh 1.5Rh 4-8 4.9 a) The model for the system is given by dT wC (Ti T) h p A p (Tw T) (2-51) dt dT mw C w w hs As (Ts Tw) h p A p (Tw T) (2-52) dt Assume that m , mw, C, Cw, hp, hs, Ap, As, και w είναι σταθερές. Αναδιατύπωση των ανωτέρω εξισώσεων από την άποψη των μεταβλητών απόκλισης και σημειώνοντας ότι mC dTw dTw dT dT , dt dt dt dt dt dT mC wC (Ti T) h p A (Tw T) dt dt dT m w w w hs As (0 Tw) h p A p (Tw T) dt Λήψη μετασχηματισμών και αναδιάταξης του Λακ. (mCs wC h p Ap) T wCTi(s) h p Ap Tw (s) (ες) 1) (mw C w s hs As h p Ap)Tw(s) h p Ap T (s) (2) Υποκαθιστώντας στο Eq. 1 για Tw (ες) από Eq. 2 , (mCs wC h p Ap)T (s) wCTi(s) h p Ap h p P Ap (mw C w s hs As h p A p) T (s) Ως εκ τούτου, wC (m w C w s hs Όπως h p A p) T (s) Ti(s) (mCs wC h p A p)(m w C w s hs Όπως h p A p) (h p P p) 2 β) γ) wC (hs Όπως p A p) T (-ες) Το κέρδος είναι Ti(s) s 0 wC (hs As h p A) hs As h p Ap , το κέρδος αναμένεται να είναι ένα if the tank was insulated so that hpAp=0. For the heated tank, the gain is not one, because the heat input changes as the T. 4-9 4.10 Changes Additional Assumptions 1. Perfect mixing in tank 2. Fixed density and special heat C. 3. What's stable? Energy balance for the tank, • VC dT • wc (Ti − T) o Q − (U o bv 2) A(T − Ta) dt Let the right side be declared with f(T,v), o VC dT f f • f (T , v) • • T v dt • T s • v s (1) f 2 • − wC − (U) bv) A T s f • • 2vva(T − Ta) • v replaces the few derivatives in the Eq. 1 and noting that the wc (U bv 2) A T – 2vva(T – Ta) v dt dT ρVC (U o bv 2) A T – 2vva(T – Ta) v dt dT Receiving the Laplace transformation and reordering · VC · VCT wc (U bv 2) A T 2vva(T Ta) V VCs wc (U bv 2) A 2vva(T Ta) T wc (U o bv 2) A s o 1 4-10 dT dT = dt dt 4.11 a) Mass balances for increasing tanks: dm1 · w1 – w2 dt dm 2 · w2 – w3 2 dt (1) (2) Ideal gas law: m1 RT M m P2V2 · 2 RT M P1V1 · Flows (Ohm law : I · (3) (4) E Driving Force · R Resistance 1 (Pc – P1) R1 1 w2 · (P1 – P2) R2 1 w3 (P2 – Ph) R3 w1 (5) (6) (7) Degrees of freedom: 11 number of parameters : 8 (V1, V2, M, R, T, R1, R2, R3) number of variables : 9 (m1, m2, w1, w2, w3, P1, P2, Pc, Ph) number of equations : 7 Ø number of degrees of freedom to be eliminated = 9 – 7 = 2 Because pc and Ph are known functions of time (i.e. , inputs), NF = 0. b) Development Model M1 RT M V2 Substitute (1) on (3) : RT Substitute (3) on (1) : dP1 · w1 – w2 dt dP2 · w2 – w3 dt 4-11 (8) (9) Substitute (5) and (6) in (8): MV1 d P1 1 1 · (Pc – P1) – (P1 – P2) RT dt R1 R2 MV1 dP1 1 1 1 · Pc (t) – ( ) P1) P2 RT dt R1 R1 R2 R2 (10) Similarly , substitute (6) and (7) in (9) : MV2 dP2 1 1 · (P1 – P2) – (P2 – Ph) RT dt R2 R3 MV2 dP2 1 1 1 · P1 – ( ) P2 Ph (t) RT dt R2 R2 R3 R3 (11) dP1 · f1 (P1 , P2) From Eq. 10 dt dP2 · f2 (P1 , P2) From Eq. 11 dt This system has the following features : i) 2nd class denominator (2 differential equations) ii) Zero-series numerator (see example 4.7 in text) W (s) (iii) The gain of 3 is not equal to the module. (It may not be because the pc units are for the two variables are different). Note that 4.12 (a) First write the constant state equations: 0 · wc (Ti – T) (Te – T) 0 · Q – ae (Te – T) Remove the constant state equations from the dynamic equations dT · T (Ti – Ti) – (T T) dt · Te (Te – Te) – (T – T) dt mc dT · Te (T – T) – T dt · DT · dt and dTe / dt · dTe / dt. Substitute mC (1) (2) deviation variables: then multiplied (1) by 1/wC and (2) by 1/(he e Ae). 4-12 hA m dT ' – (T' – Ti ') ee (T'e – T') w dt wC meCe dT 'e Q' – (T'e – T') hA dt he Ae (3) (4) Elimination of one of the output variables, T's or T'e (s), by solving (4) for this, and replacing it in (3). Because T'e(s) is the intermediate variable, remove it. Στη συνέχεια, η αναδιάταξη δίνει: mmeCe 2 meCe meCe m 1 T 'ec Q 's) wC αυτός Ae Επειδή και οι δύο εισροές επηρεάζουν τη δυναμική συμπεριφορά του T%, είναι απαραίτητο να αναπτυχθούν δύο λειτουργίες μεταφοράς για το μοντέλο. The effect of Q' on T' can occur assuming that Ti is stable at the nominal constant state value, Ti . Thus, Ti = 0 and the previous equation can be rearranged as: T'(s) 1/ we · G1 (s) Q '(s) b2 s 2 a1 b1s o 1 (1) o 4.13) Similarly, the effect of T'i on T' is achieved by assuming that Q' = Q (i.e., Q'=0): meCe s 1 o1 Ae T '(s) · G2 (s) T' i (s) b2 s 2 s b1s o 1 (Q 's) · 0) where the b1s o 1 ρ(s). Note 1/ wc · G1 (s) Q (s) X2 G2(s) T'(s) (s) = D(s) Equal to ρ(s) Note that there are two transfer functions that have the same denominator by T'(s) · G1 (s)Q '(s) G2 (s)T'i(s) (b) The restrictive behavior of m eCe that goes to zero has b 2 · 0 and b1 · m / w and simplifies the last equation with T'(s) · 1 / wC 1 Q ' (s) T i (s) m m s μ1 s μ1 w w 4-13 4.13 Mass balance yields: dm · qi – q dt (1) The term mass accumulation can be written, noting that dV=Adh=wtLdh, such as dm dV dh · · · o wt L (2) dt dt dt where the cross-sectional area of the tank is not constant. From geometric considerations, the equation is the length of one side of a right triangle whose hypotenuse is R. Thus, by combining (1) and (2) gives a nonlinear dynamic model for the tank with qi and q as inputs: dh 1 · (qi – q) dt 2 L (D – h)h To line into this equation on the operating point (h), let qi – q f (2 L (D – h) h) the terms of Taylor series ... (3) The linearized model (4) with (5) 4-14 The last partial derivative is zero, because qi – q from the constant state relationship, and the derived term in parentheses is finite for all 0 10s, Tm (t) · 20(1 – exp(−t) – 40(1 – exp(−t(t − 1 – (t − 10)) · 120 Figure S5.27 Tm vs. time 5-37 (b) when t = 0.5 s, Tm (0.5) · 20(1 – exp(−0.5)) o 120 · 127.87F when t = 15 s, Tm (2) · 20 (1 – exp(−2)) o 120 · 100.26F 5.28 a a5 a a4 1 · 1 0 2 2 o 3 o 3 2 s (2 s 1) s o s 1 (1 s 1) (2 s 1) · 2 a2 · lims o 1 o 0 SET ... (6) s (2s 1) ... 5.29 a) Energy balance: ... The time-domain ... These two inputs rate the time-domain response ... (c) Fidelity ... 6-1 (a) ... 6.1 (a) Standard form: G s · 4 o s 2 o 0 o 1 o 2 s 0 o s 1 o 2s 1 o s 0 o 1 (s) o (s o 2) b) Apply zero pole cancellation: 8 G · s · 6 · s · 1 2 s o 1 Gain =8; Pole=0.5? Zeros=None (c) 1/1 Padéapproximation: ... 6.5 Since the transfer function ... the zero ... 6.6 ... Inverse response ... Overshoot ... (no) ... 6.7 ... 6.8 ... 6.9 ... 6.10 (a) The transport function for each case is ... 6-11 ... Simulink ... Figure S6.5. Step system response. 6.6 Y (εc) K1 K (εc) 2 U (εc) 2 U (εc) s 1 2 Y (εc) K1 K1 K2 s (K1 K2 s 2 s) (s) 1 s 6-5 250 300 Βάλτε σε τυπική φόρμα για ανάλυση: ... 1.5 ... (a) X(s) ... 6.7 The transfer functions ... The approximation is 1st class with delay ... 6.8 (a) ... 6.10 ... 6.11 ... (b) Simulink ... Step response for accurate and approximate models. 6-13 50 (c) Maximum error = 0.265, in tr = 9.85s, and the position corresponding to the maximum error is graphically displayed in the image above with a black vertical line. 6.13 From solution to exercise 2.5(a), the dynamic model for isothermal function is V1 M dP1 P4 – P1 P1 – P2 RT1 dt Ra Rb (1) V2 M dP2 P1 – P2 P2 P2 – RT2 dt Rb Rc (2) Receiving Laplace transformation, and noting that P1(εc) · 0 since P1 is stable, K b P(εc) Ka (s) P (εc) K P (s) P2(εc) · 1 s ... Replacing P1(s) from Eq. 5 to 4, ... 6.14 ... T(s) ... 6.15 ... Approximation of the shortest time constant ... 6.16 ... Scan through ... 6-15 ... The longest time constant in (6) for neglect is 1. Thus, G(s) ... Approximation of the model ... 6.17 The entrance is T, (λ) ... The transfer function of time constant ... 6.18 150 148 146 ... 6.19 The mathematical model is derived on the basis of the balance of the material: dc V 1 · F0 c0 o R F2 c2 − F1c1 − Vkc1 dt dc V 2 · F1c1 − 1 o R F2c2 − Vkc2 dt ... The transfer mode can occur on the basis of the Laplace transformation: ... 6-20 ... 6.20 ... (a) ... The profit of the above transport operation is 1. The dynamic model for the process is given by Eqs. 2-45 and 2-46, which can be written as dh 1 (wi − w) dt · A (1) and ... where h is the liquid level, A is the cross-sectional area of the tank. (b) ... The profit ... 6.21 (a) ... The interaction ... 6.22 (a) ... 6-24 Let ... Then ... For the heating process, the heat capacity C of the liquid and the process (a) Considering ... 6.24 (a) ... (c) The temperature of the steam, Ts ... (iii) The liquid temperature response ... Thermal capacity ... 6.25 ... Μέση θερμοκρασία ... Το αποτέλεσμα ... T – q transfer function ... T – Ts ... T – qF ...

is independent of ts and the steam pressure Ps. T − q transfer function is second class due to interaction with the liquid level; is the product of an integrator and a first-class process. T − the qF transfer function is second class due to interaction with the liquid level; has dynamic numerator since qF affects T directly, as well as if TF oT . T − The Ts transfer function is first class because there is no interaction with the liquid level. (c) h − qF : h is constantly increasing at a steady rate. h − q : h is constantly decreasing at a steady rate. h − Ts : h remains stable. T − qF : for TF , T is initially reduced (reverse response) and then increased. After a long time, T grows like a ramp operation. T − q : T decreases, eventually at a steady rate. T − Ts : T increases with a first-class response and achieves a new stable state. 6-27 6.22 a) The two-tank process is described by the following equations in deviation variables: dh1' 1 ' 1 ' ' w1 (h1 h2 dt A1 R dh2' 1 dt A2 1 ' ' R (h1 h2 (1) (2) Laplace transforming A1RsH1' (s) RWi ' (s) H1' (s) H 2' (s) (3) A2 RsH 2' (s) H1' (s) H 2' (s) (4) From (4) (A2 Rs 1) H 2' (s) H1' (s) (5) H 2' (s) 1 1 ' H1 (s) A2 Rs 1 2 s 1 (6) or where 2 A2 R Returning to (3) (A1Rs 1) H1' (s) H 2' (s) RWi ' (s) (7) Substituting (6) with 1 A1 R 1 ' ' (1s 1) H1 (s) RWi ' (s) H 2' (s) 1 or 6-28 (8) (1 2) s 2 (1 2 s 1) s H1' (s) R(2 s 1)Wi ' (s) H1' (s) R(2 s 1) W1 ' (s) s 1 2 (1 2) s (9) (10) Dividing numerator and denominator by (1 2) to put into standard form H1' (s) [R /(1 2)] (2 s 1) W1' (s) 1 2 s 1 2 (11) Note that K R R 1 1 1 2 A1R A2 R (A1 A2) A (12) since A A1 A2 Also, let s 1 2 R 2 R 2 A1 A2 RA1 A2 1 2 R (A1 A2) A (13) so that H1' (s) K (2 s 1) Wi ' (s) s(3 s 1) (14) and H 2' (s) H 2' (s) H1' (s) K (2 s The transport functions (6), (14) and (15) determine the operation of the process of the two tanks. The process of a tank is from the following equation in the deviation variables: 6-29 dh' 1' · wi dt · A (16) Note that · , which is constant , removes. Laplace transformation and rearrangement: H ' (s) 1 / oA · Wi ' (s) s (17) Again K · 1 oA H ' (s) K Wi ' (s) s (18), which is the expected integrated relationship without zero. b) For A1 · A2 · A / 2 · 2 · AR / 2 · · · 3 · AR / 4 · (19) So · 2 · 2 · 3 We have two sets of transport functions: One-Tank Process of Two Tanks H ' (s) K · Wi Wi ' (s) (2)3 s o1) · Wi ' (s) s(3 s) H1' (s) K · ' Wi (s) s(3 s) Remarks: - Profit (K · 1/ oA) is the same for all TFs. - Each TF contains an embed element. 6-30 - However, the two TF tanks contain a pole (3 s 1) that will filter changes in the level caused by the change wi(t). - On the other hand, for this special case, we see that zero in the first tank transfer mode (H i' (s) / Wi' (s)) is greater than the pole: 2 · 3 · 3 So we need to ensure that the reinforcement of changes in h1(t) caused by zero is no more than to cancel the beneficial filtering of the pole, so as to cause the first apartment to overflow easily. Now consider the general situations of the case of the two tanks: H1' (s) K (-A2 Rs) K ()2 s o1) · ' Wi (s) o RA1 A2 s(3 s) s 1 · A o 1 · A o 1 · A o 1 · A o 1 · A o 1 · A o 1 · A o 1 · A o 1 · A o 1 · A o 1 · A o 1 · 1 · A o 1 2 (s) K · ' Wi(s) s(3 s) 1) (21) For A1 o 0 or A2 o 0 This loses the beneficial effect of the pole as the process tends to be more similar to the first-class process. · 3 · c) Optimal filtering can be found by maximizing · 3 in relation to A1 (or A2) · 3 · RA1 A2 · RA1 (A − A1) · A Find max · 3 : Setting to 0: · 3 R · a − A1) a1 (--1) · A1 A − A1 − A1 · 0 2 A1 A A1 A / 2 Thus, the maximum filtering action is achieved when A1 · A2 · A / 2. 6-31 The ratio of · 2 / · 3 determines the aid effect of zero to h1 (t). As a result, the effect of changes in wi (t) for h1 (t) will be very large, resulting in the possibility of overflow in the first tank. To sum up: The process designer would like to have A1 · A2 · A / 2 in order to achieve maximum filtering of h1 (t) and h2 (t). However, the process response should be checked for standard changes to the wi (t) to make sure that h1 does not overflow. If it does, the A1 area should be increased until it is not a problem. Note that · 2 · 3 when A1 · A, so careful study (simulations) should be done before designing the separator tank. Otherwise, leave it quiet and use the non-separated tank. 6.23 The process transfer function is Y(s) K · G(s) · 2 U (s) (0.1s 1) (4s 2) 2s where K = K1K2. The square term describes a 2nd-class system that has not been sampled by · 2 · 4 · 2 o 6-32 2 · · 2 (a) o · 0.5 For the element of the second order procedure with · 2 = 2 and this degree of sampling (· 0.5) , the short time constant, the element of the 2nd series process (-1 = 0.1) will have little effect. Actually, from 0.10. System II (air-closing valve): as the signal in the control valve increases, the flow through the valve decreases Rv < 0. 8-4 b) System I: The flow rate is very high - need to close the valve reduce controller output reverse operation controller Or: Process gain + Valve gain + Controller gain must be + (meaning reverse action) System II: Flow rate too high - need to close valve growth controller output direct acting controller. Or: c) Process Profit + Valve Profit – Controller Profit must be – (meaning direct operation) System I: Kc > 0. System II: Kc < 0. 8.5 (a) From Eqs. 8-1 and 8-2, p(t) · p o K c y sp (t) − y m (t) o (1) The liquid transmitter ratio is ym(t) = KT h(t) (2) where: h is the liquid plane KT > 0 is the gain of the direct action transmitter. The control-valve ratio is q(t) = Kvp(t) (3) where q is the manipulated flow rate Lv is the gain of the control valve. (a) Configuration (a) in the Figure. E8.5: As h increases, we want to reduce qi, the rate of input flow. For an air-closing control valve, the output p controller must be reduced. Thus as p increase reduces the direct-acting controller. 8-5 Configuration (b): As h increases, we want to increase the q, output flow rate. For an air-closing control valve, the output of the controller should be reduced. Thus since the x p increase reduces a reverse action controller. b) Configuration (a) in the Figure. E8.5: As h increases, we want to reduce qi, the rate of input flow. For an air-opening control valve, the output p controller must be reduced. Thus since the x p increase reduces a reverse action controller. Configuration (b): As h increases, we want to increase the q, the output flow rate. For an air-to-open control valve, the output of the controller should be increased. Thus as the p increase increases a direct-acting controller. 8.6 For the control PI t 1 p(t) · p o K c o(t) o e(t*)dt * · I 0 · t t · 1 o p(t) · K c ◊ e(t) o e(t*)dt * · I 0 · From e(t) = ysp – ym = 0 - 2 = - 2 Then , t 1 2 p (t) · K c · − 2 o (--2)dt * · K c · − 2 − · I 0 · I · t The initial response to t = 0 is − 2 Kc 2K c · Akata and the numeric values of the initial response and inclination from the Fig. E8.6: The inclination of the response is − - 2 Kc = 6 − 2K c = 1.2 min-1 Kc = -3 · I = 5 minutes 8-6 8.7 (a) The error signal can be described by: e(t) · 0.5t E(s) · 0.5 50 50 The PID controller transfer function is given by (Eq. 8-14): P' (s) 1 · K C · 1 o o D s E(s) · It is The replacement provides the controller with the output: 1 o Ds · 1 o is 1 2 p '(t) · 0.1 The replacement of numeric values and the addition of p · 12 mA gives: 1 pPID (t) · 12 t 2 t 2 t 0.5S (t) 3 (b) The equation for a PI controller is achieved by setting the tD to zero. 1 PPI (t) · 12 t 2 t 3 (c) The design of the controller response for both controllers is shown in the Figure. S8.7. The two auditors have similar answers. The difference is the sudden transition to t=0 that occurs with the PID controller as a result of the derivatives condition. When the specified point begins to change with a constant gradient, there is a step change in the error derivative from 0 to 0.5. The derived term on the controller gives it a start-up right when the adjustment point starts to change that the PID controller does not have. 8-7 Figure S8.7: PID Controller Output Response 8.8 From Eq. 8-25 Inspection, the derivative kick = K c a) Analog kick = K c r b) e1 = e2 = e3 = ... = ek-2 = ek-1 = 0 ek = ek+1 = ek+2 = ...= r p k −1 · p o t p k · p o K c · r o r o D r oP · 8-8 · D t t p k oi o p o k c · r o (1 o i) i · Kc p 2, ... · D r t K c r Kc p k-1 , k t r · I k+1 k+2 k+3 a) To eliminate the derivative kick, replace (ek – ek-1) in Eq. 8-25 by - (YK-Yk-1). (Note the minus sign.) 8.9 (a) Allow the fixed adjustment point to be indicated with y sp . The digital speed algorithm P is obtained with the setting of 1/-I = D = 0 in Eq. 8-27: pk = Kc (ek – ek-1) = K c (y sp − y k) − y sp − y k −1 ¦ = K c y k −1 − y k · The PD digital speed algorithm is obtained with a setting of 1/ · I = 0 in Eq. 827: pk = Kc [(ek – ek-1) + D (ek – 2ek-1 + ek-2)] t · = Kc [(-2 yk + yk-1) + D (-yk – 2yk-1 + yk-2)] t 8-9 In both cases, pk does not depend on y sp . b) For both of these algorithms pk = 0 if yk-2 = yk-1 = yk. This achieves a constant state with a y value that is independent of the value of y sp . The use of these control algorithms is not recommended if displacement is a concern. c) If the built-in function exists, then pk contains the term Kc a) · D s P 1 · K o o1 o E (s) o s os o 1 I D · t (y − sp y k). So, in stable condition, pk = 0 and yk-2 = yk-1 = yk , yk = y sp , and the displacement problem has been eliminated. 