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UNAPREĐENJE VOĐENJA PROCESA ATMOSFERSKE DESTILACIJE MAKSIMIZACIJOM PREDGRIJAVANJA SIROVINE

Sažetak

Proces atmosferske destilacije je eksperimentalno istražen. Na osnovi analize i sinteze vođenja procesa postavljen je viševeličinski dinamički matematički model procesa atmosferske destilacije. Odgovarajući model primijenjen je u svrhu optimalnog vođenja primjenom viševeličinske jedinice za vođenje. Rezultati pokazuju da se primjenom viševeličinske jedinice za vođenje postiže veća stabilnost procesa te vođenje procesa primjenom kriterija optimalnosti temeljenog na maksimizaciji temperature predgrijavanja.

Uvod

Proces atmosferske destilacije primarni je proces u preradi nafte kojim se godišnje najčešće preradi i više od 2 milijuna tona nafte. Proces se sastoji od sekcije za predgrijavanje, sekcije peći, sekcije atmosferske kolone i sekcije za stabilizaciju i splitiranje. S obzirom da veći kapaciteti prerade zahtijevaju i veću potrošnju energije, nova postrojenja se projektiraju s ciljem maksimalnog iskorištenja topline u sekciji predgrijavanja, dok se na starijim postrojenjima poboljšanja postižu rekonstrukcijama i unapređenjima postrojenja kako bi se postigao isti cilj.

Viševeličinsko optimalno vođenje već se 25 godina primjenjuje u praksi pri projektiranju novih procesnih postrojenja, te pri unapređenju postojećih. (1, 2)

Velik broj instalacija u cijelom svijetu potvrđuju da je viševeličinsko optimalno vođenje najprimjenjenija tehnologija unapređenja procesnih postrojenja.

Viševeličinsko optimalno vođenje razvijeno je na temelju teorijskih osnova vođenja procesa i razvoja primjene računala, a u praksi predstavlja niz alata koji omogućavaju kontinuirano djelovanje specijaliziranih računalnih programa na regulacijske krugove procesa, a sve u svrhu kontinuiranog provođenja kriterija optimalnosti u stvarnom vremenu. Primjenom viševeličinskog optimalnog vođenja u

praksi, provodi se optimalno vođenje, vođenje sa svrhom: maksimizacije iskorištenja kapaciteta postrojenja, održavanja konstantne kvalitete pojedinih proizvoda, minimizacija potroška energije, itd. (3)

Prednosti primjene viševeličinskog optimalnog vođenja u odnosu na klasični način vođenja je konstantno i kontinuirano provođenje zadanog kriterija vođenja, a to se u praksi očituje u povećanju stabilnosti procesa, smanjenju varijabilnosti procesnih veličina, smanjenju troškova održavanja, povećanju iskorištenja kapaciteta, smanjenju potroška energije, što sve utječe na konačni cilj povećanja dobiti.

Proces atmosferske destilacije je uz fluid katalitički kreking jedan od najzahtjevnijih rafinerijskih procesa za primjenu viševeličinskog optimalnog vođenja. Procesom atmosferske destilacije proizvodi se 8 produkata koji moraju zadovoljiti zahtjeve kvalitete proizvoda pa je primjena viševeličinskog optimalnog vođenja na procesu atmosferske destilacije veliki izazov.

Jedan od najvažnijih ciljeva za primjenu viševeličinskog optimalnog vođenja na postrojenju atmosferske destilacije je minimizacija potroška energije. Minimizacija potrošnje loživog ulja i plina postiže se boljim iskorištenjem topline predgrijavanjem sirovine. Zbog toga je za primjenu viševeličinskog optimalnog vođenja postavljen kriterij optimalnosti maksimizacija temperature ulaza sirovine u peć.

Pristup unapređenju

Sustav viševeličinskog optimalnog vođenja primijenjen je na postrojenju atmosferske destilacije Rafinerije nafte Rijeka. Proces je prikazan slikom 1 i tablicom 1.

Hladna sirovina uvodi se u sekciju za predgrijavanje kroz kondenzatore E-28A/B, gdje se zagrijava parama vrha atmosferske kolone. Sirovina se zatim provodi kroz izmjenjivače topline E-1 i E-3ABC u kojima se zagrijava pomoću lakog plinskog ulja i atmosferskog ostatka, te uvodi u odvajač parne faze C-6. Iz odvajača parne faze C-6 plinoviti se ugljikovodici odvođe u kondenzatore E-29A/B, dok se proizvod dna C-6 pomoću crpke provodi kroz izmjenjivače E-4ABCDEF, E-2A/B, E-5ABC i E-6 u kojima se predgrijava gornjim kružnim pretokom, lakim plinskim uljem, atmosferskim ostatkom i donjim kružnim pretokom, te zatim uvodi u peć F-1.

Protok sirovine se regulira pomoću regulacije protoka FC5_12. Izlazna temperatura iz peći regulira se pomoću regulacije temperature TC5. Temperatura vrha atmosferske kolone se regulira pomoću regulacije temperature TC7, dok se protoci gornjeg i donjeg kružnog pretoka reguliraju djelovanjem regulacija protoka FC14 i FC18. Protoci bočnih proizvoda teškog benzina, lakog i teškog plinskog ulja reguliraju se pomoću regulacija protoka FC2, FC3 i FC18. Protok atmosferskog ostatka regulira se u kaskadi pomoću regulacije razine atmosferske kolone LC3, dok je ulazna temperatura u peć mjerena pomoću T114.