8.10 Kc s(D s 1) D s 1 D s s s(D s 1) 1 (I D) s (1) D s 2 Kc I s (D s 1) πολλαπλασιάζοντας (I D s 2 I s) P (εc) K γ 1 (D) s (1) D s 2 E (s) Λαμβάνοντας αντίστροφα laplace μετασχηματίζει, δ 2 ε τ) δ 2 p (t) dp (t) de(t) (1) D D K c e(t) (I D) dt dd 2 dt 2 dt · b) o 1 o D s P (s) · · K c ◊ I E (s) · I s · d s o 1 -multiplying · I s 2 (D s) 1) P (s) · K c o(i) 1)(d s 1) o E (s) 8-10 o o o d c) d 2 e(t) d 2 p (t) dp (t) de(t) K e (t) ( ) I D I c I D dt dt 2 dt 2 Η προσομοίωση εκτελείται για τις ακόλουθες τιμές παραμέτρων: Kc 2 , 3 , D 0,5 , 0,1 , M = 1 Τα αποτελέσματα Simulink-MATLAB εμφανίζονται στο σχήμα S8.10. : Step Answer 22 Parallel PID with Derivative Filter PID Series with Derivative Filter 20 18 16 14 p'(t) 12 10 8 6 4 2 0 2 4 6 8 10 Time Figure S8.10. Response step for parallel PID controllers and for series PID controllers with a derivative filter. 8.11 The integral part of the auditor's action is determined by incorporating the error between the measurement and the specified point over time. As long as the error symbol remains the same (that is, if the measurement does not cross the specified point), the embedded element will continue to change monotonously. If the measurement passes the specified point, the error term will change the symbol and the embedded element will start changing in the other direction. So it will no longer be monotonous. 8-11 8.12 a) False. The controller output can be certified or the controller may be in manual mode. (b) False. Even with the integer control action, the offset can occur if the controller output is certified. Or the controller may be in manual mode. 8.13 First consider qualitatively how h2 responds to a change in the 2nd trimester. From natural estimates, it is clear that if q2 increases, h2 will increase. So if h2 increases, we want q2 to decrease, and vice versa. Since the control valve q2 is aerial, the output p level controller should be reduced in order to reduce the q2. In short, if h2 increases we want p to decrease; therefore, a reverse action controller is required. 8.14 Consider first qualitatively how the soluble mass fraction x responds to a change in the rate of steam flow, S. From natural estimates, it is clear that if the S increases, the x will also increase. So if x grows, we want S to drop, and vice versa. For a failopen (air-near) control valve, the output of p controllers must be increased in order to have the reduction S. In short, if x increases we want S to decrease, which requires an increase in the controller output p; therefore, an immediate action controller is required. 8.15 First qualitatively how the output temperature Th2 responds to a change in the flow rate of cooling water, wc. From natural estimates, it is clear that if wc decreases, Th2 will increase. So if Th2 decreases, we want wc to decrease, and vice versa. But to determine the controller's action, we need to know if the control valve is open or fails nearby. Based on safety reasons, the control valve should fail open (from air to close). Otherwise, the very hot liquid current could become even warmer and cause problems (e.g., bursting the tube or creating a two-phase flow). For an air-closing control valve, the temperature controller p output must be increased in order to have wc reduction. In short, if Th2 decreases we want wc to be reduced, which requires the output controller p to increase; therefore, a reverse action controller is required. 8-12 8.16 Two information is required to determine the action of the controller: i) Is the control valve fail open or fail near ii) Is x1 > x2 or x1 <? x2 If x1 > x2, then the mass balance is: x1w1 x2 w2 · xw · x (w1) x1 x2 x2 x1 o x2 o (x2 o) w1 o x2 w2 · x(w1) x2 w1 o w1 o x2 w2 · x (w1 w2) x2 (w1 o w2) o w1 · x (w1 o w2) x x2 o w1 (w1) w2) Since all variables in the equation are positive, then x > x2 . The only way to reduce x is to increase w2 (but x may never be less than x2). Therefore, w2 should be increased when x increases in order to have x decrease. If the control valve is open (air-close), then the composition controller output signal p should be reduced. A reverse action controller should therefore be selected. Conversely, for an airto-open control valve, a direct action controller should be used. If x1 < x2, then x1w1 x2 w2 · xw · x (w1 w2) x1 o x2 x1 · x2 − (x2 −) w1 x2 w2 · x(w1) w2 w2) x2 w1 − w1 x2 w2 · x(w1 w2) x2 (w1 w2) − w1 · x (w1 o w2) x x2 − w1 (w1) w2) Since all variables are positive , then x < x2 . If x increases, the controller should reduce it to restore it to the specified point. The only way to reduce x is to reduce w2 (although x may never be less than x1). If the control valve is damaged open (air-close), then the output signal of the synthesis controller p should be increased in order to reduce w2. Therefore, the synthesis controller should act directly. Instead, for a fail close control valve, a reverse action controller should be used. 8-13 Chapter 9 © 9.1 (a) Flow speed transmitter: 15 psig - 3 psig qm(psig)= (q gpm - 0 gpm) 3 psig · 400 gpm-0 gpm psig = 0.03 q(gpm) o 3 psig gpm πίεσης: 20 mA - 4 mA Pm(mA)= (p in. Hg 10 μέσα. Hg) 4 mA - 30 in. Hg - 10 μέσα. Hg mA = 0.8 4 mA in. Bg · Flat transmitter: 5 VDC - 1 VDC hm(VDC)= o (h(m) - 0.5 m) o 1 VDC · 10 m - 0.0. 5 m VDC = 0.421 h(m) o 0.789 VDC m · Concentration transmitter: 10 VDC - 1 VDC Cm(VDC)= · (C (g/L)-3 g/L)+1 VDC · 20 g/L - 3 g/L VDC = · 0.529 C (g/L) − 0.59VDC g/L · b) Gains, zeros and intervals are: Concentration of flow pressure level 0.03 psig/gpm 0.8 mA/in. Hg 0.421 VDC/m 0.529 VDC/g/L Zero 0 gpm 10 in. Hg 0,5 m 3 g/L Span 400 gpm 20 in. Hg 9.5 m 17 g/L Gain 9.2 [Type here] 9-1 [Type here] a) Safer conditions are achieved by lower temperatures and pressures in the flash tank. VALVE 1.- Fail near (air-to-open) VALVE 2.- Fail open (air-to-close) VALVE 3.- Fail open (air-to-close) VALVE 4.- Fail open (air-to -close) VALVE 5.- Fail near (air-to-open) Adjustment valve 1 as fail close prevents more heat from switching to flash drum and setting valve 3 as fail open to allow steam chest to drain. Adjusting valve 3 as a failure opening prevents pressure build-up in the container. Valve 4 should failopen to evacuate the system and help keep the pressure low. Valve 5 should be fail-close to prevent any additional pressure buildup. b) Steam flow in downstream equipment can cause a dangerous situation VALVE 1.- Fail near (air-to-open) VALVE 2.- Fail open (air-near) VALVE 3.- Fail near (air-to-open) VAL VIA 24.- Fail open (air-to-close) VALVE 5.- Fail near (air-to-open) Setting valve 1 as fail near (air-to-open) prevents more heat from entering flash drum and minimizes future steam production. Setting valve 2 as fail open (air-to-close) will allow the steam chest to be evacuated, setting valve 3 as fail near (air-to-open) prevents steam from escaping the vessel. Adjusting valve 4 as fail open (air-to-close) allows liquid to leave, preventing the build-up of steam. Adjusting valve 4 as fail near (air-to-open) prevents pressure buildup. c) Liquid flow in downstream equipment can cause a dangerous situation VALVE 1.- Fail near (air-to-open) VALVE 2.- Fail open (air-near) VALVE 3.- Fail open (air-near) VALVE 4.- Fail near (air-to-open) VALVE 5.- Fail near (air-to-open) Adjust valve 1 as fail close to prevent all liquids from evaporating (This will cause the flash drum to overheat). Setting valve 2 as fail open will allow the steam chest to be evacuated. Adjusting valve 3 as fail open prevents the accumulation of pressure in the drum. The valve 4 as closing the failure prevents the liquid from escaping. Valve adjustment 5 as closing failure prevents wet accumulation in drum 9-2 9.3 Note: This exercise is better understood after the material in Ch. 11 has been taken into account. a) Changing the extent of the temperature transmitter will change its steady state gain, according to eq. 9-1. Because the performance of the closed loop system depends on the profits of each individual element (see Chapter 11), 11), may be adversely affected. (b) Changing the zero of a transmitted person shall not affect his profit. Thus, this change will not affect the stability of the closed loop. c) Changing the control valve trim will change the (local) steady state gain of the control valve, dq/dp. Because the performance of the closed loop system depends on the profits of each individual element (see Chapter 11), the stability of the closed loop could be adversely affected (d) For this process, the change in the feed flow rate could affect both the steady state gain and its dynamic characteristics (e.g. time constant and time delay). Because the performance of the closed loop system depends on the profits of each individual element (see Chapter 11), closed loop stability could be adversely affected. 9.4 Starting from Eq. 9-7: q Cv · Pv Nf (l) gs (1) The drop in pressure on the valve is: Pv · P − Ps where Ps · Kq 2 (2) (3) Resolve for K by connecting the nominal values of q and Ps . First, convert the nominal value of q to m3/h units to match the metric unit version of N (parameter N = 0.0865 m3/h(Kpa)1/2 when q has m3/h units and pressure has KPa units). 9-3 qd · 0.6 m3 / min · 36 m3 / h Psd · 200 kPa Psd 200 kPa · 2 · 0.1 54 kPa/(m3 /h) 2 2 qd 36 (m3 /h)2 Now replace (3) in (2) to receive an expression for Pv in terms of q. K · Pv · P − Kq 2 (4) Substitute (4) in (1) to get: q Cv · (5) P − Kq 2 Nf (l) gs The problem determines that qd should be 2/3 of qmax (where qmax is the flow rate through the valve when the valve is fully open). 2 qd · qmax 3 3 3 qmax · qd · 36 m3 /h 2 2 3 qmax · 54m /h Now find the CV that will give qmax = 54 m3/h. Substitute q = qmax and f (l)=1 (valve fully open) in (5). qmax Cv P Kqmax 2 N gs Now that all of the variables on the right hand side of the equation are known, plug in to solve for Cv. kPa m3 P 450 kPa, K 0.154 3 , N 0.0865 , (m / h) 2 h(kPa)1/2 g s 1.2, qmax 54 m3 /h 54 Cv 3 0.0865 m h(kPa)1/2 m3 h 450kPa 0.154 m3 h m3 m3 1/2 0.0865 0.88(kPa) 0.076 h(kPa)1/2 h Cv 710.5 54 m3 h kPa 54 2 (m 3 /h) 2 3 2 (m /h) 1.2 54 9-4 9.5 Let Pv/Ps = 0.33 at the nominal q 320 gpm Ps = Pb+ Po = 40 + 1.953 10-4 q2 Pv= P - Ps = (1 –2.44 10-6 q2)PDE – (40 + 1.953 10-4 q2) (1 - 2.44 10 -6 320 2)PDE - (40 + 1.953 10 -4 320 2) · 0.33 (40 + 1.953 · 10 -4 · 320 2) PDE = 106.4 psi Let qd= q · 320 gpm For the rated Cv, the valve is completely open at 110% qd i.e., at 352 gpm or the ceiling of 350 gpm p Cv · q · v · gs − 1 2 (1 − 2.44 · 10−6 · 35 0 2)106.4 − (40 o 1.953 · 10--4 · 3502) Cv · 350 · 0.9 − 1 2 Subsequently, using Eq. 9-27, q 66.4 − 10 - 4 q 2 In 0.9 101.6 - 1 I I −1/2 The diagram of the valve characteristic is shown in Figure S9.5. From the plot of the valve featuring the graded CV 101.6, it is obvious that the characteristic is quite linear in the operating range of 250 · q · 350. The cost of pumping could be further reduced by reducing PDE to a price that will make Pv/Ps = 0.25 to q · 320 gpm. Then PDE = 100 and for qd = 320 gpm, the nominal Cv = 133.5. However, as the plot shows, the valve featuring this design is only slightly more nonlinean in the operating area. Therefore, the selected valve factor is Cv = 133.5. 9-5 400 350 300 250 q (gpm) 200 150 ------- Cv = 101.6 100 - - - Cv = 133.5 5 0 0 0.1 0.2 0.3 0.4 0.5 0.6 0.7 0.8 0.9 1 l (valve lift) Figure S9.5. Control valve characteristics. 9.6 (a) (b) There are three control valves. The choice of air-to-near versus air-to-open is based on safety reasons: i. Steam control valve: Air-open to prevent overheating of the evaporator. ii. Level control valve (adjusting the flow rate of liquid B): From air to aperture to prevent exposure of steam coils to the vapour space, which could cause the coils to burn out. iii. Pressure control valve (adjusting the flow rate of solvent D): From air to closure to avoid over-pressure of the evaporator. For the three controllers: i. Concentration controller: As the concentration of the xB product increases, we want the steam pressure, Ps to increase. Since the steam valve is air-to-air, this means that the controller output signal to the control valve (via I/P) should be increased. Therefore, the controller should act immediately. 9-6 p. Level controller: As the liquid h level increases, we want the B product flow rate to increase. Since the control valve is air-to-air, this means that the controller output signal to the control valve (via I/P) should be increased. Therefore, the controller should act immediately. iii. Pressure controller: As the pressure P increases, we want to increase the flow rate of solvent D. Since the control valve is air-to-close, this means that the controller output signal to the control valve (via I/P) should be reduced. Thus, the controller should act in reverse action. 9.7 Since the dynamic behaviour of the system will be described using deviation variables, the dynamic attribute can be analysed taking into account that the input conditions (not related to x) can be considered constant and therefore the deviations are zero. The original form is the linear homogeneous ODEON: M d 2x dx o R o K x · 0 2 gc dt dt Taking the transformation Laplace gives, X (s) · s 2 o Rs o K · 0 · gc M 2 R X (s) · s o s o 1 · 0 K · Kgc Calculate and g comparing this equation with the standard format of the second order model in (5-39) (bearing in mind that gc = 32,174 lbm ft/(lbf s2)). M 0,00965s Kg c · 2 · R 2 M R · Kg c K gc · 155.3 KM The characteristics of the valve are extremely more than multiplied and can be accurately accessed with a first-class model obtained by neglecting the term d2x/dt2. 9-7 9.8 Configuration I: This series configuration will not be very effective because a large flow rate must pass through a small control valve. Thus, the pressure drop will be very large and the flow control will be ineffective. Configuration II: This parallel configuration will be effective because the large control valve can be adjusted to provide the nominal flow rate, while the small control valve can be used to adjust the flow rate. If the small valve reaches its maximum or minimum value, the large valve can be adjusted slightly so that the small valve is about half open, allowing it to adjust the flow again. 9.9 First note the time-domain step response for a 10°C step change. Then solve the equation to find out when y(t) is equal to 5 (since the variables are in deviation variables, this represents when TM will reach 30 °C). ym (t) · KM (1 − e − t /) where M · 10o C, K · 1, and o 10s ym (t) · 10(1 − e − t /10) 5 · 10(1 − e − ta /10) ta · 6.93s Therefore, the alarm will sound 6.93 seconds after 1:10 pm. 9.10 0.1 psig · 0.5% of full-scale accuracy of 20 psig is unknown, since the true pressure in the tank is unknown resolution 0.1 psig = · 0.5% accuracy 20 psig full scale = repeatability = ±0.1 psig =±0.5% of the full scale 20 psig 9-8 9.11 Suppose the sensor/transmitter gain is module (i.e. there is no constant state measurement error). Then tm(s) 1 · T (s) (s 1) (0.1s 1) where T is the temperature measured and Tm is the measured value. To change the temperature of the ramp: 0.3 T (t) = 0.3t (°C/s) , T (s) = 2 s 1 0.3 Tm (s) · 2 (s o 1)(0.1s 1) s Tm o o 2 (s o 1)(0.1s 1) s Tm <1> <5> (t) · 0.00333e−10t 0.333e−-t μ 0.3t − 0.33 The maximum error occurs as t- : Maximum error = |0.3t − (0.3t − 0.33)| = 0.33 °C If the shortest time constant is neglected, the time domain response is slightly different for small t-values, although the maximum error (t-) does not change. T °C 6 5 4 3 2 1 0 0 5 10 15 Time 20 Figure S9.11. Response for process temperature sensor/transmitter. Orange solid line is T'(t), and purple dotted line is T'm(t). 9-9 Chapter 10 © 10.1 Assumptions: 1. Uncompressed flow. 2. The chlorine concentration shall not affect the density of the air sample. 3. T and P are about stable. The detection time, td, depends on the delay of the transfer time, the value and response time of the analyser, tr = 10 s: td = , – tr (1) delay) can be calculated as the ratio of the volume of tubing V divided by the volumetric flow rate of chlorine q: i · V q q (2) where q = 10 cm3/s and, p Di 2 L V, 4, where Di diameter is: (3) Di = 6.35 mm – 2(0.762 mm) = 4.83 mm = 4.83 x 10-3 m Substitute Di and L = 60 m in (3): V = 1.10 x 10-3 m3 Substitute Di into (2): V 1.10 x10 3 m3 100cm · · · 110 s q · 10 cm3 /s · 1 m 3 i · The substitute in (1): td = - + tr = 110 + 10 = 120 s = 2 minutes carbon monoxide (CO) is one of the most widely displayed toxic gases, especially limited for confined spaces. High concentrations of carbon monoxide can korean a person's blood in a few minutes and quickly lead to respiratory problems or [Press here] 10-1 [Press here] even death. Therefore, long detection time would not be acceptable if the hazardous gas is CO. 10.2 (a) Start with a mass balance in the tank. Then solve the equation to find out how long it takes for the height to drop from 1 m to 0.25 m. dV (t) · −C h(t) dt Adh(t) · −C h(t) dt dh C ¦ − h0.5 dt A C h −0.5 dh − dt A 0.25 h tf −0.5 dh · − 1 0 C dt A 2(0.25[m] − 1[m]) − − −1m0.5 · C (t f − 0) A C tf A A 1[m0.5] C o (0 .5) 2 [m 2] tf · [m0.5] 2.5 0.065[m/min] t f · 12.1[min] tf the alarm will sound at 5:12:06 am (b) To find out how much liquid has leaked from the tank , calculate the volume difference between the start level and the alarm level. 2 1 V · Vh · 1m − Vh · 0.25 m · · o [m] · 1[m] − 0.25[m] · · 0.59m3 · 2 0.59m3 of liquid has leaked when the alarm sounds. 10.3 10-2 Key safety concerns include: 1. Early detection of leaks in the environment 2. Over-investment of the flash drum 3. Maintain enough fluid level so that the pump is not a cavity. 4. Avoid driving liquid to the gas. These concerns can be addressed by the following means. 1. Leak detection: sensors for hazardous gases shall be located near the flash drum. 2. Over compression: Use a high pressure switch (PSH) to turn off the feed when a high pressure occurs. 3. Liquid stock: Use a low-level switch (LSL) to shut down the pump in case of low level. 4. Liquid stimulation: Use a high level alarm to turn off food if the liquid level becomes too high. This SIS system appears in the Schema. S10.3 with conventional control loops for pressure and liquid level. Steam PSH PT LSH S PC P Feed LT LC LSL S Liquid Figure S10.3: SIS System 10.4 10-3 The proposed alarm/SIS system is shown in Figure S10.4: S PSH C O L U M N Figure S10.4: Proposed alarm system/SIS The tubular valve is normally closed. But if the pressure in the column exceeds a specified limit, the High Pressure Switch (PSH) activates an alarm (PAH) and causes the valve to fully open, so the pressure in the tank. 10.5 Set k as the number of sensors that work correctly. We want to calculate the probability that k · 2 , P (k · 2) . Because k = 2 and and = 3 are mutually exclusive events (see Appendix F), P(k · 2) · P(k · 2) o P(k · 3) These probabilities can be calculated from the dionymal distribution 1 3 1 P(k) · o 0.05 (0.95)9 2 · 0.135 · 2 3 0 P(k) · · · 0.05 · (0.95)3 · 0.857 · 3 10-4 (1) n if the notation, It refers the number of combinations of n objects taken r · 3 at a time, when the order of r objects is not significant. So · · · 3 and · 2 3 · · 1. From EU 1, · 3 P (k · 2) · 0.135 · 0.857 · 0.992 1 See any typical probability or statistics book, e.g., Montgomery D.C. and G.C. Runger, Applied Statistics and Engineering Probability, 6h edition, Wiley, New York, 2013. 10.6 Solenoid switch: S = 0.01 Level switch: LS = 0.45 Level alarm: A = 0.3 Notation: PS = the probability that the solenoid switch fails PLS = the probability that the level switch fails PA = the probability that the level alarm fails PI = the probability that the interlock system (solenoid & level switch fails) We wish to determine, P = the probability that both safety systems fail (i.e., the original system and the additional level alarm) Because the interlock and level alarm systems are independent, it follows that (cf. Appendix F): P = PI PA (1) From the failure rates, the following table can be constructed, in analogy with Example 10.4: Component R P=1–R Solenoid: Level switch: 0.45 0.638 0.362 Level alarm 0.3 0.741 0.259 10-5 Assume that the switch and solenoid are independent. Στη συνέχεια, PI = PS + PSW - PSW PI = 0,01 + 0,362 – (0,01)(0,362) PI = 0,368 Υποκατάστατο σε (1): P = PI PA = (0,368)(0,259) = 0,095 Μέσος χρόνος μεταξύ των αποτυχιών, MTBF: Από (10-6) έως (10-8): R = 1 – P = 1 – 0,095 = 0,905 = - ln (0.0095 = 905) = 0,0998 MTBF 1 10,0 έτη 10,7 Let P2 = η πιθανότητα να μην λειτουργεί σωστά ούτε το ροόμετρο D/P. Στη συνέχεια, το P2 και η σχετική αξιοπιστία, R2, μπορούν να υπολογιστούν ως (βλ. προσάρτημα ΣΤ): P2 = (0,82)2 = 0,672 R2 =1 - P2 = 1 - 0,672 = 0,33 Για τον υπολογισμό της συνολικής αξιοπιστίας του συστήματος, αντικαταστήστε το R2 = 0,33 για την τιμή αξιοπιστίας για ένα μόνο μετρητή ροής D/P, 0,18, στον υπολογισμό R του παραδείγματος 10.4: 5 R - Ri R (0,33)(0,95)(0,61)(0,55)(0,64) i 1 R 0,067 Έτσι , the addition of the second D/P flowmeter has increased the overall reliability of the system from 0.037 (for example 10.4) to 0.067. 10-6 10.8 Let P3 = the probability that none of the 3 D/P flowmeters are working properly. Then P3 and the relative reliability, R3, can be calculated as (see Appendix F): P3 = (0.82)3 = 0.551 R3 =1 – P3 = 1 – 0.551 = 0.449 To calculate the overall reliability of the system, replace R3 = 0.449 for the reliability value for two D/P flows (R2=0.33) in the calculation of R from the calculation of R by the of R from the calculation of R from the calculation of R from the calculation of R from the calculation of R from the calculation of R from the calculation from the calculation of R from the calculation of R from the calculation from the calculation of R from the calculation of R from the calculation from the calculation from R from the calculation of R from the calculation from the calculation from R from the calculation from calculation R2=0.33) to the calculation of R from the calculation of R from the calculation of R from the calculation from the calculation from the calculation of R the calculation from the calculation of R from the calculation from the calculation from the calculation from R from the calculation of R from the calculation from the calculation of R from the 10.7: 5 R - Ri (0.449)(0.95)(0.61)(0.55)(0.64) i 1 R 0.092 Έτσι, η προσθήκη του τρίτου ροόμετρου D/P αύξησε τη συνολική αξιοπιστία του συστήματος από 0,067 (για άσκηση 10,7) σε 0,092. 10.9 Ας υποθέσουμε ότι ο διακόπτης και το σωληνοειδές είναι ανεξάρτητα. Από τα δεδομένα ποσοστού αποτυχίας, μπορεί να κατασκευαστεί ο ακόλουθος πίνακας, σε αναλογία με το παράδειγμα 10.4: διακόπτης πίεσης στοιχείων R P=1–R 0,34 0,712 0,288 Solenoid διακόπτης/βαλβίδα: Ας υποθέσουμε ότι ο διακόπτης και το σωληνοειδές είναι ανεξάρτητα. Στη συνέχεια, η συνολική αξιοπιστία του συστήματος κλειδώματος είναι, R = (0,712)(0,657) = 0,468 = - ln (0,468) = 0,760 MTBF 1 10-7 1,32 έτη Κεφάλαιο 11 11,1 D1 Gd1 D2 Gd2 X 2 Ysp Km ~ Ysp +- E Gc P U Gv B Gp X3 + + X1 + + Gm 11,2 1 Gc (s) K c 1 I s Η λειτουργία μεταφοράς κλειστού βρόχου για αλλαγές σημείου ρύθμισης δίνεται από τον Eq. 11-36 11 1 1 , με Kc αντικαθίσταται από K c 1 I s 1 1 K c K IP K v K p K m 1 I s (s 1) Η (εc) H sp (εc) 1 1 1 K c K IP K v K p K m 1 I s (s 1) όπου Kp = R = 1,0 min/ ft2 , και = RA = 3,0 λεπτά. Note also that $\tau I = \tau = 3.0$ min. Solution Manual for Process Dynamics and Control, 4th edition Copyright ©

2016 by Dale E. Seborg, Thomas F. Edgar, Duncan A. Mellichamp, and Francis J. Doyle III 11-1 Y psi ft 3 / min min mA KOL K c K IP K v K p K m (5.33) 0.75 0.2 1.0 2 4 3.2 mA psi ft ft 3s + 1 1 3.2 (3s) (3s + 1) H ' (s) 3.2 1.0 = = = ' Hsp (s) 1 + 3.2 (3s + 1) (1) 3s + 3.2 0.94s + 1 3s 3s + 1 (s) For H sp (3 2) 1 s s h(t) 1 e 1.07 t 0.94ln 1 h(t) h(t) 2.5 ft h(t) 0.5 ft t 0.65 min h(t) 3.0 ft h(t) 1,0 ft t, h(t 0,65 min) 2,5 ft h(t) 3,0 ft 11,3 Gc (s) K c 5 ma/ma Ας υποθέσουμε m = 0, v = 0 και K1 = 1, στο σχήμα 11.7. a) Offset = Tsp ( ) T ( ) 5 F 4.14 F 0.86 F b) K K m K c K IP K v 2 T (s) s 1 Tsp (s) K 1 K m K c K IP K v 2 s 1 Using the standard current range of 4-20 ma, 11-2 Km 20 ma 4 ma 0.32 ma/ F 50 F K v 1.2 , K IP 0.75 psi/ma , =5 min , Tsp (s) T (s) 7.20K 2 s(5s 1 1.440K 2) T ( ) lim sT (s) s 0 T ( ) 4.14 F c) 5 s 7.20K 2 (1 1.440K 2) K 2 3.34 F / psi From Fig. 11.7 , από τι 0 Pt( ) K v K 2 T ( ) και Pt( ) 1,03 psi Pt K v K 2 Ti K1 T , Pt 3,74 psi Pt ( ) Pt Pt( ) psi 11.4 (a) Controlled variable: c3 Manipulated variable: q2 Disturbance variable: c2 (note: q1 and c1 kept constant.) If c2 changes, then q2 must be adjusted to keep c3 at the specified point. 11-3 C'2(s) lb sol ft3 ^ C'3sp(s) C'3sp(s) lb sol ft3 Km ma E(s) +- Gc ma psi C'3m(s) ma Q'2(s) Gv GIP ma X1 P'T(s)) P's Gd X2 Gp USGPM C'3(-s) Gm lb sol ft3 (b) Gm (s) • K m e – - s assuming • m = 0 m Gm(s) • (20 – 4)ma –2 s ma – 2 s e • • • 2.67 lb sol lb sol/ft 3 • (9 – 3) 3 ft 1 Gc (s) • K c • • 1 o s I • G IP(s) • K IP • 0.3 psi/ma Gv(s) • K v • (10 – 20) USGPM USGPM – • 1.67 (12 – 6) psi For the process and function of disturbance transfer : Total material balance for the tank, USgallons dh • q1 q 2 – q3 • 7,481 A ft 3 • dt As h is kept constant at 4 ft from the overflow pipe: 0 • 10 • 15 – q3 (1) So q3 • 25 Component balance for soluble, 11-4 + + C'3(-a) 7.481 Ah d (c3) • q1c1 q2 c2 – q3 c3 dt (2) Line each term on the right side of Eq. 2 as described in section 4.3: q1c1 q1 c1 q1c'1 q'1 c1 q 2 c 2 • q 2 c 2 q 2 c' 2 q' 2 c 2 (3) q 3 c3 • q 3 c3 q 3 c 3 q q ' 3 c3 In a stable state : 0 • q1c1 o q 2 c 2 – q3 c3 (4) Put (2) in deviation variable taking into account (3) and (4): 7.481 Ah dc'3 • q 2 c ' 2 o q ' 2 c 2 – q 3 c ' 3 – q ' 3 c3 dt As q'3 = q'2: 7,481 Ah dc'3 • q 2 c ' 2 q ' 2 c 2 – q 3 c ' 3 – q ' 2 c3 dt 7.481 Ah dc'3 • q 2 c ' 2 ' q ' 2 (c 2 – c 3) 3) – 3 c ' 3 dt (5) Taking Laplace convert and rearrange gives K 1 K2 Q2 (s) C 2 (s) (s) (6) os os o 1 os 1 c – c3 q lb sol/ft 3 7.481Ah where K1 • 2 , K 2 • 2 • 0.6 and • • 0.08 • 15 minutes q3 q3 q3 q3 USGPM C3 (s) from a • • D 2 / 4 • 12.6 ft 2 , and h • 4 ft . Therefore, G p(s) • 0.08 0.6 and Gd(s) • 15s o 1 15s 1 (c) Closed loop responses for disturbance changes and adjustment point changes can be obtained using algebra block diagram for the block diagram in part (a). 11-5 Therefore, these responses will only change if any of the transfer functions in the blocks of the diagram change. i. c 2 changes. Then block transfer mode G p (s) changes due to K1. Therefore Gc(s) need to be changed, and retuning is required. ii. Changes in km. Narrow loop transfer functions change, therefore Gc(s) must be adjusted to compensate for changes in Gm and Km. Pi controller needs to be reset. iii. Km remains unchanged when zero is set. The controller does not need to be reset. To verify the results of linearity, the nonlinear model is used: d (c 3) 7.481 Ah • q1c1 q2 q2 – q3 c3 dt q1 o q 2 • q3 Response step c3 to q2: (Profit 0.077 compared to linear gain (Kp) 0.08 in 6 Eq. response c3 to c2: (Profit 0.6 compared to linear gain (Kd) 0.6 in Eq. 6) 11-6 Results are consistent with linearity. 11.5 (a) From Eq. 11-26 we get the closed loop transfer function for changes that point Km Gc Gv Gp Y = Ysp 1 + Gc Gv Gp Gm Substitution of information from the problem gives 4 Y 4 4 s(s + 4) = = 2 4 Ysp 1 + s (s+ 4) + 4 s + 4 s(s + 4) 1 Or in standard format (Eq. 5-40), with t = 2 and g= 1 Y 1 = Ysp 1 s 2 + s + 1 4 (b) Given a unit step change at the specified point we have 11-7 Y(s) = s (s 2 4 + 4s + y 4) Using the Final Value Theorem we get 4 4 lim sY(s) = 2 = =1 s + 4s + 4 4 s⟶0 Therefore y(∞) = 1 (c) As the step change is a unit step change , and we have shown that y(∞) = 1, we can say that offset = 0. This is consistent with the fact that the profit of the total transport operation is 1, so there will be no compensation. Normally analog control does not eliminate compensation, but does for this integration process. d) Using Eq. 5-50 or taking the inverse Laplace Transformation of the answer given above we get y • t • 1 – • 1 o 2t o e--2t The replacement of the value of 0.5 for t gives y • t • 0.264 (e) We can say from the answer above that the answer above that the answer will not be oscillation, the answer will not be oscillation, • 1 . 11.6 For the analog controller, gc(s) • K c Assume that the level transmitter and control valve have negligible dynamics. Then Gm(s) • K m Gv (s) • K v 11-8 The block diagram for this control system is the same as in Fig.11.8. Hence eqs. 11-26 and 11-29 can be used for closed loop responses to adjustment and load point changes, respectively. Transport modes G p (s) and G d (s) are as given in Eqs. 11-66 and 11-67, respectively. a) Substituting for Gc, Gm, Gv, and Gp into Eq. 11-26 gives 1 Km Kc Kv Y 1 As Ysp s 1 1 1 Kc Kv K m As A where Kc Kv Km (1) For a step change in the setpoint, Ysp (s) M / s M /s Y (t) lim sY (s) lim s M s 0 s 0 s 1 Offset = Ysp (t) Y (t) M M 0 b) Substituting for Gc, Gm, Gv, Gp , and Gd into (11-29) gives 1 1 K c K v K m Y (s) As D(s) s 1 1 1 Kc Kv K m As where It is given by Eq. 1. For a step change in the disorder, D(s) • M / s – M /(K c K v K m) –M Y (t o) • lim sY (s) • lim s • • s o0 s 1) Kc Km km –M Offset = Ysp (t o) – Y (t o) • 0 – • • Kc Km μ 0 Therefore, is not removed for a step change in the disorder. 11-9 Using diagram block algebra Y • G d D q pU (1) • ~ U • Gc Ysp – Y – G pU • From (2), • (2) GcYsp – GcY ~ 1– Gc G p Replacement for U in Eq. 1 1 G (G c p • ~ ~ ~ ~ G p) Y • Gd (1 – Gc G p) D G p Gc Ysp Therefore, G p Gc Y • ~ Ysp 1 g gc (G p – G p) and Gd (1 – Gc G p) Y • D 1 g (G p – G p) 11.8 The available information can be translated as follows 1. The outputs of both tanks have a q0 flow rate at all times. 2. Up to (s) • 0 3. Since an energy balance indicates a first-row transfer function between T1 and Q0 , T (t) • 1 – e –t / • 1 T ( ) or 2 • 1 – e – 12 / • 1 , • 1 = 10.9 minutes 3 11-10 Therefore T1 (s) 3 F / (--0 ,7 5 gpm) 4 • • • Q0(s) 10.9s o 1 10.9s o 1 T3 (s) (5 – 3) • F /(--0.75gpm) 2.67 • • Q0 (s) • 2s • 1 o 2s 1 4. for t2(s) = 0 T1(s) (78 – 70) • F /(12 – 10)V 4 • V1(s) 10s 1 10s 1 T3(s) (90 – 85) • F /(12 – 10)V 2.5 • V2 (s) 10s 1 1 10s 1 5. 5 • 2 =50 min or • 2 = 10 minutes Since the input and output flow rates for tank 2 are q0 and tank volumes are equal, T3 (s) q0 / q0 1 • T2 (s) • 2 s o 1 10.0s 1 6. V3(s) • 0.15 T3(s) 30 7. T2 (t) • T1 • t – t1 (t – 0.5) • 60 T2 (s) • (e) 0.5 s T1 (s) (s) Using these transfer functions, the block diagrams are as follows. 11-11 (a) (b) (c) The configuration of the control in Part (a) shall provide the best control. As shown in the above block diagrams, the feedback loop contains, in addition to gc, only a first-class process in part a), but a second series of co-conspirator-time-delay process in part b). Therefore, the controlled variable responds more quickly to changes in the manipulated variable for part (a). 11-12 11.9 The given block diagram is equivalent to Gd Ysp E' +- +~ ~ Y1-Y2 P Gc ~ ~ G *(1-e-□s) For the inner loop, let Gc P • Gc • ~ ~ E 1 gc G * (1 – e – -s) In the outer loop, we have Gd G Y • D 1 g gc g Substitute for G c , Y • D Gd G GC G 1 o ~ ~ * 1 g g (1 – e – – -s) d G 1 g gc g * (1 – e – □s) Y • ~ ~* D 1 g g g (1 – e – e) gc g 1 1.10 a) Draw CLT F: Y Y3 Y2 G3 Z G2 P 11-13 + + G D Y Y • G3 (D o Y1) G2 K c E Y G3 D G3 G1 K C E G2 K E Y G3 D (G3G1 K c G2 K c) E E • – K mY Y • G3 D – K c (G3G1 g2) K mY G3 Y • D 1 k c (G3G1 g2) K m b) Characteristic equation: 1 k c (g3g1 g2) k m 0 4 5 1 kc o o 0 s – 1 2s o 1 5(2 s 1) o 4(s – 1) 1) Kc (s – 1) (s – 1) (s – 1)(2s 1) k c 5(2s) 1) 4(s – 1) 2s 2s 2s – s – 1 k c (10s 5 4s – 4) • 0 2s 2s 2s 2 (14K c – 1) s (K c – 1) • 0 Necessary K c 1/14 and K c 1 For a 2nd series characteristic equation, these conditions are also sufficient. Therefore, K c is 1 for closed loop stability. 11.11 11-14 (a) CF's) Kg/m3 ~ C'sp(s) Km Kg/m3 ma +- GL X1 E ma P' GPI + + Pv' KIP ma Q'A Gv m3/min psig X2 Gp + + GD C'm(s) C'TL Gm ma Kg/m3 C' GTL Kg/m3 b) Transmission line: Transmission line volume = / 4 (0.5 m)2(20m)= 3.9 3 m3 Nominal flow rate on line = q A o q F • 7.5 m 3 / min Time delay on line = 3.93 m 3 • 0.52 minutes 7.5 m 3 /min GTL (s) • e –0 ,52 s Composition transmitter: Gm(s) • K m • (20 – 4) ma ma ma • 0.08 3 (200 – 0) kg/m kg/m 3 Controller from the ideal controller in Eq. 8-14 • 1 ~ E (s) o K c o D s C sp (s) – C m (s) P (s) • K c o1 • • I s ~ In the above equation , set C sp (s) 0 to receive only the derivative at the process output. Then, 1 G PI (s) • K c • • 1 o I s 11-15 C's) GD(s) • K c • D s with Kc >0, as the controller should be reverse action, since P(t) should be increased when Cm(t) decreases. I/P converter K IP • (15 – 3) psig psig • 0.75 (20 – 4) ma ma Control valve Gv (s) • Vv • v s o 1 5 v • 1 , Rv • dq A dpv • v • 0.2 minutes • 0.03(1 / 12) (ln 20)(20) pv • pv q A • 0.5 • 0.0.0.0. 17 0.03(20) 0.03(20) pv 3 12 pv 3 12 0.5 0.17 0.33 K v (1/12)(ln 20)(0,33) 0,082 Gv (εc) pv 3 12 m 3 /min psig 0,082 0.2s 1 Διαδικασία Ας Ας cA είναι σταθερή για καθαρό A. Ισοζύγιο υλικών για το A: V dc q A γ A q F c F (q A q F) c dt Γραμμοποίηση και γραφή σε μεταβλητή απόκλισης 11-16 (1) V dc c A q A q F c F q A q F)c c q A dt Λήψη Vs (q A q F) C (-α) (γ A γ)Q A (εc) q F C F (εc) (2) Από Eq. 1 σε σταθερή κατάσταση , dc / dt • 0 , c • (q A c A) q F c F) /(q A o q F) • 100 kg/m 3 Substitution of numeric values in Eq. 2, 5s o 7.5-C (s) • 700 Q A(s) o7 C F (s) 0.67 s o 1-C (s) • 93.3 Q (s) (s) 0.93 C F (s) 93.3 0.67s (1 0.93 Gd (s) • 0.67s o 1 G p(s) • 11.12 Stability limits are derived from the Eq attribute. 11-83. Therefore, if a change of organs affects this equation, then the stability limits will change and vice versa. a) The transmitter gain, Km, changes as the extent changes. Thus Gm(s) changes and the characteristic equation is affected. Stability limits are expected to change. b) Zero in the transmitter does not affect km gain. Therefore gm(s) remains unchanged and stability limits do not change. c) Changing the control valve trim changes the G. This affects the characteristic equation and the stability limits are expected to change as a result. 11-17 11.13 (a) 1 Kc • 0 ' 5s 2 o 6 s o 1 o K c • 0 o 5s o 1 o o 1 o Application the square type produces the roots: s – 6 36 – 20 • 1 o K c • 10 To have a stable system, both roots of the characteristic equation must have negative real parts. As a result, 20 o1 o Kc o 0 Kc −1 1 K c • 1 o Is) 1 o 0 • A o 5 3 o 5s 3 o 3 o 1 o 5s 3 6 s 2 o s o K c s o K c • 0 o 5s o 1 o 1 o 1 • 3 2 When • 0.1 , 0.5s o 0.6s o 0.1 o1 o 1 o Kc o Kc • 0 Using direct substitution, and set • j • : • 0.5 • Re: Im: 3 o 0.1 o1 o K c • • j – 0.6 • 2 o K c 0 –0.6o2 + Kc = 0 –0.5o3 + 0.1(1 + Kc)h = 0(2) • o 0 : K cm • 0.136 (1) To have a stable system , we have: 0 o K c o 0.136 3 2 When • • 1, 5s o 6s o1 o 1 o Kc • Kc • 0 Set s • j • • • • • 1 o Kc os o Kc • 0 Set s • • 1 •

the roots of the polynomial attribute to become positive. However, for first or second order processes, increasing Kc will not cause instability. d) Always true. The increase in the auditor's profit will reduce the shift. However, if profit increases too much, oscillations may occur. Even with oscillations the displacement will continue to decrease until the system becomes unstable. 14-10 14.15 (a) Figure S14.15a Bode plot of GOL. (Kc = 10) 1 Kc (4 s 1)(2 s 1) Cannot become unstable – maximum phase of 2nd series superagreed process (GOL) is -180 degrees. GOL · GGc · b) Figure S14.15b Provision plot of GOL. (Kc = 10) 14-11 GOL · GGc · (5s 1) K c 1 o 1 / 5s Kc o (4s 1)(1)(1 2s 1) 5s(4s 1)(2s 1) Cannot become unstable - maximum phase angle of (GOL) is -180 degrees, while at low frequency the integrator has -90 degree phase angle. (c) Figure S14.15c A plot of GOL is proposed. (Kc = 10) GOL · GGc · s o1 2s o 1 s(4s 1) GOL · GGc · K c K c · (4s o 1)(2s o 1) s(4s o 1) 2s o 1 Cannot become unstable - maximum phase angle is greater than -180. 14-12 (d) Figure S14.15d Provision plot of GOL. (Kc = 10) GOL · GGc (1 - s)Kc 1 · Kc (4s 1)(2s o 1) (4s o 1)(2s 1) It can become unstable - maximum phase of 270 degrees. (e) Figure S14.15e A plot of GOL is proposed. (Kc = 10) 14-13 GOL · GGc · e -s Kc (4s 1) It can become unstable due to time delay at high frequency. 14.16 Using standard plot, bode diagram 2 10 size (abs) 1 10 0 10 -1 10 -2 10 0 10 1 10 2 parallel row with filter phase (deg) -45 -90 -13 15 -180 -225 -270 -2 10 -1 10 0 10 1 10 2 Frequency (rad/sec) Figure S14.16 Predict plot for exercise 13.8 Transfer mode multiplied by PID Controller Transfer Mode. Two cases: a)Parallel to Series with Derivative Filter (f=0.2). Range ratios: Ideal PID controller: AR = 0.246 = 0.367 In series PID controller: AR=0.294 in · 0.74 There is a 19.5% difference in AR between the two controllers. 14-14 14.17 (a) Method examined in section 6.3 (1) · 12e -0.3s (8s 1) (2.2s) Method examined in section 7.2.1: Response step of Z(s) 12 10 X: 10.10.1. 09 Y: 7.2 Value 8 6 4 X: 4.034 Y: 2.4 Turning point 0 0 10 20 30 40 50 Time 60 70 80 90 100 Figure S14.17a (from Fig. 14.17a) 0 response step(s) Based on Figure S14.17, we can obtain time stamps in 20% and 60% answer: t20 = 4.034; t60 = 10.09- 20/t = 0.4. Based on 60 t in figure 7.7, we have 60 = 2.0; g = 1.15, so we have t = 5.045. Using t the slope of the inflection point we can estimate the time delay to be 0.8. So we have: 12e -0.8 s G2 (s) · 25.45s 2 o 11.60 s 1 Bode plot in Figure S14.17a, we can receive i = 0.8; t = 15 – 0.8 = 14.2 12e –0.8 s G3 (s) · 14.2 s o 1 Comparison of three estimated models and the exact model in the frequency sector using bode plots: 14-15 Bode Diagram 50 Size (dB) 0 -50 G (s) -100 G1 (s) G2 (s) -150 G 43 (s) -200 Phase (deg) -1.152 -2.304 -3.456 -4.608 -10 10 -2 10 -1 0 10 0 10 1 10 2 Frequency (rad/s) Figure S14.17b Comparison of estimated models. 14.18 The initial transfer model is 10(2s) -2 s G(s) · (20s 1)(4s 1)(s 1) (a) the transfer function obtained with the use of section 6.3 is: G s 10 e -2s (20 s 1 o e -5 s o 20 s 1 o 14-16 3 10 Chart Bode 1 Size (abs) 10 0 10 -1 10 -2 G(s) 10 -3 G1 (s) -360 Phase (deg) -720 -1080 -1440 -1800 -2160 -2520 -2880 -3 -2 -1 10 10 10 10 0 10 1 Frequency (rad/sec) Figure S14.18 Bode plots for accurate and approximate models. As shown in Fig. S14.18, the approach is good at low frequencies, but not as good at higher frequencies. 14.19 a) G · G p G m 2e –1.5 s 0.5e –0.3s 3e –0.2 s 3e –2 s (s o 1) (5s 1) 3s 1 2s 1 (60s) 1(60s 5s o 1) (2s o 1) o) oh case where f = –180° : · c. 0.152 AR () · 0.227 · K cu · · 4.41 AR () 14-17 Figure S14.19a Bode plot to find the plot. Simulation results with a different Kc are displayed in the Figure S14.19b. Kc gel: You see that the system becomes unstable as predicted. Kc · 60 o 0.5 · 5 · 62.5, · 2.5 o 2 o 2 · 9.5 The approximate model FOPTD: 14-18 G · 3e –9.5 s 62.5 s 1 Kc = 0.586(0.5 / 62.5) 62.5 · Kc = 0.586(0.5 / 62.5) 62.5 · 1.10 1 o 1 62.5 · 1.10 1 o 9.5 s · 62.19s, GOL · GGc you receive from f = –180: · c · 0.153 AR ( ) · 0.22 ... K cu · · 4.02 AR (c) Figure S14.19c Pronominer of the FOPTD model. 14.20 Using the Bode plot, at a phase angle of -180°, we demand that K c K V K p K m o 1 G p(s) · 0.5 GM · 0.5 Gm · 1.0 14-19 Gp gain = 1.0 for all · s · (a) 2 · rad · 0.628 10 minutes The critical gain is easily found by K c K V K p K m K c · K cu (1.5)(1)(1)point 1 · 1, or K cu · 2.0 b) The phase angle of G c G · Gcu = phase angle of e –!!! , or F = –p (rad) (Eq. 14-33) when F = -180° = –p = p Επειδή, ψ = rad 10 τότε: λεπτό. 10 λεπτά 2 14.21 a) Χρήση Eqs. 14-56 και 14-57 2 2 5 1 1 ... 2 2 25 ... 14-20 Διάγραμμα bode 2 10 1 10 0 10 -1 10 -2 10 -3 10 -4 Amplitude ratio 0 Phase angle (deg) -150 -200 -250 -300 -350 -2 10 -1 10 0 10 1 10 Συχνότητα (rad/sec) (a) Σχήμα S14.21a Οικόπεδο Bode που ... 180° και πάνω, or AR = 0.4695 στη συνέχεια AR OL = 1 = Kcu(1.025) ... Kc του 10, Kcu = 1/1.025 = 0.976 Το σύστημα είναι σταθερό για Kc = 0.976 Για Kc = 0.2, set AROL = 1 and solve for ω to obtain ω = 0.1404 Then ... = –133.67 From Eq. 14-61, PM = 180 + –46.47 d) From Eq. 14-60 GM = 1/ 1 = ... Therefore AR OL = 1.025 Kc 1.025 Kc = 1.025 Kc = 11.7 or Kc = 0.574 Bode Diagram 2 10 1 AR/Kc 10 0 10 -1 10 -2 10 -150 Phase (deg) -180 -200 -250 -300 -350 -2 10 -1 10 0 10 1 10 2 Frequency (rad/sec) Figure S14.21b Solution for part b) using Bode plot Bode Diagram 2 10 1 AR/Kc 10 0 10 -1 10 -2 10 -150 Phase (deg) -180 -200 -250 -300 -350 -2 10 -1 10 0 10 1 10 2 Frequency (rad/sec) Figure S14.21c Solution for part c) using Bode plot 14-22 From modifying the solution to the two tanks in Section 6.4 ... R1 (A1 R1 A2 R2) s R2 1 R2 s Y ... R1 = 0.5, R2 = 2, A1 = 10, A2 = 0.8 Gp (s) = 0.5 8s ... Solving, ... c = 1.369 rad/min ... Kcu = 10.96 ... For Gv = Kv = 2.5, For Gm = 1.5 ... Substituting ... c = 1.369 rad/min, From Eq. 14-61 ... PM ... Υποκατάσταση ... 8.52 ... 0.3 ... 2.0 ... Using Eq. 12-11, the PI settings are Kc = 1.34, τI = 3.86 min ... Using Eq. 14-58 ... Using Eq. 14-56 ... AR OL ... 3.825 ... 0.7362 14-24 1.5 ... Kc 2 ... 0.5 ... 2 ... For the profit margin: GM = 1/Kc = 1.358 Resolve Eq.(14-61) for PM = 180 + ... = ... 0.625 ... Replacement in Eq. 14-57 gives ... g = –172.7° Therefore , PM phase margin = 180 + g = 7.3° 14.23 a) K = 2, c = 1, . - = 0.2 , oc=0.3 Using Eq. 12-11, the PI settings are Kc = 1.34, τI = 3.86 min ... Substituting for ... gives Kc = 1.34. Then, from Eq. 14-56, Kc · AR OL ... Closed loop response ... The controller designed in part (a) (Direct Composition) provides better performance by giving a first-class answer. Part (b) the controller produces a large excess. 14-26 14.24 a) Use Eqs. 14-56 and 14-57 AR OL ... Diagram Bode ... Figure S14.24a Plot bode by Use of eq.14-61 ... PM = 180° – 30° = 150° From the observatory versus ... 1.72 rad/min 14-27 2 10 From plot of Since AR OL = 1 ... Kc From the plot of Ac ... AR OL ... 0.144 ... 0.3 ... 6.94 0.144 From the plot of ... 0.44 0.44 ... 14-28 Based on ... Therefore Kc = 0.0326 Kc = 0.326 From Eq. 14-60, GM = 1/Kc = 3.07 Bode Diagram ... Frequency (rad/sec) Figure S14.24b Solution for part b) using bode plot 14-28 2 10 Bode Diagram 2 10 1 AR/Kc 10 0 10 -1 10 -2 10 -3 10 ... Frequency (rad/sec) Figure S14.24c Solution for part c) using Bode plot 14.25 a) Schematic diagram: TC Hot Liquid TT Cold Liquid Mixing Point 14-29 Sensor 2 10 Block Diagram: TR + Valve Controller Mixing Process Gc Gv Gp - Transfer Line GTL Gm b) GvGGGm = Km = 6 ma/ma GTL = e-8s GOL = GvGGGMGL = e-8s / (0.5s + 1) ... c) The critical frequency corresponds to an open loop gain phase of the phase angle -180° ... Ziegler-Nichols decomposition ratio settings 14/ PI controller: Kc = 0.45 Kcu = 0.45(0.167) = 0.45 Kcu = 0.45(0.167) = 0.075 ... 0.0826 ... d) Ziegler-Nichols PI είναι Kc = 0.45 Kcu ... 1.2 Therefore ... 0.75 ... Find Pu: Pu = 2 oc = π/8 rad/s ... Note that for this unusual procedure, ar process is independent of frequency ... The NOTE: The MATLAB version of PID control uses the following controller settings: ki=Kc/I and kd=Kc·D. d) Derivative control action improves closed loop response by reducing settlement time, at the expense of a more oscillating response. 14.26 Kcu and · c are taken using Eqs. 14-56 and 14-57. Including the GF filter in these equations gives ... Kcu = 4.251 Kcu = 3.536 for · F = 0 · F = 0.1 · F = 0.1 Therefore, ... decreases by 17% ... filtering the measurement leads to worse control performance ... ∠G(s) · 0.012 (0.0017s) o 1) Gm(s) · 0.12 (0.0 0.0.0) 024s 1) Use Eq. 14-61 ... tan-1(0.083) ... tan-1(0.024) Resolve, ... c = 18.19 rad/min Use Eq. 14-60 14-32 ... 5.624 ... Therefore, ... 4.2 ... Kcu = 12.97 Pu = 2π/ ... = 0.345 λεπτά Χρησιμοποιώντας τις ρυθμίσεις Ziegler-Nichols PI είναι Kc = 0.45 Kcu = 5.84 ... Ι = Pu/1.2 = 0.288 λεπτά ... Υποκατάσταση ... 14-33 ... 14-34 ... Kc = 14.50 and ... In this way: ... = 11.78 rad/min Substituting into Eq. 14-57 gives ... = 1 ... 5.624 ... 0.0095 ... 14-34 ... 0.651 14-33 Using Eq. 14-60, GM = 1/Ac = 1.54 Using Eq. 14-56 for ... gives AR OL ... 11.78 rad/min Using Eq. 14-57 gives ... –166.87 Using Eq. 14-61 PM = 180° + ... = 13.2 ... 14.28 From exercise 14.28, G(vs) · 5.264 0.0083 s 1 2 0.0325 ... 0.0162 ... (0.024s 1) ... Therefore, the closed loop transfer function is the PI controller is GC ... GC Gv G GGL ... 6.3175 0.21061 14-35 1.0 · 1 ... 0.16 ... Therefore, the profit margin is GM = 1/Ac = ... Actual axis Figure S14.28a The Nyquist plot of the system ... Given by fitting, At the same time ... 0.5 ... Therefore, the profit margin is GM = 1/0.5 = 2 14-36 Chapter 15 15.1 For uniformly flow control a Gd · Kd ... d ... which may differ more from Kp in Eq. 15-2. Because the new Kc depends on both d and you, 15.2 By default, the ratio station sets um = KR · um = um o KR ... Gd ... Gp ... Gm ... If Kp ... Kd ... Kv ... This implies that the Kd(s) ... Χρησιμοποιώντας ... έλεγχος τροφοδοσίας είναι ... Σχεδίαση feedforward ελεγκτή ... 15-21 ... κατάστασης ... GF (ex) ... Kd Kt Kv Kp ... Λογάριθμος ... Ziegler-Nichols ... (b) the set point change ... K ... 1/(1 o 1s) ... To find gains Kc and ... from Eq. 15-28. The steady state ... flow rate of the heat exchanger ... dynamic feedforward controller is given by Eq. 15-21 ... Mass balance in tank 1 is A1 dh1 ... q1 q 2 – C1 h1 dt where A1 is the transverse area of tank 1. Linearization and adjustment q 2 ... leads to A1 dh1 ... q1 ... h1 ... dt 2 h1 Take Laplace transformation, H 1 ... For tank 2 ... The mass balance in tank 2 is A2 dh2 ... q3 q4 ... q5 dt Using deviation variables, setting ... and taking the transformations Laplace gives ... A2 H2 (s) ... Q3(s) Q5(s) ... G Gc Gv ... Km ... Feedforward controller design A dynamic model will be developed based on the following assumptions: 1. Perfect mixing 2. Isothermal function 3. Fixed volume Component balances: dcA ... q(c Ai – c A) – V (k1c A – k2cB) dt dc V B ... qcB V (k1c A – k2cB) dt V Linear, dc ... a11cA ... a12cB ... b1q ... dc ... V B a21c ... a22cB - b2 q dt V 15-9 (1) (2) (1) a11 ... q – k1 , V a12 · k2 a21 · k1 , d , q , V q – k2 V c b2 ... B V a22 ... b1 · cA i – c A , V (3) c Ai – cA and c A indicates the nominal constant state value Take Laplace transforms and resolves after replacing the first equation for CA ... (s) in the second equation. The result is: ... Solving for ... and Q(s), ... Rearrangement and substitution of (3): ... For Gm ... Km ... b) Substitute ... G(s) ... c) Design of the feedforward controller ... Reverse or direct action controller; From ch. 11, we know that for the closed loop system to be stable, Kc ... Kv ... Kp ... Km > 0 The available information indicates that Kv ... Kp ... assuming that q is still the manipulated variable. So Kc should have the same mark as Kp and we need to specify the mark of Kp. From (5) we have Kp ... a21b1 – b2 a11 ... a22 – a12a12 Substitute from (3) and simplify to get ... ...Therefore, the (4 s 1) ... 6 Because both the numerator and the denominator terms are positive, Kp ... 0, and Kc > 0. Finally, in View So Kc should be positive. Conclusion: The feedback controller should act in reverse action. c) The advantages of using a fixed-state controller are that the calculations are quite simple and a detailed process model is not required. The downside is that the control system may not perform well during transitional conditions. To decide whether or not to add dynamic compensation, we need to know whether the controlled cb variable is affected faster, or more slowly, by the cAi disturbance variable than it is by the manipulated variable, q. If the response times are quite different, then dynamic compensation of 15-11 could be beneficial. A unstable state model (or experimental data) will be required to resolve this issue. Then, 15.9 The block diagram for the feed forward and feedback control system is shown in Figure 15.12. a) No (b) Power controllers are required from ... Gv ... 0.75 psi/mA ... Gv ... 2.5 Gv(s) ... 2.5 ... e-2s ... 1 ... Gm ... Kc ... 0.75 ... 0.333 ... 4.71 ... The ideal dynamic feedforward controller is given by Eq. 15-21: Gf ... Gd K I PGt GvGp (15-21) Substituting the individual transfer functions into Eq. 15-21 gives, G f (s) ... 0.417 ... 0.0833s ... 1 (1) The static (or steady-state) version of the controller is simply a gain, Kf: Kf = - 0.417 (2) 15-12 Note that G(s) in (1) is physically unrealizable. In order to produce a naturally liquid dynamic controller, the non-performing controller in point 1 shall be approached by a lead delay unit. In proportion to example 15.5: G f (s) ... –0.417 0.1833s o 1 0.0183s 1 (3) Equation 3 resulted from (1) from (i) omitting the delay in time, (ii) the addition of the 0.1-minute delay to the delivery time constant and (iii) the introduction of a short time constant ... =0.1833 0.0183 for ... = 0.1. (c) Feedback controller Define G as ... 25 ... 2.6 x10 ... 0.0833 ... 4.71 ... 0.758 ... G ... GP GvG p Gm First, approximate G as a FOPTD model, G using Skogestad's half-rule method in Section 6.3: ... 4.71 ... 0.5 ... 0.0833 ... 4.75 min ... 1 ... 0.4 ... 0.0833 ... 0.14 min Thus, ... 0.01 ... 0.4 ... 0.14 ... 0.5 ... 1 The ITAE controller settings are calculated as ... Kc ... 0.977 ... 0.680 ... 0.14 ... 0.859 ... 4.752 ... 0.977 ... Kc ... 134 ... 0.680 ... 0.142 min Combined feedforward-feedback control This control system consists of the dynamic feedforward controller of point (b) and pi controller of point (c). Closed loop responses to a +0.2 step change to x1 for the two forward feed controllers are displayed in the Shape. S15.9a. The dynamic forward power controller is superior to the static power controller 15-13, because both the maximum deviation from the specified point and the precipitation time are smaller. Figure S15.9b shows the combined feed feedback control system provides the best control and is superior to the PI controller. A comparison of the Figs. S15.9a and S15.9b shows the addition of feedback control significantly reduces settlement time due to the very high value of Kc that can be used because the time delay is too short. (Note that ... 0.14/4.75 = 0.0029.) Image. S15.9a. Comparison of static and dynamic power controllers for step disturbance +0.2 to x1 in t = 2 minutes x 8 x 10 -3 7 6 5 FB FF-FB 4 3 2 1 0 -1 0 2 4 6 8 10 12 14 16 18 20 time (min) Image S15.9b. Comparison of feedback control and feedforward feedback controllers for step disturbance +0.2 x1 at t = 2 min. 15-14 15.10 a) For steady-state conditions, Gp=Kp, Gd=Kt, Gv = Gm = Gt = 1 (Bring up Eq. 15-21 ... Gd ... Gf G p ... = G p (1)(1)(1) b) From Eq. 15-21, ... 0.5e ... 80s and Gd ... 0.75 ... (s) ... 10s 60s ... 1 0.5 ... 10s ... 1 ... 0.5 ... 1 ... 1 Using Table 12.1, a PI controller is obtained from G, Kc ... 95 K c 0.95 K p 2 ... I ... (30 ... 20) ... 0.5 ... As shown in ... Fig S15.10a ... ... αρνητικό ... υπερβολικά σε ... The FF controller ... Σχήμα S15.10a ... Με το FF-FB ... 15-16 15.11 ... the ideal FF controller design equation is given by, GF ... Gd Gt Gv G p (15-21) But, and gives, GF ... Gv ... GP ... Υποκατάσταση ... GF 2(s + 1) ... (3s 1)(5 ...) In this case, you need a physically realizable controller ... Therefore, ... 2(3s ... 1)(5s ... 1) ... 1.5 ... 3s 1 ... 1 ... 20 ... 0.2 ... 1 ... 15.12 a) ... 1.5 minutes ... 1 GF controller ... S15.12a; the FF-FB controller provides the best control with a small maximum deviation and without compression. The oscillation due to the feedback controller can be depreciated using a larger value of the design parameter, c. 15-19 Figure S15.12a: Controller Comparison for step change in fuel gas purity from 1.0 to 0.91 t = 10. Top: full scale; Bottom: enlarged scale. 15.13 Fixed-state feedforward design ... 0 · q1 · q3 (1) 0 · q3 q2 · q4 · q5 (2) Resolve (4) for x3 q3 and substitute in (3), ... Substituting ... x2 q2 ... x4 q4 ... Rearranging ... (7) Replace numeric values: ... Note: If winnings of the transmitter and control valve are available, then an expression relating to the output signal of the power controller, p(t), with the measurements, Xm(t) and q5m(t) can be developed. Dynamic compensation: It will be required due to the additional dynamic delay introduced by the tank on the left side. Current disturbance 5 affects x3, while q5 does not. 15.14 The three xD control strategies are compared to Figs. S15.14a-b for the disturbance of the packing in the composition of the feed. The FF-FB controller is slightly superior because it minimizes the maximum deviation from the specified point. Note that the design of the PCM power controller ignores the two time delays, which are quite different. Thus, the feedforward controller over-fixes and is not as effective as it could be. 15-21 Fig. S15.14a. Comparison of feedback control and forward control and feedback control for step change in feed composition from 0.5 to 0.55 to t = 0. 15-22 Chapter 16 16.1 The difference between systems A and B lies in the dynamic lag of measurement elements Gm1 (primary loop) and Gm2 (secondary loop). With a faster measuring device in A, better control action is achieved.

S15.14a. Comparison of feedback control and no control for step change in feed composition from 0.5 to 0.55 to t = 0. Image. S15.14b. Comparison of forward feed control and feedback control for step change in feed composition from 0.5 to 0.55 to t = 0. 15-22 Chapter 16 16.1 The difference between systems A and B lies in the dynamic lag of measurement elements Gm1 (primary loop) and Gm2 (secondary loop). With a faster measuring device in A, better control action is achieved. Additionally, for an overlay control system to function properly, the secondary control loop response must be faster than the primary loop. Therefore, system A should be faster and yield better closed loop performance than B. Because the Gm2 in system B has a noticeable delay, cataract control has the potential to improve the overall closed loop more than for system A. Little improvement in system A can be achieved by controlling cataracts versus conventional feedback. Comparisons are shown in the S16.1a/b. PI controllers are used in the outer loop. Pi controllers for both system A and system B are designed on the basis of Table 12.1 (c) . P controllers are used in internal loops. Due to the different dynamics, the proportional gain of system controller B is about a quarter as large as the system controller gain A: Kc2 = 1 System B: Kc2 = 0.25 · I=15 · I=15 Kc1=0.5 Kc1=2.5 0.7 Ca standard feedback 0.6 0.5 Output 0.4 0.3 0.2 0.1 0 0 10 20 30 40 50 60 70 80 90 100 time Figure S16.1a System A. Comparison of D2 responses (D2=1/s) for PI control and conventional control. Solution Manual for Process Dynamics and Control, 4th Edition Copyright © 2016 by Dale E. Seborg, Thomas F. Edgar, Duncan A. Mellichamp, and Francis J. Doyle III 16-1 Compared to the two numbers, it appears that the typical feedback results are essentially the same, but the sequential response to System A is much faster and has a much less absolute error than what to control waterfall of B 0.7 Cascade Standard feedback 0.6 0.5 Output 0.4 0.3 0.2 0.1 0 0 10 20 30 40 50 time 60 70 80 90 100 Figure S16.1b System B . Comparison of D2 responses (D2=1/s) for sequential control and conventional PI control. Figure S16.1c Block diagram for System A 16-2 Figure S16.1d Block diagram for System B 16.2 a) The transfer function between Y1 and D1 is Y1 D1 Gd 1 Gc 2Gv 1 Gc1 G p Gm1 1 Gc 2GvGm 2 and that between Y1 and D2 is G p Gd 2 Y1 D2 1 Gc 2Gv Gm 2 Gc 2GvGm1Gc1G p using Gv 5 s 1 , Gd 2 1 , 16-3 Gd 1 1 , 3s 1 4 , Gm1 0.05 , (2s 1)(4s 1) For Gc1 = Kc1 and Gc2 = Kc2, we obtain G m 2 0 .2 Gp 8s3 (14 8Kc 2) s 2 (7 6 Kc 2) s Kc 2 1 Y1 D1 24s 4 (50 24 Kc 2) s3 [10 Kc 2 (9 3Kc1)]s (35 26 Kc 2) s 2 o Kc 2 (1 o Kc1) 1 Y1 4(s 1) · 3 2 D2 8s (14 o 8K c 2) s The following figures show the following step load responses for Kc1=43.3 and Kc2=25. Note that both responses are constant. Keep in mind that the critical gain for Kc2=5 is Kc1=43.3. The increase in Kc2 stabilizes the controller as planned. -3 1 12 x 10 0.8 10 0.6 8 0.4 6 Output output 0.2 0 4 -0.2 2 -0.4 0 -0.8 0 5 10 15 time 20 25 30 -2 0 0 5 10 10 15 time 20 25 30 Figure S16.2a Responses for unit load change to D1 (left) and D2 (right) b) The typical equation for this system is 1+Gc2GGGm2+Gc2GGGm1Gc1Gp = 0 (1) Let Gc1=Kc2 and Gc2=Kc2. substituting all transport functions in (1), we have 8s3 o (14 o 8Kc 2) s 2 (7 o 6 Kc 2) s o K c 2 (1 k c1) o 1 · 0 (2) Now we can use the direct substitution: −8 · 3 o 7 o 6 K c 2 · · j – (14 o 8K c 2) · 2 o K c 2 (1 k c1) o 1 · 0 (3) j : 8 7 6 K c 2 200 1 1 Therefore, for normal (positive) values Kc1 and Kc2, 16-4 24 Kc 2 2 66 Kc 2 o 66 Kc 2 o 45 4Kc 2 The results are shown in the table and picture below. Note the almost linear variant of Kc1 ultimate with Kc2. This is because the right side is very nearly 6 Kc2 +16.5. For larger Kc2 values, the stability margin in Kc1 is higher. There do not appear to be non-linear effects of Kc2 for Kc1, especially at high Kc2. Kc1,u · There is no theoretical ceiling for Kc2, except that large values can cause the valve to kort for small set-point or load changes. Kc1.u 33.75 34.13 38.25 43.31 48.75 54.38 60.11 65.91 71.75 77.63 83.52 89.44 9 5.37 101.30 107.25 113.20 119.16 125.13 131.09 137.06 160.00 140.00 120.00 Kc1, final Kc2 1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 20 100.00 80.0 0 160.00 40.00 20.00 0.00 0 5 10 15 Kc2 Figure S16.2b Kc2 effect on critical profit Kc1 c) With built-in action in the inner loop , Gc1 · K c1 1 Gc 2 · 5 · 1 o · 5s Replace all transfer functions in the characteristic equation yields 1 5 1 o 5-1 o (0.2) o 5-1 o (0.0.1 (0.2 o 5-1 o (0.0.1 05) K c1 · 5s s o 1 · 5s s o 1 4 · 0 (4 s o 1) (2 s o 1) Rearrangement gives 16-5 20 8s 4 54s 3 o 45s 2 o (12 o 5K c1) s o K c1 o 1 · 0 Now we can use direct substitution: −54 · 3 o12 o 5K c1 · j o 8 · 4 – 45 · 45 · 4 2 k c1 · 1 · 0 j : −54 · 3 · 12 o 5 K c1 · · 0 8 · 4 – 45 · 2 k c1 · 1 · 0 Solve the above equations , and we will get: K c1,u · 44.2 The final K c1 is 44.2, which is close to the result, as for analog only control of the secondary loop. With built-in action only in the outer loop, 1 Gc1 · K c1 · 1 o · 5s Gc 2 · 5 Replacing the transfer functions in the characteristic equation. 1 o 5 5 1 4 (0.2) o 5 (0.05) K c1 · 1 o 0 s o o 1 s o1 · 5s (4s 1)(2s 1) 8s 4 o 54s 3 o 37s 2 o (6 o 5K c1) s o K c1 · 0 Now we can use the direct substitution: −54 · 3 o 6 o 5K c1 · · j o 8 · 4 − 37 · 2 o K c1 · 0 j : −54 · 3 o 6 o 5 K c1 · · 0 8 · 4 − 37 · 2 k c1 · 0 Resolve the above equations, and we receive: K c1,u · 34,66 Therefore, Kc10) for the flow control loop to function properly. Two alternative control strategies are taken into account: Method 1: use a default feed flow rate when Pcc > 80% Let: Pcc = output signal from the synthesis controller (%) ~ Fsp (internal) adjustment point for the power flow controller (%) Control strategy: 16-32 ~ ~ If Pcc > 80%, Fsp · Fsp, low ~ where Fsp , low is a set default flow rate that is lower than normal ~ value, Fsp nom . Method 2: Reduce the power flow when > 80% Control Strategy: ~ ~ If Pcc > 80%, Fsp · Fsp nom − K (Pcc - 80 >%) Pcc HS + - K - + ~ Fsp 80 % ~ Note: A check should be made to ensure that 0 · Fsp · 100% ALTERNATIVE selective control system B.A is recommended: 16-33 Figure S16.22. Recommended selective control system Both control valves are A-O and the transmitters are direct action, so the controller must be reverse action. When the output concentration decreases, the controller output increases. Therefore, this signal cannot be sent directly to the power valve (it will open the valve). Using a high selector that selects the highest of these signals can solve the problem .- Flow transmitter .- Output concentration controller Therefore, when the signal from the output controller exceeds 80%, the selector holds it and sends it to the flow controller so that the feed flow rate decreases. 16-34 16.23 ALTERNATIVE delay A.Time.- Compensation for time delay of use, e.g., Smith forecast waste concentration variable. Changes in the pH of the tank occur due to these unforeseen changes. Process gain changes also (c, f. literature curve for strong acid-strong base) Variable waste flow rate.- Use FF control or qbase ratio in qwste. Measure qbase .- This suggests that you may want to use cataract control to compensate for upstream pressure changes, etc. Alternative B.Several advanced control strategies could provide improved process control. A selective control system is usually used to control pH in waste water treatment. The proposed system is shown below (pH T = pH sensor, pH C = pH controller) Figure S16.23. Proposed selective control system. where S represents a selector (< or >, to be specified) This combination uses multiple variables to handle to control a single-process variable. When the pH is too high or too low, a signal shall be sent to the selectors either in the waste stream or in the base flow controllers. The exact configuration of the system depends on the transmitter, controller and valve gains. In addition, a Smith forecast indicator for the pH controller is proposed due to the long time delay. There would be other possibilities for this process, such as an adaptive control system or a cataract control system. However, the above system can be pretty good Necessary information: 16-35 .- Descriptions of measuring devices, valves and controllers; direct action or reverse action. . - Model of the procedure for applying the Smith Prediction Index 16.24 To change the adjustment point, the closed loop transfer function with built-in controller and fixed-state procedure (Gp Kp) είναι: 1 K G G I s P KP 1 C P Y Ysp 1 G G 1 I s K P 1 K I C P s 1 I s P KP Ως εκ τούτου επιτυγχάνεται απόκριση πρώτης τάξης και μπορεί να επιτευχθεί ικανοποιητικός έλεγχος. Για μεταβολή διαταραχής (Gd = Gp): Y Gd KP K (s) I P I D D G G 1 o 1 K P · I s o K P · I C P s o1 · I s KP Therefore, a first order response is also received for a change of disorder. 16.25 MV: INSULIN CV pump flow rate: DV body sugar level: food intake (sugar or glucose) The standard PID control algorithm could be used to provide a basic level of control. However, it may be subject to saturation in order to maintain blood glucose within the stated limits. Feedforward control could be used if the effect of meal intake (disorder) can be quantified depending on its glucose level. Then, the insulin injection can predict the effect of the meal by taking preventive measures before the change in blood glucose becomes sensed. A trap of an FF/FB control could be that high insulin pump flow rates may be required in order to keep blood glucose within the desired range, and pump flow rate can be saturation. Another improvement would be adaptive control, which would allow the controller to automatically tune for a given man in order to obtain a better response (each person's body chemistry is different). One drawback of adaptive control 16-36 is that it can be too aggressive and cause rapid changes in blood glucose. A less aggressive adaptive controller could use profit planning, where a higher controller gain is used when the blood glucose level goes too high or too low. 16.26 If the feed temperature is too high, the slave controller will feel the temperature increase and increase the signal in the refrigerant valve, which will increase the flow of the refrigerant to reduce the temperature of the food. The master controller will feel a slight increase in temperature in the reactor and increase the specified point of the slave controller, which in turn will increase the flow rate of the refrigerant for a second time. In this case, both the slave and the master controller work together to counteract the disorder. As a result, the disturbance is treated quickly and the temperature of the reactor is only slightly affected. If the feed flow rate is too high, the temperature of the feed exiting the heat exchanger will increase. The slave controller will feel this and act as above by increasing the refrigerant flow rate. The increased flow rate of the higher temperature supply to the reactor will most likely increase the temperature of the reactor and the main controller will change the specified point of the slave controller accordingly. Again the master controller The slave controller is working together to neutralize the disorder. 16.27 16-37 Figure S16.7a Control waterfall exothermic chemical reactor Figure S16.7b Diagram block of control over an exothermic chemical reactor D1 : Reactor temperature D2 : Cooling water D3 : Temperature of the reactor wall The control system measures the temperature of the reactor wall to collect information on temperature gradients in the contents of the tank, tank, in a designated spot, and regulates cooling water makeup. The main advantage of the new cataract control strategy is that the temperature of the reactor wall is close to a possible temperature disturbance of 16-38 gradients in the contents of the tank and the associated feedback loop can react quickly, thus improving the closed loop response. 16.28 For a linear algebraic one-input-two-output model displayed in Eqs. (1)~(2): y1 · K12u1 o b1 (1) y2 · K 21u2 o b2 (2) The y1 output can reach the y1sp adjustment point with the u1 setting based on Eq. (3): y1sp − b1 u1 · (3) K12 But for y2 production, it is determined by the Eqs combination. (2) and (3), and cannot be arbitrarily determined which leads to compensation. 16-39 Chapter 17 17.1 Using Eq. 17-9, the filtered values of xD are shown in Table S17.1 time(min) 0 1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 20 0 0.495 0.815 1.374 0.681 1.889 2.078 2.668 2.533 2.908 3.351 3.336 3.564 3.419 3.917 3.884 3.871 3.924 4.300 4.252 4.409 0.8 0 0.396 0.731 1.245 0.794 1.670 1.996 2.534 2.533 2.833 3.247 3.318 3.515 3.438 3.821 3.871 3.871 3.913 4.223 4.246 4.376 0.5 0 0.248 0.531 0.953 0.817 1.353 1.715 2.192 2.362 2.635 2.993 3.165 3.364 3.392 3.654 3.769 3.820 3.872 4.086 4.169 4.289 Table S17.1. Un filtered and filtered data. In order to obtain the analytical solution for xD, the set(s) F(s) · 1 in the given transport operation, so that 1 5 5 1 F (s) · 5 · − 10s 1 s(10s) · s o 1 10 Receiving reverse transformation Laplace X D (s) · xD(t) = 5 (1 − e-t/10) A graphical comparison appears in Fig. S17.1 Solution Manual for Process Dynamics and Control, 4th Edition Copyright © 2016 by Dale E. Seborg, Thomas F. Edgar, Duncan A. Mellichamp and Francis J. Doyle III 17-1 4.5 4 3.5 3 XD 2.5 2 1.5 1 noisy alpha data = 0.5 alpha = 0.8 analytical solution 0.5 0 0 2 4 6 8 10 12 time (min) 14 16 18 20 Fig S17.1. Graphical comparison for noisy data, filtered data and analytical solution. As it decreases, the filtered data give a smoother curve than in the case without a filter (=1), but this noise reduction is negotiated by increasing the deviation of the curve from the analytical solution. 17.2 The exponential filter output in Eq. 17-9 is yF (k) · ym (k) o (1 − o) yF (k 1) (1) Replace k from k-1 to Eq. 1 gives yF (k − 1) · ym (k − 1) o (1 − ·) yF (k − 2) Replacing yF (k − 1) from (2) to (1) gives yF (k) · ym k) (1 −) ym (k − 1) μ (1 − o) 2 yF (k − 2) Sequential substitution of yF (k − 2) , yF (k − 3) ,... gives the final form 17-2 (2) k −1 yF (k) · (1 −)i · ym (k − i) o (1 − ·) k yF (0) i · 0 17.3 Table S17.3 lists the un filtered production and, from Eq. 17-9, the filtered data for periods 1.0 and 0.1. Notice that for the sampling period of 0,1, the and filtered outputs were taken in 0.1 time increments, but refer only to 1.0 intervals to maintain brevity and facilitate comparison. The results show that for each t value, the data become smoother as it decreases, but at the expense of the delay relative to the average production y(t)=t. In addition, the lower sampling period improves filtering by providing smoother data and less delay for the same value. t 0 1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 20 · =1 0 1.421 1.622 3.6 206 3.856 4.934 5.504 6.523 8.460 8.685 9.747 11.499 11.754 12.699 14.470 14.535 15.500 16.987 17.798 19.140 19.575 =0.8 0 1.137 1.525 2.870 3.659 4.679 5.339 6.286 8.025 8.553 9.508 11.101 11.624 12.484 14.073 14.442 15.289 16.647 17.568 18.825 19.425 t=1 =0.5 0 0.710 1.166 2.186 3.021 3.977 4.741 5.632 7.046 7.866 8.806 10.153 10.954 11.826 13.148 13.841 14.671 15.829 16.813 17.977 18.776 =0.2 0 0.284 0.552 1.083 1.637 2.297 2.938 3.655 4.616 5.430 6.293 7.334 8.218 9.115 10.186 11.055 11.944 12.953 13.922 14.965 15.887 =0.8 0 1.381 1.636 3.227 3.916 4.836 5.574 6.571 8.297 8.688 9.741 11.328 11.770 12.747 14.284 14.662 15.642 16.980 17.816 19.036 19.655 t=0.1 =0,5 0 1.261 1.678 3.200 3.973 4.716 5.688 6.664 8.044 8.717 9.749 1 1.078 11.778 12.773 14.051 14.742 15.773 16.910 17.808 18.912 19.726 =0,2 0 0,877 1.647 2.779 3.684 4.503 5.544 6.523 7.637 8.533 9.544 10,6 58 11.556 12.555 13.649 14.547 15.544 16.605 17.567 18.600 19.540 Πίνακας S17.3. Un filtered and filtered output for sampling periods 1.0 and 0.1 17-3 Graphic comparison: 20 18 16 14 y(t) 12 10 8 6 · · · · 4 2 0 0 2 4 6 8 10 times, t 12 14 16 18 20 Figure S17.3a. Graphic comparison for t = 1.0 20 18 16 14 y(t) 12 10 8 6 · · · · · · 4 2 0 0 2 4 6 8 10 time, t 12 14 16 Figure S17.3b. Graphic comparison for t = 0.1 17-4 18 20 17.4 Using Eq. 17-9 for ◊ = 0.2 and · = 0.5, Eq. 17-18 for N * = 4, and Eq. 17-19 for y =0.5 , the results are listed and plotted below. · =1 0 1.50 0.30 1.60 0.40 1.70 1.50 2.00 1.50 t 0 1 2 3 4 5 6 7 8 (a) (a) 2 0 0.30 0.30 0.56 0.53 0.76 0.91 1.13 1.20 · =0.5 0 0.75 0.53 1.06 0.7 3 1.22 1.36 1.68 1.59 b) N*=4 0 0.38 0.45 0.85 0.95 1.00 1.30 1.30 1.30 1.30 1 1.40 1.68 (c) y=0.5 0 0.50 0.30 0.80 0.40 0.90 1.40 1.90 1.50 Table S17.4. Un filtered and filtered data. 2.5 2 y(t) 1.5 1 · · · · · · · · · 0.5 N*=4 y=0.5 0 0 1 2 3 4 times, t 5 6 7 8 Figure S17.4. Graphical comparison for filtered data and uneded data. 17-5 17.5 Configuration: 1 Gp = 2s+1; d(t) = 1 + 0,2 sin(t) · tF = 0 (without filtering) or 3 To do this problem, create the Simulink diagrams below. Note that the filter is represented by a first order transfer operation with a time constant of fine tF. This appears by performing the Laplace transformation of equation 17.4 in the book. Figure S17.5a. Diagram of blocks when using a filter on the output with a time constant of 3 minutes. A sinusoidal wave of frequency 1 and width 0.2 is the input. Figure S17.5b. Block diagram when no filter is used on the output. A sinusoidal wave of frequency 1 and width 0.2 is the input. Simulation of the diagram for 50 minutes: 17-6 1.4 Filtered without filter 1.2 Closed loop response 1 0.8 0.6 0.4 0.2 0 0 5 10 15 20 25 30 Time/ minute 35 40 45 50 Figure S17.5c. First-order process response to disturbance, d(t)=1+0.2sin(t), with and without exponential filter. From figure S17.5c, we can see that the filter will significantly reduce oscillation with the cost of causing a lag in the first 10 minutes. 17.6 Y(s) · 1 1 X (s) · s o1 s μ1 s , then y(t) = 1 − e-t For noise level 0.05 units, many different values are tested in Eq. 17-9, as shown in figure. S17.6a. While the filtered output for ◊ = 0.7 is still quite noisy, that for ◊ = 0.3 is very sluggish. Thus, ◊ = 0.4 seems to offer a good compromise between noise reduction and adding lag. Therefore, the designed first-class filter for the noise level 0.05 units is · = 0.4, which corresponds to · F = 1.5 according to Eq. 17-8a. Noise level = 0.05 17-7 1.4 · · · · · · · · · · 1.2 1 y() (t) 0.8 0.6 0.4 0.2 0 0 2 4 6 8 10 t 12 14 16 18 20 Figure S17.6a. Digital filters for noise level = 0.05 Noise level = 0.1 1.4 1.2 1 y(t) 0.8 0.6 · · · · · · · · · · · · · 0.4 0.2 0 0 5 10 15 20 t Figure S17.6b. Digital filters for noise level = 0.1 Noise level = 0.01 17-8 1.4 1.2 y(t) 0.8 · 0.6 0.4 0.2 0 0 0 5 10 15 20 t Figure S17.6c. Response for noise level = 0.01; no filter required. Similarly, for the noise level of 0.1 units, a good compromise is · =0.2 or · F = 4.0, as shown in the Figure. S17.6b. However, for the noise level of the units 0.01, no filter is required as shown in the Figure. S17.6c. thus =1.0, F = 0 17.7 y(k) = y(k-1) 0.21 y(k-2) + u(k-2) k 0 1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 u(k) 1 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 u(k-1) 0 1 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 Plotting this results 17-9 u(k-2) 0 0 1 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 0 y(k) 0 0 1.00 1.00 0.79 0.58 0.41 0.29 0.21 0.14 0.10 0.07 0.05 0.03 0.02 0.02 0.01 0.01 0.01 0.00 1.2 1 0.8 y 0.6 0.4 0.2 0 0 2 4 6 8 10 k 12 14 16 18 20 Figure S17.7. Graphical simulation of the difference equation The constant state value of y is zero. 17.8 Using the Simulink and STEM routine to convert the continuous signal into a pulse range, 12 10 8 Tm'(t) 6 4 2 0 0 5 15 20 25 time 30 35 40 45 50 Figure S17.8. Discrete time response for the change Therefore, the maximum value of the recorded temperature is 80.7°C. point is reached in t = 12 minutes. 17-10 17.9 a) Y (z) 2.7 z −1 (z) 2.7 o 8.1 − 1 · U (z) z 2 − 0.5 z 0.06 z 2 − 0.5 z 0.5 z 0.5 z 0.5 z 0.5 z 0.06 and denominator by z2 Y (z) 2.7 z −2 o 8.1z −3 · U (z) 1 − 0.5 z −1 μ 0.06 z −2 Then Y (z)(1 − 0.5 z −1 μ 0.06 z −2) · z)(1 2.7 z −2 o 8.1z −3) or y(k) = 0.5y(k-1) − 0.06y(k-2) + 2.7u(k-2) + 8.1u(k-3) Simulation of the difference equation yields k 0 1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 20 u(k) 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 17-11 u(k-3) 0 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 y(k) 0 0 2,70 12,15 16,71 18,43 19,01 19,20 19,26 19,28 19 28 19,28 19,29 19,29 19,29 19,29 19,29 19,29 19,29 19,29 19,29 19,29 β)20 18 16 14 12 y 10 8 6 4 2 0 Simulink Difference Equation 0 2 4 6 8 10 k t 12 14 16 18 20 Figure S17.9. Simulink response to unit step change to u c) The constant state value of y can be found in the z =1 setting. In this way, y =19.29 This result is in line with the above data. 17.10 1 Gc(s) · 2 · 1 o · 8s replacement s (1-z-1)/t and accounting for t=1 2.25 − 2 z −1 Gc (z) · 2 · 1 · · 1 (1 − z −1) · 8(1 − z) Using Simulink-MATLAB, the simulation for changing a unit step on the e(t) controller error signal is shown in the Figure. S17.10 17-12 70 60 50 b(k) 40 30 20 10 0 0 5 10 15 k 20 25 30 Figure S17.10. Open loop response for unit step change 17.11 (a) Y (z) 5(z 0.6) · 2 U (z) z − z 0.41 Dividing both the numerator and the denominator by z2 Y (z) 5 z −1 o 3z −2 · U (z) 1 − z −1 o 0.41z −2 Then Y (z) (1 − z −1 · 0.41z −2) · U (z)(5 z −1 3z −2) or y(k) = y(k-1) − − 0.41y(k-2) + 5 (k-1) + 3u(k-2) b) Simulation of the difference equation yields 17-13 u(k) 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 3 4 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 c) u(k-1) 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 u(k-2) 0 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 1 y(k) 5 13.00 18.95 21.y 62 21.85 20.99 20.03 19.42 19.21 19.25 19.37 19.48 19.5 4 19,355 19,54 19,52 19,51 19,51 19,51 Using Simulink-MATLAB, the simulation for a unit step change in u odds 25 Difference equation Simul 20 15 y 10 5 0 0 2 4 6 8 10 k t 12 14 16 18 20 Figure S17.11. Simulink response to unit step change to u d) The constant state value of y can be found in the z =1 setting. In this way, y =19.51 This result is in line with the above data. 17-14 17.12 a) 1 1 − z −1 7 6 Output 5 4 3 2 1 0 0 1 2 3 3 3 4 4 5 Time b) 1 1 o 0.7 z −1 Exit 1 0.8 0.6 0.4 0.2 0 0 1 2 Time 1 1 − 0.7 g −1 3 2.5 2 Production c) 1.5 2 1 0.5 0 0 1 2 Time 17-15 1 (1 0.7 of)(1 0.3z 1) 1 Έποδος 0.8 0.6 0.4 0.2 0 0 1 2 3 4 5 3 4 5 5 Ρα ε) 1 0.5 of 1 (1 0.7 of 1) (1 0.3z 0.3z) 1 Production 0.8 0.6 0.4 0.2 0 0 1 2 Time f) 1 − 0.2 z −1 (1 0.6 z −1)(1 1 − 0.3z −1) 1 0.8 0.6 0.4 0.2 0 0 1 2 17-16 Conclusions: .- A pole in z = 1 causes instability. .- Poles only on a positive real axis give oscillation free answer. .- The Poles on the negative real axis give a oscillating response. .- Poles on the positive real axis mitigate the oscillating reactions. .. - Zeros on the positive real axis increase oscillations. .- Zeros closer to z = 0 contribute less to increased oscillations. 17.13 Using Simulink, the response to a unit setting point change appears in the Figure. S17.13a 1.8 1.6 1.4 Output 1.2 1 0.8 0.6 0.4 0.2 0 0 5 10 15 20 Time 25 30 35 40 Figure S17.13a. Closed loop response to unit adjustment point change (Kc = 1) Therefore the controlled system is stable. The final controller gain for this process lies in the test and error 17-17 8 7 6 Output 5 4 3 2 1 0 0 5 10 15 20 Time 25 30 35 40 Figure S17.13b. Closed loop response to unit adjustment point change (Kc =21.3) Then Kcu = 21.3 17.14 Using Simulink-MATLAB, these final gains are located: t = 0.01 2 1.8 1.6 1.4 Production 1.2 1 0.8 0.6 0.4 0.2 0 0 1 2 3 Time 4 5 6 Figure S17.14a. Closed loop response to unit adjustment point change (Kc =1202) t = 0.1 17-18 2 1.8 1.6 1.4 Output 1.2 1 0.8 0.6 0.4 0.2 0 0 5 10 15 Time image S17.14b. Closed loop response to unit adjustment point change (Kc =122.5) t = 0.5 2 1.8 1.6 1.4 Output 1.2 1 0.8 0.6 0.4 0.2 0 0 5 10 15 Time image S17.14c. Closed loop response to unit adjustment point change (Kc =26.7) Therefore t = 0.01 t = 0.1 t = 0.5 Kcu = 1202 Kcu = 122.5 Kcu = 26.7 As noted above, reducing the sampling time makes the allowed controller gain increments. For small prices t, the final profit is large enough to guarantee a wide range of stability. 17-19 17.15 Using Simulink-MATLAB Kc = 1 1.4 1.2 Output 1 0.8 0.6 0.4 0.2 0 0 5 10 15 20 25 times 30 35 40 45 50 Figure S17.15a. Closed loop response to unit adjustment point change (Kc =1) Kc = 10 1.8 1.6 1.4 Output 1.2 1 0.8 0.6 0.4 1 0.2 0 0 0 5 10 15 20 25 time 30 35 40 45 50 Figure S17.17.17.17 15b. Closed loop response to unit adjustment point change (Kc =10) 17-20 Kc = 17 2.5 2 Output 1.5 1 0.5 0 -0.5 0 5 10 15 20 25 times 30 35 40 45 50 Figure S17.15c. Closed loop response to unit adjustment point change (Kc =17) the maximum controller gain is Kcm = 17 17.16 Gv(-a) = Rv = 0.1 ft3 / (min)(ma) Gm(s) = 4 0.5s 1 To obtain Gp(1) s), write the mass balance for the tank as A dh · q1 q1 q 2 − q3 dt Using deviation variables and taking Laplace convert As H (s) · Q1 (s) q2 (s) − Q3 (s) Therefore , G p(s) · H (s) −1 −1 · Q3 (s) Up to 12.6s Using Simulink-MATLAB, Kc = -10 1.4 1.2 1 y(t) 0.8 0.6 0.4 0.2 0 0 5 10 15 20 25 time 30 35 40 45 50 Figure S17.16a. Closed loop response to unit adjustment point change (Kc = -10) Kc = -50 1.8 1.6 1.4 1.2 y(t) 1 1 0.6 0.4 0.2 0 0 5 10 15 20 25 time 30 35 40 45 50 Figure S17.16b. Closed loop response to unit adjustment point change (Kc = -50) 17-22 Kc = -9 2 3.5 3 2.5 2 y(t) 1.5 1 0.5 0 -0.5 -1 1 -1.5 0 5 10 15 20 25 time 30 35 40 45 50 Figure S17.16c. Closed loop response to unit adjustment point change (Kc = -92) Therefore, the closed loop system is fixed for -92 < Kc < 0 As mentioned above, it takes place after a change at the adjustment point. 17.17 a) The closed loop response for set-point changes is GcG(s) Y(s) · Ysp(s) 1 gcG (s) then Gc (z) · 1 (Y/Ysp) G 1 − (Y/Ysp) We want the closed loop system presents a first row plus dead response time, e − hs (Y / Ysp) · -s o1 or (1 − A) z − N −1 (Y / Ysp) · 1 − Az −1 In addition, 17-23 where A = e-t/- G (s) · e −2 s 3s 1 or G(z) · 0.284z −3 1 − 0.716z −1 Thus, the resulting digital controller is Dahlin's Eq. 17-66 controller. Gc (z) · (1 − A) 1 − 0.716 z −1 1 − Az −1 − (1 − A) z − N −1 0.284 (1) If value -=1 is taken into account, then A = 0.368 and Eq. 1 is 0.632 1 − 0.716 z −1 Gc (z) · 1 − 0.368 z −1 − 0.632 z −3 0.284 (2) b) (1-z-1) is a denominator factor in Eq. 2, showing the presence of integrated action. Then, no offset occurs. (c) From Eq. 2, the denominator of Gc(z) contains a non-zero z-0 term. Therefore, the controller is of course feasible. d) First adjust the process time delay to maintain a zero order by adding t/2 to get a time delay of 2 + 0.5 = 2.5 minutes. Then, obtain the CONTINuOS PID controller setting based on the ITAE setting relationship (adjustment point) in table 12.3 with K = 1, · =3, - = 2.5. So KKc = 0.965(2.5/3) − 0.85 , Kc = 1.13 o/ = 0.796 + (-0.1465)(2.) 5 /3), , ,I = 4.45 · D/0.929(2.5/3)0.308 = , · D = 0.78 Using the position format of PID (Eq. 8-26 or 17-55) 1 Gc (z) · 1.13 · 1 o 0.225 · 0.78(1 − z −1) −1 · 1 − z 2.27 − 2.89 z −1 · 0.88 z −2 · 1 − z −1 Using Simulink-MATLAB, examine the performance of the controller: 17-24 1.4 1.2 1 y(t) 0.8 0.6 0.4 0.2 0 0 5 10 15 20 25 Time 30 35 40 45 50 Figure S17.17. Closed loop response to change unit step at specified point. Therefore, the performance shows 21% excess and also wobbles. 17.18 (a) C2's(s) t ysp(z) ysp(z) Km +- E(z) P(z) D(z) M(s) H(s) G v)G d(s) Q2's Gp (s) B(s) B(s) B(s) Gm(s) The transport functions in the various blocks are as follows. Km = 2.5 ma / (mol solute/ft3) Gm(s) = 2.5e-s 17-25 + + C3's(s) C3(s) H(s)= 1 − e −s Gv(s) = Rv = 0.1 ft3/min.ma To obtain gp(s) and G d(s), write the soluble for the rest of the tank as V3 · q1c1 q2 (t) − q3c3 (t) dt Line and use the deviation variables V dc3 · q2 c2 o c2 q2 − q3 c3 dt Receiving Laplace transformation and substitution 30SC3(s) · 1.5Q2 1.5Q2 (s) 0.1C 2 (s) − 3C3 (s) Therefore, b) G p (s) · C3 (s) 1.5 0.5 · Q2 (s) 30s (3 10s) 1 (a) s) 0.1 0.033 · C2 (s) 30s 3 10s 1 G p (z) · C3 (z) 0.05 · Q2 (z) 1 − 0.9 z −1 Analog controller gives a first command exponential response to a unit step change in integral C2. This controller will also provide a first order response to adjustment point changes. Therefore, the desired response could be determined as (Y/Ysp) · 1 1 17.19 17-26 HG p (z) K mGc (z) Y · Ysp 1 hg p Gm (z)Gc (z) Resolve for Gc(z) Gc (z) · Y Y HG p (z) K m − HG pGm (z) Ysp (1) Since the process is not time-delayed, N = 0. Therefore Y · · Ysp (1 − A) z −1 · 1 − Az −1 d Extra HG p (z) · z −1 1 − z −1 HG p Gm (z) · z −2 1 − z −1 Km = 1 Replacement in (1) gives (1 − A) z −1 1 − Az −1 Gc (z) · 1 z z −2 (1 − A) z −1 − 1 − z −1 1 − z −1 1 − z −1 1 − Az −1 Back (gc (z) · (1 − A) − (1 − A) z −1 1 − Az −1 − (1 − A) z −2 Using Simulink-MATLAB; the closed loop response is displayed for different values of A (actually different values of -) : -=3 A = 0,716 17-27 -=1 - = 0,5 A = 0,368 A = 0,135 5 4,5 4 3,5 y(t) 3 2.. 5 2 1.5 1 -- Closed loop response to change the drive step in the disturbance. 17.20 The closed loop response for changing the adjustment point is HG(z) K vGc (z) Y Ysp 1 hg (z) K v K m (z) Gc (z) Therefore Gc (z) Y Ysp 1 HG (z) K K K Y v m Ysp The process transfer function is 2.5 G(s) · 10s 1 or HG (z) · 17-28 0.453z −1 1 − 0.819z −1 (- = 0 thus N = 0) Minimum prototype controller implying - = 0 (e.g., One (0) . Then, Y z 1 Ysp Therefore the controller is 1 0.819z 1 z

1 Gc (z) 0.453z 1 0.2 (0.2)(0.25) z 1 Simplifying, Gc (z) z 1 0.819z 1 0.819z 1 0.091z 1 0.023z 2 0.091 0.023z 1 17.21 a) From Eq. 17-71, the Vogel-Edgar controller is GVE (1 a1 z 1 a 2 z 2)(1 A) (b1 b2)(1 Az 1) (1 A)(b1 b2 z 1) z N where A = e−Δt/τ = e−1/5 = 0.819 Using z-transforms, the discrete-time version of the second-order transfer function yields a1 = -1.625 a2 = 0.659 b1 = 0.0182 b2 = 0.0158 Therefore GVE (1 1.625z 1 0.659z 2)0.181 (0.0182 0.0158)(1 0.819z 1) 0.181(0.0182 0.0158z 1) 0.181 0.294z 1 0.119z 2 0.034 0.031z 1 0.003z − 2 Using Simulink-MATLAB, the controlled variable y(k) and the output of the p(k) controller to change the unit step in ysp are displayed. 17-29 Controlled variable y(k): 1.0 0.8 0.6 0.4 0.2 0 0 5 10 15 20 25 Figure S17.21a. Controlled variable y(k) to change unit step in ysp. Controller output p(k): 5.5 5 4.5 4 p(k) 3.5 3 2.5 2 1.5 1 0.5 0 5 10 15 20 25 Figure S17.21b. Controller output p(k) to change unit step in ysp. 17-30 Dahlin Controller, the output of the controller and the controlled variables are presented. The controller is G DC (z) ... 3.5 3 2.5 p(t) 2 1.5 1 0.5 0 -0.5 -1 0 5 10 15 20 25 30 35 40 45 50 Figure S17.22a. Controller output for Dahlin controller. 1.4 1.2 Output 1 0.8 0.6 0.4 0.2 0 0 5 10 15 20 25 30 35 40 45 50 Figure S17.22b. Closed loop response for Dahlin controller. 17-31 So there is no ringing (this is expected for a first-class system) and no setting is required for ringing. PID (ITAE setpoint) For this controller, adjust the process time delay for the zero-order hold by adding Δt/2, and K=2, τ=10, θ=1.5 obtain the continuous PID controller tunings from Table 12.4 as KKc = 0.965(1.5/10) ... = 4.98; τI = 1.28; τD = ... By using Simulink, 4 3.5 3 2.5 p(t) 2 1.5 1 0.5 0 -0.5 -1 0 5 10 15 20 25 30 35 40 45 50 Figure S17.22c. Controller output for PID (ITAE) controller 17-32 1.4 1.2 Output 1 0.8 0.6 0.4 0.2 0 0 5 10 15 20 25 30 35 40 45 50 Figure S17.22d. Closed loop response for PID controller (ITAE). Dahlin's controller is better since it does not have ringing and the PID controller has more oscillations. The Dahlin controller gives closed loop performance than PID because it includes time delay compensation. 17.23 From Eq. 17-66 to a1 = e−15/10 = 0.819, N=5, and A = e−1/1 = 0.37, the dahlin controller is G DC (z) ... Using Simulink-MATLAB, the output of the controller and the controlled variables are displayed: 17-33 ... Figure S17.23. Controller output for Dahlin controller. ... This is expected for a first-class system. 17.24 Dahlin Controller Using Table 17-1 with K=0.5, τ = 1.0, θ = 0.5 ... Using Simulink, the controller output is displayed: 5 4 p(k) 3 2 1 0 -1 -2 0 5 10 15 20 25 30 Figure S17.24a. Controller output for Dahlin controller. This is the first one, no ringing. Parameters for Dahlin controller: Vogel-Edgar controller ... Figure S17.24b. Controller output for the Vogel-Edgar controller. As noted in the Figure, S17.24 a, ringing happens for Dahlin's controller. Vogel-Edgar controller is GVE (z) = ... Using the Simulink-MATLAB, the output of the controller and the controlled variable are shown below: 17-35 ... Figure S17.24b. Controller output for the Vogel-Edgar controller. 17.25 a) Material Balance for the tanks, dh1 ... q1 ... dt R dh2 ... By using Excel-Solver, ... 17-37 ... Figure S17.25. The controller output for dahlin controller. As noted in the Figure, S17.25, the controller output does not oscillate. c) This controller is naturally liquid since the z-0 factor in the denominator is non-zero. Thus, the controller is of course feasible for all values of τ ... A conservative initial guess for τ would be equal to the open loop time constant, that is, τ = 4.9 minutes. If the accuracy of the model is reliable, a bolder guess will include a smaller τ, say 1/3 rd of open-loop fixed time. In that case, the initial guess would be τ = (1/3) 4.9 =1.5 min. 17.26 ... Substituting ... Conversion of the controller transfer function into a difference equation: p(k) ... By using Simulink-MATLAB, discrete and continuous responses are compared ... Note that b1=0.5 , b2 = −0.333 and a1 = −0.833 ... 1.0 Output 0.8 0.7 0.6 0.5 Continuous Response Discrete Response Discrete Response Discrete Response 0 0.4 0 5 10 15 20 25 Time 30 35 40 45 50 Figure S17.26. Comparison between discrete and continuous controllers. 17-39 Chapter 18 18.1 McAvoy has reported the PI controller settings shown in Table S18.1 and the pig adjustment point responses. S18.1a and S18.1b. When both controllers are designed with 2N settings, unwanted damped oscillations occur due to control loop interactions. The multi-signal adjustment method results in more conservative settings and more sluggish responses. T17 Controller Pair – R T4 – S Tuning Method Single LoopZ-N Single LoopZ-N Kc -2.92 4.31 (min) 3.3 18.1 15 T17 – R T4 – S Multiloop Multiloop -2.59 4.39 2.58 2.58 Table S18.1. Controller settings for example 18.1 1.6 1.4 1.2 T17 1.0 0.8 0.6 Single loop adjustment (one loop in manual) Single loop adjustment (both loops in automatic) Multi-signal adjustment 0.4 0.2 0 0 5 10 15 Time (min) Figure S18.1a. Set the point answers for exercise 18.1. Analysis for the T17 Solution Manual for Process Dynamics and Control, 4th Edition Copyright © 2016 by Dale E. Seborg, Thomas F. Edgar, Duncan A. Mellichamp, and Francis J. Doyle III 18-1 1.6 1.4 1.2 T4 1.0 0.8 0.6 Coordination of a loop (one loop in the manual) Coordination of a loop (both loops in automatic) Multisetup setting 0.4 0.2 0 0 5 10 15 Time (min) Figure S18.1b. Set point responses for exercise 18.1. Analysis for T4 18.2 The characteristic equation is identified by identifying any of the four transport functions Y1(s)/Ysp1(s), Y1(s)/Ysp2(s), Y2(s)/Ysp1(s) and Y2(s)/Ysp2(s), and setting its denominator equal to zero. In order to specify, e.c./Y1/Ysp1(s), Ysp2 = 0 in Eq 18.3b and use the algebra block diagram to obtain C1(s) ... GP12 GC1 (1 Gp11 ...) 1 + Gc2 Gp22 Therefore characteristic equations are (1 + Gc1 Gp12) (1 + Gc2 Gp22) − Gc1 Gc2 Gp12 Gp21 = 0 If either Gp11 or Gp22 is zero, this is reduced to (1 + Gc1 Gp11) = 0 or (1 + Gc2 Gp21) = 0 So the stability of the overall system simply depends on the stability of the two individual feedback control loops in Fig. 18.3b if Gp12 and Gp21 is damaged. 18.3 Consider the bar diagram for control pair 1-1/2-2 in Fig 18.3a but including a sensor and valve transfer function (Gv1,Gv2), (Gm1, Gm2) for each output (y1,y2). The following expressions are easily derived. Y(s) = ... (If Eqs. 1 to 3 are resolved for the output response to variations in specified points, the result is 18-3 Y (s) = [I + Gp(s)Gc(s)Gm(s)]−1 Gp(s)Gc(s)Gm(s) Ysp(s) where I is the identity matrix. In terms of the component transfer function matrix h11, h12, h21, h22 ... (4) where h11(s) = Gc1(s) Gv1(s) Gm1(s) h12(s) = Gp12(s) Gc2(s) Gv2(s) Gm2(s) h21(s) = Gp21(s) Gc1(s) Gv1(s) Gm1(s) h22(s) = Gp22(s) Gc2(s) Gv2(s) Gm2(s) The inverse of V, if exists, is ... The inverse is given by ... and the closed-loop transfer functions are (see book notation) T11(s) ... T12(s) ... T21(s) ... T22(s) ... 18-4 From Eqs. 6-91 and 6-92 and by natural reasoning, it is obvious that although h is affected by both manipulated variables, T is affected only by wh and is 18-4 regardless of w. Therefore, T can only be combined with wh. So the coupling based on this for the control system is T-wh, h-w. 18.5 System transfer mode matrix ... The characteristic equation of the system is ... For (a) Kc1 = −1; Kc2 = 0 ... coupling is unstable Step Response of Y 1 4 1-1-1/2-2 coupling 3.5 3 2.5 2 1.5 0.5 0 1.0 10 20 30 40 50 Time 60 70 80 Figure S18.5a. Response step of Y1 18-5 90 100 Step Response of Y 2 6 1-1/2-2 mating 5 4 3 2 1 0 10 20 30 40 50 Time 60 70 80 90 100 Figure S18.5b. Response step of Y2 18-6 90 100 Step Response of Y 1 0.7 1-1/2-2 match 0.6 0.5 0.4 0.3 0.2 0.1 0 0 10 20 30 40 50 60 70 80 90 100 Time Figure S18.5c. Step Response of Y 1 18-7 Step Response of Y 2 1.2 1.0 0.8 0.6 1 0.4 0.2 0 0 10 20 30 40 50 60 70 80 90 100 Time Figure S18.5d. Y2 Step Response (c) Kc1 = 1; Kc2 = 2: coupling is unstable Step Response of Y 1 46 6 x 10 1-1/2-2 coupling 5 4 3 2 1 0 -1 -2 0 50 100 150 Time 200 250 Figure S18.5e. Step response Y1 18-8 Step Response of Y2 18-7 Step Response of Y 2 46 1.5 x 10 1-1/2-2 coupling 1 0.5 0 -0.5 -1 -1.5 -2 -2.5 -3 -3.5 0 50 100 150 Time 200 250 Figure S18.5f. Step response of Y2 18.6 a) Calculate the steady-state gains ... 18.7 (a) GC1 = K1 ... The gain of the coupling ... For positive relative gains requires XD-R, XB-S. (ii) This coupling also appears appropriate from dynamic estimates; due to the delay in the column, R affects XD earlier than XB, and S affects XB earlier that XD. (a) The corresponding fixed-state profit table is 18.7 ... Pairing for positive relative gains requires XD-R and XB-S. b) The same coupling is recommended on the basis of dynamic issues. Transfer modes between XD and R contain less dead time and shorter time constant, so XD will respond very quickly to changes in R. For the XB-S pair, the time constant is not important due to the dead time is significantly shorter and the response will be quick as well. 18-9 18.8 a) From Eq. 6-105 G p11 (s) ... Therefore ... replacement in Eq. 18-34, ... the ideal disconnection control system will be like in Fig.18.9 where Y1, Y2 Ultimate value of ... controller ... 18.9 ... 18.10 ... OPTION B: Controlled variable: T17, T30 Manipulated variables: u1, u2 The corresponding fixed-state gain matrix is ... Using the formula in Eq. 18-34, we see the RGA is ... OPTION C: Controlled variable: T24, T30 Manipulated variables: u1, u2 The corresponding constant state gain matrix is ... Using the formula in Eq.18-34, we receive −1 = 290 So the RGA is ... Option C is the best coupling, because the values are highest for the first controller. Otherwise, the relationship between elements would be difficult for configuring multi loop controllers, such as 1 − 2, 2 − 1. ... The material balance in the tanks is ... By taking the Laplace transform, A1sH1(s) ... The four fixed-state process gains are determined using Eqs. 3 and 4 ... Thus RGA is ... Replacement of the specific numeric values; the RGA is ... The RGA is q1-h1 the best match for coupling loops q1-h2 and q2-h1, q2-h2. 18.11 (a) The ... Then, with inspection of the Eqs. (3) and (4) in the solution of exercise 18-10, GP(s) ... Therefore ... 18-13 For the relevant gains to be positive, the preferred coupling is h1-q1, h2-q2. 18.13 ... Thus (A1s K1 ...) ... K1 and K2 are defined in the solution of exercise 18.10. The block diagram for coupling h1-q1 / h2-q2 is identical to Fig.18.3a with the addition of load. Thus the mark D passes through a fan Gd1 which is added to the summer with the production of Y1. Similarly, the summer leading to Y2 is affected by the signal D(s) passing through the block Gd2. 18.12 F = 20 u1 (P1 − P) ... P = 30 u2 (P1 − P2) (1) (2) Taking P0 and P2 to be stable, Eq. 1 gives 18-14 ... F ... 20 (P0 − P1) ... 30M 2 1 J ... Ou1 ... u2 ... 20 u2 3 0 u2 (3) Using Eq. 18-24, ... P0 − P1) , P2 − P2) = 10 P ... 2/3 30(P1 − P2) Replacing in (7), F ... Therefore, the best controller pairing is F-u1, P-u2. 18.13 a) Basic balances for the tank, ... Replacing ... Similarly, ... By replacing numerical values ... Thus, ... Similarly ... Substituting numerical values ... Therefore ... 18.14 In this case, RGA analysis is not required. Manipulated and controlled variables are: Controlled variables: F1, P1 and I Manipulated variables: m1, m2, m3 Basically, coupling could be done based on dynamic estimates, so that time constants and dead times in response should be as low as possible. The level of interface I can be easily controlled with m3 so that any change in the adjustment point is controlled by opening or closing the valve at the bottom of the carafe. The manipulated variable m1 could be used to control the F1 input rate. If F1 is removed from the adjustment point, the valve will respond quickly to control this change. 18-17 The pressure of the P1 carafe is controlled by m2. In this way, pressure changes will be dealt with quickly. This control configuration is also used in distillation columns. 18.15 OPTION A: Controlled variable: Y1, Y2 Manipulated variables: U1, U2 The corresponding fixed-state profit table is 3 −0.5 K ... −10 Using in Eq.18-34, we receive -11 = 6 So the RGA is ... OPTION B: Controlled variable: Y1, Y2 Manipulated variables: U1, U3 The corresponding fixed-state gain table is ... 2 Using the formula in Eq.18-34, we have -11 = 0.67 So the RGA is ... 18-19 0.33 0.67 from Bristol accounting initial recommendation , controlled and manipulated variables are coupled so that the corresponding relative gains are positive and as close as possible to one. Thus, OPTION B leads to better control configuration. 18.16 The process shape appears below q2 q1 MIX T2 = 140 F T1 = 70 F q3 T3 = 110 F Figure S18.16. Process system (a) Fixed state material balance: q1 + q2 = q3 (1) 18-19 Constant state energy balance: q1C(T1-Tref)+q2C(T2-Tref) = q3C(T3-Tref) (2) By replacing (1) at (2) and resolving: q1 = 9/7 qm q2 = 12/1 27 gpm (b) The fixed-state profit table K: ... The RGA analysis: ... 18-20 ... 18.17 a) Dynamic model: Mass balance: ... Energy balance: ... Control values: h1 U ... Variables: 14 inputs, w1, w2, w3, w4, T1, T2, T3, T4, Tc, xc, x3, ... 18-21 ... Υποκατάστατο ... Θα είναι δύσκολο να ελεγχθεί ο T4, διότι ούτε x3 ούτε f έχει μεγάλη σταθερή επίδραση στον T4. 18.18 α) Ισοζύγιο μάζας: F ... F1 ... F2 Fw ... F2 w2 ... 0,4 F2 CV: w, F, MV: F1 και F2. Γραμμοποιείται η διαδικασία στο σημείο λειτουργίας, όπως περιγράφεται στο σημείο 18.2.2. ... F2 ... F1 ... F2 ... 0,4 F2 K 1 21 ... 0,025 ... F2 ... F1 ... F2 ... 0,4 F ... 0,4 F2 K 22 ... 0,025 ... F2 ... F1 (β) RGA: -11 ... F1 F1 ... F2 1 1 ... 2 K ... 0,4 F2 / F F1 ... F2 F 1 ... 12 21 1 ... 2 K11 K 22 ... 0.4F ... 0.4 F2 ... / F Έτσι, η συστοιχία RGA είναι 18-25 ... F F2 ... F ... F2 F F F2 ... F ... F − F2 ... F ... c) -11 • 1 1 • • 0.5 K12 K 21 1 o 0.025 / 0 .025 1− K11 K 22 0.5 0.5 Thus, the RGA array becomes: ... , in this case, either the coupling is 0.5 0.5 recommended. 18.19 (a) (i) Static estimates: Coupling according to RGA data closest to +1: H1 – Q3, pH1-Q1, H2-Q4, pH2 – Q6 ii) Dynamic issues: Coupling results in the shortest time constants for tank 1. It is also dynamically better for tank 2 because it avoids the large ratio of 0.8. (iii) Natural estimates The proposed coupling makes sense because the controlled variables for each tank are combined with the incision flows for that one tank. Because the pH is more important than the level, we could use the coupling, H1 – Q1 / pH1-Q3 , for the first tank to provide better pH control due to the shorter time delay (0.5 vs. 1.0 minutes). b) The new profit matrix for problem 2 to 2 is 18-26 0.42 0.41 K −0.32 0.32 From Eq. 18-34, -11 • 1 • 0.506 (0.41)(--0.32) 1• (0.42)(0.32) 0.1 506 0.494 L- 0.494 0.506 RGA coupling: H2 – Q4 / pH2-Q6. The pairing also avoids the large delay of 0.8 min. 18.20 Since level is tightly controlled, there is a no accumulation, and a material balance yields: Overall: wF – E wS – wP 0 Solute: wFxF - wPxP 0 (1) (2) Controlled variable: x ΔP , wΔF Manipulated variables: wΔP , wΔs From (1): wF = wS E + wP From (2): xP wF x wF F (ws E wP) wP wP Using deviation variables: 18-27 (3) wΔF wΔs E wΔP Linearizing (3): xP wP ... x F Ews x ΔP ... 2 wP ... xP ... ws (wΔs) wP , ws ... x F ... wΔP ... ws ... P ... (5) Then the steady-state gain matrix is w ΔP x ΔP ... x F Ew s ... w 2 P ... w ΔF wΔs ... xF E ... • • ... wp 1 E Using the formula in Eq.18-34, we will acquire Ews 1 • - 22 wp Ews o wp 1 o Ews wp -12 • - 21 • 1 − -11 • Ews • o o wp -11 = So, the RGA is Ews • Ew o w p • s • L • o wP • Ews o wp Ews o wp Ews Ews o wp so, if Ews wP , the coupling should be x ΔP - w ΔP / w ΔF - wΔs So if the Ews o wp , the coupling should be is x ΔP - wΔs / w ΔF - w ΔP 18-28 18.21 a) The corresponding fixed-state gain the table is −0 ,04 −0.0005 K • −0.02 0.22 Using the formula in Eq. 18-34 , we have received So the RGA is -11 = 1.16 1.16 −0.16 L • −0.16 1.16 b) Matching for positive relative gains requires y1-u1 and y2-u2. 18.22 For the higher dimension process (n>2) the RGA can be calculated from the expression -ij = Kij Hij where Hij is the element (i,j) of H = (K-1)T Using MATLAB, 62.23 −122.17 58.02 K • • −84.47 170.83 −83.43 • 1.95 −14.85 13.09 -1 62.2 3 −84.47 1.95 H • • • • 122,17 170,83 −14,85 • 58,02 −83,4 3 13.09 Thus, the RGA is 18-29 1.89 210.34 −211.18 • L • • 390.95 406.. 58 −14.642 • 181.60 −194.39 13.80 This RGA analysis shows the control difficulties for this process due to the interaction of the control loop. For example, if the mappings are 1-3, 2-2, 3-1, the third loop will experience difficulties in closed loop operation. But other options are no better. SVA Analysis: Determinant of K = K = 0.0034 The status number = CN = 1845 Since the determinant is small, the required adjustments to you will be very large, resulting in excessive control actions. In addition, this example shows that Table K is poorly adjusted and very sensitive to small fluctuations in its elements. 18.23 Application of SVA analysis: Determinant of K = K = -6.76 Status number = CN = 542.93 Large status number indicates poor preparation. Therefore, this process will require large changes to the handled variables in order to affect the controlled variables. Some outputs or inputs should be eliminated to achieve better control and singular value decomposition (SVD) can be used to select the variables to eliminate. Using the MATLAB SVD command, the unique values in Table K are: 21.3682 6.9480 1.1576 0.0394 18-30 Note that 3/4 > 10, then the last unique price can be neglected. If we eliminate one input and one output variable, there are sixteen possible pairings that appear in table S18.23, along with the CN term number. Pairing number 1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 Controlled variables y1,y2,y3 y1,y2,y3 y1,y2,y3 y1,y2,y3 y1,y2,y4 y1,y2,y4 y1,y2,y4 y1,y2,y4 y1,y3,y4 y1,y3,y4 y1,y3,y4 y1,y3,y4 y2,y3,y4 y2,y3,y4 y2,y3,y4 y2,y3,y4 Manipulated variables u1,u2,u3 u1,u2,u4 u1,u3,u4 u2,u3,u4 u1,u2,u3 u1,u2,u4 u1,u3,u4 u2,u3,u4 u1,u2,u3 u1,u2,u4 u1,u3,u4 u2,u3,u4 u1,u2,u3 u1,u2,u4 u1,u3,u4 u2,u3,u4 CN 114.29 51.31 398.79 315.29 42.46 30.27 393.20 317.15 21.21 16.14 3897.2 693.25 24.28 20.62 1332.7 868.34 Table S18.23. CN for different 3x3 mappings. Based on the minimum status number, it is recommended to pair 10 (y1-u1,y3-u2,y4-u4). The RGA for the reduced variable set is 1.654 −0.880 0.226 L • • • 0.785 3.742 −1.957 • 0 . 1312 −1.8615 2.7304 18-31 18.24 1-2/2-1 the controller coupling has a greater stability area compared to 1-1/2-2. RGA: -11 • 1 1 • • 2.28 K12 K 21 1.5 • 1.5 1 − 1 − 2 • 2 K11 K 22 2.28 − 1.28 • • 1.28 2.28 the coupling of the controller must be 1-1/2-2 to avoid negative values. The stability analysis is based on results and uses the numeric range of the controller to receive a stable closed loop response. RGA is based on static gain process (Kij) analysis, which shows only the loop opening steady state behavior. For this problem, 1-2/2-1 matching has a larger stability area, which means choice of Kc1 and Kc2 has a larger margin with guaranteed stability. However, around the stable state, the negative RGA indicates a combat control loop, which may be vulnerable to process noise. Thus, 1-2/2-1 matching should be avoided in this case. 18-32 Chapter 19 19.1 From the definition of xc, 0 • xc • 1 f(x) = 5.3 x e (-3.6x +2.7) Leave three starting points at [0.1] to be 0.25, 0.5 and 0.75. Calculate x4 using Eq. 198,. x1 0.25 f1 8.02 x2 0.5 f2 6.52 x3 0.75 f3 3.98 x4 0.0167 For the next iteration, select x4, and x1 and x2 from F1 and F2 is the largest between f1, F2, F3. Thus, successive iterations are x1 0.25 0.25 0.25 f1 8.02 8.02 8.02 8.02 8.02 x2 0.0.. 5 0.5 0.334 0.271 xopt = 0.2799 F2 6.52 6.52 7.92 8.06 x3 0.017 0.334 0.271 0.280 F 3 1.24 7.92 8.06 8.06 x4 0.334 0.271 0.280 does not need 7 function ratings 19.2 As shown in the plan, there is both a minimum and a maximum value of the air/fuel ratio so that the thermal efficiency is non-zero. If the ratio is too low, there will not be sufficient air to maintain combustion. On the other hand, problems in combustion will occur when too much air is used. Maximum thermal efficiency is achieved when the air/fuel ratio is stoic. If the amount of air is excessive, relatively more heat will be absorbed by the air (mainly nitrogen). However, if the air is not sufficient to maintain total combustion, thermal efficiency will also decrease. Process Dynamics and Control Solution Manual, 4th Edition Copyright © 2016 by Dale E. Seborg, Thomas F. Edgar, Duncan A. Mellichamp and Francis J. Doyle III 19-1 19.3 Using Excel-Solver, this optimization problem is quickly resolved. The selected starting point is (1.1): Initial values Final values X1 X2 1 1 0.776344 0.669679 max Y= 0.55419 Restrictions 0 • X1 • 2 0 • X2 • 2 Table S19.3. Excel solution Therefore, the optimal point is (X1*, X2*) =(0.776, 0.700) and the maximum value of Y is Ymax = 0.554 19.4 Let N be the number of lots / year. Then np • 300,000 Since the goal is to minimize the cost of annual production, only the required amount should be produced annually and not more. That is, NP = 300,000 a) (1) Minimize the total annual cost, $ $ lot 0.4 hour min TC = 400,000 +2P 50 N lot • lot • hours • years $ + 8 000 P0.7 • • • years The replacement of N by (1) minimum TC = 400,000 + 3x107 P–0.6 + 800 P0.7 19-2 b) There are three limitations in P i) P • 0 N is integer. That is, (300,000/P) = 0, 1, 2,... (iii) The total production time is 320 x 24 hours/year lot hr (2 P0.4 + 14) 14) • N • • lot • 7680 years old • Replacing the N by (1) and simplifying 6 • 105P-0.6 + 4.2 • 106P-1 • 7680 c) d (TC) • 0 • 107 (--0.6) P −1.6 o 800(0.7) P −0.3 dP 1/1.3 P opt 3 .107 (--0.6) −800(0.7) lot 2931 lb d 2 (TC) • 3 • 107 (--0.6)--(1.6) P −2.6 μ 800(0.7)(–0.) 3) P −1.3 2 dP 2 d (TC) • 2.26 • 10--2 0 therefore minimum 2 dP P • Popt Nopt = 300,000/Popt = 102.35 is not integer. Ως εκ τούτου, ελέγξτε για Nopt = 102 και Nopt = 103 Για Nopt = 102, Popt = 2941.2, και TC = 863207 Για Nopt = 103, Popt = 2912.6, και TC = 863209 Ως εκ τούτου βέλτιστη είναι 102 παρτίδες των 2941,2 lb/ παρτίδας Ο χρονικός περιορισμός είναι 6 105 P--0,6 4,2 106 P--1 6405,8 7680 , ικανοποιημένοι 19-3 19,5 Ας x1 είναι ο ημερήσιος ρυθμός τροφοδοσίας του αργού No.1 σε bbl/ημέρα x2 είναι το ημερήσιο ποσοστό τροφοδοσίας του αργού No.2 σε bbl /day Στόχος είναι να μεγιστοποιηθεί το κέρδος max P = 3,00 x1 + 2,0 x2 Υπόκεινται σε περιορισμούς βενζίνης: 0,70 x1 + 0,41 x2 6000 κηροζίνη: 0,06 x1 + 0,09 x2 2400 μαζούτ: 0,24 x1 + 0,50 x2 12.000 Με τη χρήση του Excel-Solver , Initial values Final values max P = x1 1 0 x2 1 14634.15 29268.3 Restrictions 0.70 x1 + 0.31 x2 y 0.06 x1 + 0.09 x2 0.24 x1 + 0.60 x2 6000 1317 7317 Table S19.5. Excel Solution Therefore, the optimal point is (0, 14634.15) Slow No.1 = 0 bbl/day Slow No.2 = 14634.15 bbl/day 19-4 19.6 The objective function is to maximize revenue, max R = -40x1 +50x3 +70x4 +40x5 –2x1-2x2 (1) *Balance in column 2 x2 = x4 + x5 (2) * From column 1, 1.0 x1 = x2 • 1.667(x4 x5) 0.60 0.4 x 3 = x2 • 0.667(x4) x5) 0.60 (3) (4) Inequality restrictions are x4 • 200 x4 • 400 x1 • 2000 x4 • 0 x5 • 0 (5) (6) (7) (8) The limited operating range for column 2 imposes additional inequality restrictions. The average solvent is 50 to 70% of the bottom. i.e. 0,5 • x4 • 0,7 or x2 0,5 • x4 • 0,7 x4 • x5 Rewrite in linear form, 0.5 x2 ≤ x4 ≤ 0.7 or 0.5 (x4 + x5) ≤ x4 ≤ 0.7 (x4 + x5) Simplification, x4 –x5 • 0 0.3 x4 –0.7x5 • 0 (9) (10) No additional restriction is required for heavy solvent. The fact that the heavy solvent will be 30 to 50% of the bottom is ensured by limiting the average solvent and the total balance in column 2. 19-5 Using Excel-Solver, Initial Values Final values max R = Limitations x2 - x4 - x5 x1 - 1.667x2 x3 - 0.667x2 x4 x1 - 1.667x2 x4 - x5 0.3x4 - 0.7x5 x1 1 1333.6 x 2 1 800 x3 1 533.6 x4 1 400 x5 1 400 13068.8 0 7.467E-10 -1.402E-10 400 400 1333.6 0 -160 Table S19.6. Excel solution So the optimal point is x1 =1333.6, x2 =800; x3=533.6, x4 = 400 and x5 = 400. Replacing (5), the maximum revenue is $13,068/day and the percentage of output flows in column 2 is 50 % for each stream. 19.7 The aim is to minimise the sum of squares of errors for the material balance, i.e., min E = + 11.3 – 92.1)2 + (wA (wA (wA) –94.2)2 + (wA + 11.6 –93.6)2 Subject to wA • 0 Resolve in detail, dE • 0 • 2 (wA + 11.3 – 92.1) + 2 (wA +10.9 –94.2) dwA +2(wA + 11.6 –93.6) Resolve for wA... wA opt = 82.0 Kg/hr 19-6 Check for minimum, d 2E • 2 o 2 o 6 0 , therefore minimum 2 dwA 19.8 The reactor equations are: dx1 • k1 x1 (1) dt dx2 • k1 x1 − k2 x2 (2) dt When k1 • 1.335*1010 e—75000/ o8.31*T • ; k2 • 1.149*1017 e–125000/ o8.31*T • Using MATLAB, this differential equation system can be solved using the command ode45. In addition, we need to apply the fminsearch ino command in order to optimize the temperature. In this way, the results are: Top • 360.92 K ? x2,max • 0,343 MATLAB code: %% Exercise 19,8 function y = Exercise_19_8(T) k10 = 1,335*10^10; % min^(-1) k20 = 1.149*10^17; % min^(-1) E1 = 75000· % J/(g.mol) E2 = 125000· % J/(g.mol) R = 8,31· % J/(g.mol.K) x10 = 0,7· % mol/L x20 = 0· % mol/L k1 =k10*exp(-E1/(R*T))· k2 = k20*exp(-E2/(R*T))· time = [0,6]; % of the time period; initial_val = [x10, x20]; options = odeon('RelTol',1e-4,'AbsTol',[1e-4 1e-4]); [~,X] = odeon45(@reactor, time, initial_val, options). y = -X(end,2); % Due to the investigation, the opposite function should be dx = reactor(t,x) dx = zeros(2,1); % Column carrier Dx(1) = -k1*x(1); dx(2) = k1*x(1)-k2*x(2)· end %% Exercise 22.10 main clear all;clc; close it all; T_range = [200, 500]· T = fminsearch(@Exercise_19_8, 200)· x2_max =-Exercise_19_8(T)· 19-7 19.9 By using Excel-Solver: Initial values Final values 1 2.907801325 0 1.992609 Time 0 1 2 3 4 5 6 7 8 9 10 11 Equation 0 0.065457105 0.200864506 0.350748358 0.489635202 0.607853765 0.703626108 0.778766524 0.836422873 0.879953971 0.912423493 0.936416639 Data 0 0.0583 0.2167 0.36 0.488 0.6 0.692 0.772 0.833 0.888 0.925 0.942 SUM= Square Error 0 5.12241E-05 0.000250763 8.55929E-05 2.67388E-06 6.16816E-05 0.000135166 4.57858E-05 1.17161E-05 6.47386E-05 0.000158169 3.11739E-05 0.000898685 Hence the optimal values are 1 2.9; 2 1.99 .. 19.10 Let x1's gallons of suds mixed x2 are gallons of premium mixed x3 are gallons of water mixed Aim is to minimize the cost min C = 0.3x1 + 0.40x2 (1) Subject to x1 + x2 + x3 = 10,000 (2) 0.03 x1 + 0 .060 x2 = 0.050 • 10,000 19-8 (3) x1 • 2000 (4) x1 • 9000 (5) x2 • 0 (6) x3 • 0 (7) The problem given by Eqs. 1, 2, 3, 4, 5, 6, and 7 is optimized using excel-Solver, initial values Final values Objective mode x1 1 20 00 3533.33333 x2 0 7333.333 x3 0 666.666667 Restrictions x1+x2+x3 0.03x1+0.06x2 x1 x1 x2 x3 10000 50 0 00 2000 2000 7333.333333 666.666667 = = > = =10000 500 2000 9000 0 0 We achieve: suds = 2000 gallons; premium = 7333.3 gallons; water = 666.7 gallons, with a minimum cost of $3533.3. 19.11 Let xA be bbl / of A produced xB to bbl / day of B produced Objective is to maximize the gain max P = 10xA + 14xB (1) Subject to raw material 120xA+ 100xB • 9,000 (2) Warehouse space restriction: 0.5 xA + 0.5 xB • 40 (3) Production time limit: (1/20)xA + (1/10)xB • 7 (4) 19-19-19-19-19 9 Initial values Final values max P = xA 1 20 xB 1 60 1040 Restrictions 120xA+ 100xB 0.5 xA + 0.5 xB (1/20)xA + (1/10)xB 8400 40 7 Table S19.11. Excel solution So the optimal point is xA = 20 and xB = 60 The maximum profit = $1040/day 19.12 PID controller parameters are usually obtained using either process model, process data or computer simulation. These parameters are kept constant in many cases, but when operating conditions vary, supervisory control could include optimizing these configuration parameters. For example, using process data, Kc , I and D can be calculated automatically to maximize profits. A comprehensive analysis of the procedure is required in order to achieve this type of optimal control. Supervisory and regulatory control is complementary. Of course, supervisory control can be used to adjust the parameters of either an analog or digital controller, but feedback control is required to keep the controlled variable at or near the specified point. 19.13 Assuming steady state behavior, the optimization problem is, max f = D e Subject to 0.063 c –D e = 0 0.9 s e – 0.9 s c – 0.7 c – D c = 0 19-10 (1) (2) -0.9 s e + 0.9 s c + 10D – D s = 0 D, e, s, c 0 (3) where f = f(D, e, c, s) Excel-Solver is used to solve this problem, c D e s 1 1 1 1 0.479031 0.045063 0.669707 2.079784 Initial values Final values max f = 0.030179 Constraints 0.063 c –D e 0.9 s e – 0.9 s c – 0.7 c – Dc -0.9 s e + 0.9 s c + 10D – Ds 2.08E-09 -3.1E-07 2.88E-07 Table S19.13. Excel solution Thus, the optimal value of D is equal to 0.045 h-1 19.14 Material balance: Total : FA + FB = F Component B: FB CB + VK1CA – VK2CB = F CB Component A: FA CAF + VK2CB + VK1CA = FCA So the optimization problem is: max (150 + FB) CB Subject: 0.3 FB + 400CA − 300CB = (150 + FB)CB 45 + 300 CB – 400 CA = (150 + FB) CA FB • 200 CA, CB, FB • 0 19-11 Using Excel-Solver, the best values are FB = 200 l/hr CA = 0.129 mol A/l CB = 0.171 mol B/l 19.15 Material balance: Total : FA + FB = F Component B: FB CB + VK1CA – VK2CB = F CB Component A: FA CAF + VK2CB – VK1CA = FCA So the optimization problem is : max (150 + FB) CB Subject to : 0.3 FB + 3 o 106e(-5000/T)CA V − 6 • 106e(-5500/T)CB V = (150 + FB)CB 45 + 6 • 106 (-5500/T)CB V – 3 • 106e(-5000/T) CA V = (150 + FB) CA FB • 200 300 • T • 500 CA, CB, FB • 0 Using Excel-Solver, the best values are FB = 200 l/hr CA = 0.104 mola/l CB = 0.177 mol B/l T = 311.3 K 19-y 12 Chapter 20 20.1 (a) The unit step response is ... Ως εκ τούτου , y (t) = 3S (t − 2) 2) + 2e − (t − 2)/10 − 3e − (t − 2)/15 Si = y (iDt) = y (i) = b) {0, 0, 0.0095, 0.036, 0.076, 0.13...} Evaluate the expression for y(t) in part a) y(t) = 0.99 (3) ≈ 2.97 in t = 87. So, N = 87, for 99% full answer. 20.2 (a) Note that G (s) = Gv (s)G p (s) Gm (s) . From Figure 12.2, ym(s) 4(1 − 3s) (s) = G= P(s) (15s + 1)(5s + 1) (1) for unit step change, P (s) = 1 / s , and (1) becomes: Ym (s) = 1 4(1 − 3s) s (15s + 1)(5s + 1) © Solution Manual for Process Dynamics and Control, 4th Edition Copyright © 2016 by Dale E. Seborg, Thomas F. Edgar, Duncan A. Mellichamp, and Francis J. Doyle III 20 - 1 Partial Fraction Extension: A B C 1 4(1 − 3s) Ym (s) = + + + s (15s + 1) (5s + 1) s (15s + 1) (5s + 1) (5 1) (2) where: A = A B= 4(1 − 3s) = 4 (15s + 1)(5s + 1) s =0 4(1 − 3s) = −108 s (5s + 1) s = − − 1 15 = C 4(1 − 3s) = 16 s (15s + 1) s = − 1 5 Substitute in (2) and take the reverse Laplace Transformation : 36 16 ym (t) = 4 − e − t /15 + e − t /5 5 5 (3) b) The new constant state value is obtained from (3) to be ym(∞)=4. For t = t99, ym(t)=0.99ym(∞) = 3.96. Substitute in (3) 36 16 3.96 = 4 − e − t99 /15 + e − t99 /5 5 5 (4) Resolve (4) for t99 gives t99 ≈ 77.9 minutes So we determine that Dt = 77.9/30 ≈ 3 minutes Table S20.22. Συντελεστές απόκρισης βημάτων k 2 3 4 5 6 7 8 9 10 t (min) 3 6 9 12 15 18 21 24 27 30 Si -0,139 0,138 0,578 1.055 1.511 1.919 2.27 2 2.573 2.824 3.034 k 11 12 13 14 15 16 17 18 19 20 t (min) 33 36 39 42 45 48 51 54 57 60 20 - 2 Si 3.207 3.3.3 49 3.467 3.563 3.642 3.707 3.760 3.803 3.839 3.868 k 21 22 23 24 25 26 27 28 29 30 t (min) 63 66 69 77 5 78 81 84 87 90 Si 3.892 3.912 3.928 3.941 3.951 3.960 3.967 3.973 3.978 3.982 20.3 Από τον ορισμό της μήτρας S, που δίνεται σε Eq. 20-28 , for P=5 , M=1, with Si taken from exercise 20.1, • S1 o 0 o S o 0.01811 2 s = S 3 o = 0.06572 o S 4 o 0.1344 from 0.2174 from 0.2174 from 0.2174 from 0.2174 from 0.2174 from 0.2174 from 0.2174 from 0.2174 from 0.2174 From 0,2174 from 0, From 0.217420-65: Kc = (STS)-1ST Kc = [0 0.2589 0.9395 1.9206 3.1076] = Kc1T Because Kc1T is defined as the first row of Kc , Using the given analytical result, 1 Kc1T = 5 [S1 S 2 S 3 S 4 S 5] 2 • (S i) i =1 Kc1T = 1 [0 0.01811 0.06572 0.1344 0.2 174] 0.06995 Kc1T = [0 0.2589 0.9395 1.9206 3.1076] which is the same as the answer received above using (20-65). 20.4 The step response is taken from the step response of the analytical unit as in example 20.1. The Kc feedback table is obtained using Eq. 20-65 as in example 20.5. These results are not mentioned here for the sake of brevity. The closed loop response for point changes and disturbance is shown below for each case. The MATLAB MPC toolbox was used for the simulations. 20 - 3 i) For this model horizon, the step response is over 99% complete as in example 20.5. therefore, the model is good. Teh Teh and the disturbance responses shown below are non-oscillating and have long break-up times Outputs 1.5 1 y 0.5 0 0 10 20 30 40 50 60 70 80 90 100 70 80 90 100 Time Manipulated Variables 2 1.5 u 1 0.5 0 0 10 20 30 40 50 60 Time s20.4a. Auditor (i); changing the adjustment point. Outputs 0.8 0.6 y 0.4 0.2 0 0 10 20 30 40 50 60 70 80 90 100 70 2 80 90 100 Time Manipulated Variables 0 -0.5 u -1 -1.5 0 10 20 30 40 50 60 Chronograph S20.4b. Auditor (i); change of disturbance. 20 - 4 ii) The answer shown below shows the same excess, shorter precipitation time and unwanted ringing in u compared to part i). The disturbance response indicates a lower peak value, a lack of oscillations, and a faster adjustment of the manipulated input. Outputs 1.5 1 y 0.5 0 0 10 20 30 40 50 60 70 80 90 100 70 80 80 9 0 100 Time Manipulated Variables 15 10 5 u 0 -5 -10 0 10 20 30 40 50 60 Time Shape S20.4c. Auditor (ii); changing the adjustment point. Outputs 0.4 0.3 y 0.2 0.1 0 0 10 20 30 40 50 60 70 80 90 100 70 28 0 90 100 Time Manipulated Variables 0 -0.5 u -1 -1.5 0 10 20 30 40 50 60 Time s20.4d. Auditor (ii); change of disturbance. 20 - 5 (iii) The responses to the adjustment point and the disturbances presented below show the same trends as Part i). Outputs 1.5 1 y 0.5 0 0 10 20 30 40 50 60 70 80 90 100 70 80 9 0 100 Time Manipulated Variables 10 5 u 0 -5 -10 0 10 20 30 40 50 60 Year Figure S20.4e. Auditor (iii); changing the adjustment point. Outputs 0.4 0.3 y 0.2 0.1 0 0 10 20 30 40 50 60 70 80 90 100 70 28 0 90 100 Time-manipulated Variables 0 -0.5 u -1 -1.5 0 10 20 30 40 50 60 Time s20.4f. Auditor (iii); change of disturbance. 20 - 6 (iv) The responses of specified points and loads presented below show the same trends as in Parts (i) and (ii). Compared to part (iii), this controller has a longer penalty for manipulated input and therefore leads to a smaller and less oscillating entry effort at the expense of greater excess and settlement time for the controlled variable. Outputs 1.5 1 y 0.5 0 0 10 20 30 40 50 60 70 80 90 100 70 80 90 100 Time manipulated variables 4 3 2 u 1 0 -1 0 10 20 30 40 50 60 Year Old Figure S20.4g. Auditor (iv); changing the adjustment point. Outputs 0.5 0.4 0.3 y 0.2 0.1 0 0 10 20 30 40 50 60 70 80 90 100 100 100 270 80 90 100 Time-manipulated Variables 0 -0.5 u -1 -1.5 0 10 20 30 40 50 60 Year Figure S20.4h. Auditor (iv); change of disturbance. 20 - 7 20.5 There are many sets of M, P, and R values that meet the given limitation for a unit load change. One such set is M=3, P=10, R=0.01 as shown in exercise 20.4(iii). Another set is M=3, P=10, R=0.1 as shown in exercise 20.4(iv). A third set of values is M=1, P=5, R=0 as shown in the exercise 20.6 (Using the MATLAB model prediction toolbox) As shown below, controller (a) provides a better disturbance response with less maximum output deviation and less control effort. However, the controller (a) is poorer for a change of specified point, point, ringing at the manipulated entrance. Outputs 1.5 1 y 0.5 0 0 10 20 30 40 50 60 70 80 90 100 70 80 80 9 0 100 Time Manipulated Variables 15 10 5 u 0 -5 -10 0 10 20 30 40 50 60 Time s20.6a. Auditor (a); changing the adjustment point. 20 - 8 Exits 1.5 1 y 0.5 0 0 10 20 30 40 50 60 70 80 90 100 70 8 8 0 90 100 Time Manipulated Variables 15 10 5 u 0 -5 -10 0 10 20 30 40 50 60 Chronograph S20.6b. Auditor (a); change of disturbance. Outputs 0.4 0.3 y 0.2 0.1 0 0 10 20 30 40 50 60 70 80 90 100 70 2 80 90 100 Time Manipulated Variables 0 -0.5 u -1 -1.5 0 10 20 30 40 50 60 Time Shape S20.6c. Auditor (b); changing the adjustment point. 20 - 9 Exits 0.4 0.3 y 0.2 0.1 0 0 10 20 40 30 60 50 70 80 90 10070 80 90 100 Time Manipulated Variables 0 -0.5 u -1 -1.5 0 10 20 30 40 30 60 50 Year Figure S20.6d. Auditor (b); change of disturbance. 20.7 The non-crushed MPC control law has controller gain matrix: Kc = (STQS+R)-1STQ for this exercise, the parameter values are: m = r = 1 (SISO), Q=I, R=1 and M=1 So (20-65) becomes Kc = (STQS+R)-1STQ Which reduces to a series vector: Kc = [S1 S 2 S 3 ... S P] P • S i =1 20 - 10 2 i +1 20.8 (Use MATLAB Model Prognostic Control Toolbox) a) M=5 vs. M=2 XD XB 1.5 1 y(t) 0.5 0 -0.5 0 20 40 4 060 80 100 120 140 Time (min) 0.2 R S 0.1 u(t) 0 -0.1 -0.2 0 20 40 60 80 100 120 140 Time (min) Figure S20.8a1. Simulations for P=10, M=5 and R=0.1I. 2 XD XB 1.5 1 y(t) 0.5 0 -0.5 0 20 40 60 80 100 120 140 Time (min) 0.. 2 R S 0.1 u(t) 0 -0.1 -0.2 0 20 40 60 80 100 120 Time (min) Figure S20.8a2. Simulations for P=10, M=2 and R=0.1I. 20 - 11 140 b) R=0.1I .vs R=I 2 XD XB 1.5 1 y(t) 0.5 0 -0.5 0 20 40 60 80 100 120 140 Time (min) 0.2 R S 0.1 u(t) 0 -0.1 -0.2 0 20 40 60 80 100 120 140 Time (min) Figure S20.8b1. Simulations for P=10, M=5 and R=0.1I. 2 XD XB 1.5 1 y(t) 0.5 0 -0.5 0 20 40 60 80 100 120 140 Time (min) 0.2 R S 0.. 1 u(t) 0 -0.1 -0.2 0 20 40 60 100 120 140 Time (min) Figure S20.8b2. Simulations for P=10, M=5 and R=I. Notice that the higher control horizon M and the lower input weighting factor R, the more control effort is required. 20 - 12 20.9 The GP's open loop unit step response(s) is e −6 s 1 - 10 - 1 − (t − 6) / 10 = L-1 e −6 s − y (t) = L-1 = S (t − 6) 1 − e s 10 s + 1 10 s + 1 s [ ] When testing and error, y(34) < 0.95, y(36) > 0.95. Therefore NDt =36 or N = 18. The coefficients {S i } are taken from the expression for y(t) and the forecast controller is obtained in accordance with the procedure set out in example 20.5. Closed loop responses for a unit configuration point change are shown below for the three controller design configuration sets. 20.10 Note: These results were created using the PCM oven unit, option MPC c) Change CO2 adjustment point adjustment point responses in Figs. S20.10a and . S20.10b demonstrate that the increase in the elements of matrix R makes the controller conservative and leads to more sluggish reactions. 20 - 13 Figure S20.10a. Change in CO2 adjustment point from 0.922 to 1.0143 for P=20, M=1 and Q = diag [0.1, 1]. The two series represent R = diag [0.1, 0.1] and R = diag [0.5, 0.5]. 20 - 14 d) Step disturbance in hydrocarbon flow rate Disturbance responses in Figure Fig. S20.10b is sluggish after an initial period of oscillation, and both MVs change very slowly. When the diagonal elements of matrix R increase to 0.5, the disturbance reactions are even slower. Figure S20.10b. The two series represent R = diag [0.1, 0.1] and R = diag [0.5, 0.5]. 20.11 Repeat 20.10 for R [0.1 0.1], Q = [0.1 1] and (a) M=1 and (b) M=4 First we evaluate the controller's response to a gradual change in the oxygen concentration adjustment point from 0.922 to 1.0143. 20 - 15 Figure S20.11a. Gradual change in the oxygen concentration adjustment point for P=20, Q = diag [0.1, 1], R = diag [0.1, 0.1] and M=1 or M=4. Then we will try a step change in the hydrocarbon flow rate from 0.035m3/min to 0.038m3/min. 20 - 16 Figure S20.11b. Gradual change in fuel gas flow rate for P=20, Q = diag [0.1, 1], R = diag [0.1, 0.1] and M=1 or M=4. 20.12 Note: These results were created by using the PCM distillation column section, MPC option For parts a) and (b), the step response for the models was created in the workspace. The PCM distillation column unit then opened. The controller parameters were entered into the MPC controller as defined in parts (a) and (b). Subsequently, the tests described in parts (c)-e) were carried out for each controller. The results are presented below. c) Change step at the xD adjustment point from 0.85 to 0.8 20 - 17 Figure S20.12a. Change step at setting point xD from 0.85 to 0.8 for Q=diag [0.1 0.1] and Q=diag [0.5 0.5]. 20 - 18 (d) Change step at setting point xB from 0.15 to 0.20 Figure S20.12b. Change step at setting point xB from 0.15 to 0.2 for Q=diag [0.1 0.1] and Q=diag [0.5 0.5]. 20 - 19 e) Change step in column feed flow rate from 0.025 to 0.03 Figure S20.12a. Change step in column feed flow rate from 0.025 to 0.03 for Q=diag [0.1 0.1] and Q=diag [0.5 0.5]. 20.13 We repeat problem 20.12, but this time we are looking at the case where R = [0.1 1], Q = [0.1 0.1] and M=1 or M=5. The same three tests shall be repeated from 20.12. c) Gradual change of xD from 0.85 to 0.8 20 - 20 Figure S20.13a. Change step at xD setting point from 0.85 to 0.8 for M=1 and M=5. 20 - 21 (d) Change step at the xB adjustment point from 0.15 to 0.20 Figure S20.13b. Change step at the xB adjustment point from 0.15 to 0.2 for M=1 and M=5. 20 - 22 e) Step change in column feed flow rate from 0.025 to 0.03 Figure S20.13c. Change step in flow rate from 0.025 to 0.03 for M=1 and M=5. 20 - 23 Chapter 21 © 21.1 No. It is desirable that the minimum value of the output signal be greater than zero in order to easily detect the damage to the instrument. Thus, for a conventional electronic instrument, an output signal mA indicates that a malfunction has occurred, such as a power failure. If the instrument range was 0-20 mA, instead of 4-20 mA, the output signal could be zero during normal operation. Thus, organ failures would be more difficult to detect. 21.2 The difference between measurement 6.0 and the sample condition, 5.75, is 0.25 pH units. Because the standard deviation is s = 0.05 pH units, this measurement is five standard deviations from the average. If the pH measurement is distributed normally (a hypothesis), then fig. 21.7 indicates that the probability of the measurement being less than or equal to five standard deviations from the average is 0.99999943. Thus, the probability of p of a measurement being greater than five standard deviations from the average is only p =1 - 0.99999943 = 5.7x10-7. Therefore, the probability of a measurement being greater than five standard deviations from the average is half of this value, p/2 or 2.85x10-7. A very small value! 21.3 Make the usual SPC assumption that temperature measurement is distributed normally. According to Eq. 21-6, the probability that the measurement is within three standard deviations from the average is 0.9973 Thus, the probability that a measurement is beyond these limits during routine operation is p = 1- 0.9973 = 0.0027. From Eq. 21-19, the average ARL run length among false positives is, 1 ARL • 366 samples p 21-1 [Press here] [Press here] [Press here] [Press here] So, for a sampling period of one minute, on average we would expect a false positive every 366 minutes. Thus, for a period of eight hours, the expected number of false alarms N is given by: N (8 h)(60samples / h) • 1.31 1 false alarm 366 samples/false alarm 21.4 Let's indicate the desired probability. (a) p = (0.95)3 = 0.857 b) p = (0.05)3 = 1.25 x 10-4 c) There is a much better approach. The median of the three measurements is much less sensitive to sensor failure. Thus, it should be used instead of average. 21.5 A recording of the data in Figure S21.5 does not indicate abnormal behavior. 0.85 0.84 0.83 0.82 Impurity (%) 0.81 0.8 0.79 0.78 0 2 4 6 8 10 Sample number Figure S21.5. Impurity data for exercise 21.5. The following statistics and graph limits can be calculated from the data: UCL = T + 3 = 0.8 + 3(0.021) = 0.863 % LCL = T - 3 = 0.8 - 3(0.021) = 0.737 % 21-2 Figure S21.5 shows that all eight data points are within the limits of the Shewhart chart. A standard CUSUM chart (k=0.5, h=5) also shows no chart violations, as the CUSUM chart limit is h = 5 • • , and neither C+ or C- calculated by Eq. 21-21 and 21-22 exceed this limit. CUSUM calculations are shown in table S21.5. Table S21.5. CUSUM calculations for exercise 21.5 Impurity (%) Deviation from target 1 0.812 0.012 0.0015 0 2 0.791 -0.009 0.0015 0 3 0.8 41 0 0.041 0.0320 0 4 0.814 0.014 0.0355 0 5 0.799 -0.001 -0.001 0 6 0.833 0.033 0.0580 0 7 0.815 0.015 0.0625 0 8 8 8 190.807 0.007 0.0625 0 CUSUM + CUSUM Day - 21.6 (a) The Shewhart diagram for rain data is shown in Fig. S21.6a. The following data were calculated from the data for 1870-1919: s = 7.74 in. x =18.6 in. UCL = 41.9 inches. LCL = - 4.71 in. (actually zero) Rainfall exceeded a chart limit for only one year, 1941. 21-3 50 UCL 40 Rainfall (in) 30 20 10 0 LCL -10 1860 1880 1900 1920 1940 Year 1960 1980 2000 Figure S21.6a. Shewhart chart for rainfall data. The control diagram for the standard deviation of subgroup data (for each decade) is shown in the Figure. S21.6b. The following data were calculated for the data in the subgroup before 1940: s = 6.87 in. UCL = B4 s = (1,716)(6.87 inches) = 11.8 inches. LCL = B3 s = (0.284)(6.87 inches) = 1.95 inches. Subgroup data does not violate chart limits for 1940-1990. 12 Standard deviation (in) (b) UCL 10 8 6 4 2 0 1940 LCL 1950 1960 1970 Year (end of decade) 1980 1990 Figure S21.6b. Typical deviations for sub-groups. 21-4 21.7 The CUSUM and EWMA control diagrams for the period 1900-1960 are shown in Figure S21.7. The Shewhart chart and data also appear at the top for comparison purposes. The following graph statistics and limits were calculated from the data for 1900 to 1929: s = 7.02 in. Shewhart CUSUM EWMA control diagram x =19.2 inches. UCL (in.) 40.2 35.1 27.1 LCL (in.) - 1.9 (actually zero) 0 11.2 Rainfall exceeded a Shewhart chart limit for only one year, 1941 the wettest year in the data set. The CUSUM chart has both high (C+) and low (C-) chart violations in the original period, 1900-1929. Two subsequent low-limit violations occurred after 1930. After each CUSUM violation, the corresponding amount was reset to zero. No chart violations occur for the EWMA chart and the entire dataset. The CUSUM diagram shows that the period from 1930 to 1950 had two dry periods, while the Shewhart diagram identifies a wet spell. Rainfall during the 1950s was quite normal. 21-5 Rainfall (in) 40 UCL 20 0 LCL 1900 1910 1920 1930 1940 1950 1960 CUSUM 40 UCL 30 C + C - 20 10 0 1900 1 910 1920 1930 1940 1950 1960 1910 1920 1930 1940 1950 1960 EWMA 30 25 UCL 20 15 10 1900 LCL Year Fig. S21.7. Control diagrams for rainfall data. 21.8 In general, it is preferable to draw un filtered measurements because they contain most of the information. However, it is important to be consistent. Therefore, if the limits of the control diagram were calculated on the basis of un filtered data, un filtered measurements should be designed for subsequent monitoring. if chart limit calculations were based on filtered data, filtered measurements should be designed. 21.9 The control diagrams in figure. S21.9 does not show any violations of

the control diagram. Thus, the performance of the procedure is considered normal. The CUSUM chart was designed using the default values K = 0.5s = 0.5s and H = 500 = 5s where s s the standard deviation of the sample. The EWMA chart was designed using -=0.25. 21-6 BOD (mg/L) 40 UCL 20 LCL 0 0 5 10 15 20 25 30 15 20 25 30 15 20 20 20 20 Sample number 20 25 30 CUSUM 30 UCL 20 C+ 10 0 0 EWMA 30 C- 5 10 5 10 UCL 20 10 LCL 0 Figure S21.9. Control charts for the BOD data in example 21.5. 21.10 By definition, Cp USL − LSL 6▦ (21 − 25) Because the standard population deviation ▦ is not known, it should be replaced by an estimate, ▦aaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaaa Leave ▦ = s where s is the standard deviation of the sample. The standard deviation of bod data is s = 5.41 mg/L. Substitution gives, Cp • 35 − 5 6 (5.41) • 0.924 21-7 Cpk Capacity Index is defined as: min [x − LSL, USL − x] 3▦ C pk (21 − 26) The sample average for BOD data is x = 20.6 mg/L. Replacing numeric values in (21-26) gives: C pk • min [20.6 − 5, 35 − 20.6] • 0.887 3(5.41) Because both capacity indicators are less than one, the product specifications are not met and the process is considered to be underperforming. 21.11 By default, Cp USL − LSL 6▦ (21 − 25) Because the standard population deviation ▦ is not known, it should be replaced by an estimate, ▦yes . Leave ▦ = s where s is the standard deviation of the sample. The standard deviation of solid data is s = 56.3 mg/L. The substitution gives, Cp • 1600 1600 − 1200 6 (56.3) • capacity index 1.18 CPk [x − LSL, USL − x] 3▦ (21 − 26) The sample average for solid data is x = 1413 mg/L. Replacing numeric values in (21-26) gives: C pk • min [1413 − 1200, 1600 − 1413] • 1.26 3(56.3) Because both capacity indicators are well below the acceptable value of 1.5, the process is considered to be underperforming. 21-8 21.12 The new data is shown in a T2 diagram in the Figure. S21.12. A chart violation occurs for the second data point. Because one of the six measurements is beyond the chart limit, it appears that the process behavior may be abnormal. However, this measurement may be an extreme value and further research is therefore recommended. Additional data should also be collected before concluding that the operation of the process is abnormal. Note that the previous control chart limit of 11.63 from example 21.6 is also used in this exercise. 14 13 12 UCL 11 T2 10 9 8 7 6 5 0 1 2 3 4 Sample number 5 Figure S21.12. T2 Control diagram and new sewage data. 21-9 6 Chapter 22 22.1 Microwave Operating States Condition Open the door Place the food inside Close the door Set the time Heat up food Cooking complete Fan Light Timer Rotating Base Microwave Generator Door Switch OFF ON OFF OFF OFF ON OFF OFF ON OFF OFF OFF ON OFF OFF OFF ON OFF OFF OFF ON OFF OFF OFF ON OFF OFF OFF ON OFF OFF OFF OFF OFF Safety Issues: o Door switch is OFF before the microwave generator is turned ON. o Fan always ON when microwave generator is ON. 22.2 Input Variables: ON STOP EMERGENCY Output Variables: START START (1) (0) Solution Manual for Dynamics and Process Control, 3rd Edition Copyright © 2011 by Dale E. Seborg, Thomas F. Edgar and Duncan A. Mellichamp, and Francis J. Doyle III 22-1 Truth Table On 1 0 1 0 1 0 1 0 STOP 1 1 0 0 2 0 1 0 0 EXECUTIVE 1 1 1 1 0 0 0 0 START/STOP 0 0 0 0 0 0 1 0 The truth status table is used to find the logical law concerning output inputs: ON □ STOP □ EXECUTIVE • EXECUTIVE Application of Binary Algebra we can receive an equivalent expression: ON □ (STOP □ EMERGENCY) • ON □ (STOP) (EMERGENCY) Finally, the diagrams of binary logic and ladder are given in figure S22.2 : Binary logic diagram: In the stop emergency ladder logic diagram 22-2 start CR1 CR1 CR1 stop CR2 CR2 CR3 TH CR3 m shape S22.2. 22.3 A 0 1 0 1 B 0 1 1 Y 1 1 0 1 From the truth table it is possible to find the logical function that gives the desired result, A□ B Since a NAND gateway is equivalent to an OR gateway with two un undoed inputs, our expression is reduced to: A B □ A • B Finally, the binary logic diagram is given in figure S22.3. Figure A B S22.3. 22-3 22.4 information flowchart START input valve open output valve closing Stop No L>LH stirrer for yes yes stop no T = Tsetpoint input valve closed output valve open valve stirrer open input valve open stirrer off yes stop LTH heat off open V2 P2 for No L = L0 Close V2 P2 OFF 22-7 Ladder Logic Diagram: R1= Pump 1 R2= Valve 2 R3= Starting heater CR2 R4= Pump 2 CR1 L1 CR1 CR3 TH L0 CR3 CR4 TH CR2 L0 CR4 Consecutive operating diagram: 1 Initial step B Thermal filling 1 V1 Q Fill 2 1 Temp V1 L1 Full TH 1 4 V1 L0 Figure S22.5. 22-8 22.6 Information flow diagram: START L L