Maksimizacija temperature predgrijavanja, koja omogućava smanjenje potrošnje loživog ulja i plina, postavlja se kao jedan od kriterija vođenja. S ciljem postizanja zadanih kriterija rukovodstvo Rafinerije nafte Rijeka odlučilo se za provođenje unapređenja vođenja procesa primjenom sustava viševeličinskog optimalnog

vođenja. U tu je svrhu primijenjen programski sustav tvrtke Honeywell: Robust Model Predictive Control Technology (RMPCT). Osnova pristupa unapređenju procesa atmosferske destilacije je detaljna analiza procesa, sinteza viševeličinskog vođenja, te definiranje kriterija optimalnosti.

Budući da klasičan način vođenja temeljen na pristupu kojim se proces definira na osnovi više sustava s jednim ulazom i izlazom ne omogućuje provedbu optimalnog vođenja, primijenjen je pristup opisa procesa kao sustava s više ulaza i izlaza.

Kao kriterij optimalnosti unapređenja vođenja u praksi postavljen je kriterij minimizacije potroška energije.

Iako su destilacijski procesi nelinearni, u uskom se području oko radne točke dinamika procesa opisuje viševeličinskim linearnim dinamičkim modelom, pa je viševeličinski dinamički model procesa atmosferske destilacije razvijen primjenom diskretne jednadžbe, prikazanom u tablici 2.

Iz definirane jednadžbe slijedi da se prema promjenama procesnih veličina u prošlosti mogu predvidjeti buduće promjene. Zbog toga se na stvarnom postrojenju provodi eksperimentalno istraživanje s ciljem određivanja jednadžbe prikazane tablicom 2.

Prvi korak u stvaranju viševeličinske jedinice za vođenje je definiranje ulaznih, poremećajnih i izlaznih veličina, prikazanih u tablici 3.

Prije testiranja postrojenja svaka je nezavisna veličina provjerena u prethodnom testiranju. Prethodnim testiranjem je provjereno i zaključeno da li se sve nezavisne veličine mogu mijenjati i u kojoj mjeri, kako se testiranjem ne bi narušila sigurnost rada postrojenja. Broj ulaznih veličina i vrijeme do stacionarnog stanja određuju plan eksperimenta. Test se izvodi namjernim izvođenjem poremećaja različitim po dužini trajanja i po veličini amplitude. Pri testiranju procesa izvode se dovoljno veliki poremećaji kako bi se izbjegla pojava šuma.

Za vrijeme izvođenja testa eksperimentalni podaci se prikupljaju pomoću računala i programskog sustava Process Historical Data (PHD) tvrtke Honeywell. Najvažniji podaci pri provedbi eksperimentalnog istraživanja prikazani su tablicom 4.

U svrhu identifikacije viševeličinskog dinamičkog matematičkog modela primijenjena je metoda jediničnog impulsnog niza. Model identifikacije određen je za svaki par ulaznih i izlaznih veličina. Pritom je cjelokupna analiza podataka, određivanje parametara modela kao i procjena valjanosti modela provedena pomoću RMPCT programskog sustava tvrtke Honeywell.

Dinamički matematički modeli zavisnosti ulaznih, poremećajnih i izlaznih veličina prikazani su slikom 2.

U svrhu analize valjanosti modela, modeli odziva izlaznih veličina na poremećaje promatraju se u 3 važna uvjeta: da li model ima fizičkog smisla, da li model opisuje eksperimentalne podatke, te da li je nesigurnost modela prihvatljiva? Na ta pitanja odgovori su sljedeći: modeli imaju fizičkog smisla, opisuju eksperimentalne podatke i nesigurnost modela je prihvatljiva.

Maksimizacija predgrijavanja

Opisani viševeličinski dinamički matematički model je prihvaćen kao dio viševeličinske jedinice za vođenje, te se primjenjuje u svrhu optimizacije i unapređenja vođenja procesa. Djelovanje viševeličinske jedinice za vođenje je kontinuirano, a temelji se na određenom viševeličinskom dinamičkom matematičkom modelu, te omogućava provođenje kriterija optimalnosti svake minute.

Viševeličinska jedinica za vođenje instalirana je na računalu i programskom sustavu Application module (APP) tvrke Honeywell, a vođenje procesa se odvija pomoću programskog sustava RMPCT. Svake minute djelovanjem na ulazne veličine izvršava se kriterij optimalnosti maksimizacije temperature predgrijavanja. Osim tog kriterija, kao kriteriji optimizacije mogu se postaviti i drugi kriteriji, npr. minimizacija iscrpka atmosferskog ostatka. Određivanje kriterija optimizacije ovisi o ekonomskoj bilanci postrojenja.

U svrhu maksimizacije temperature predgrijavanja, viševeličinska jedinica za vođenje može djelovati na 6 ulaznih veličina: ukupan protok sirovine, FC5_12.SP, temperatura izlaza iz peć, TC5.SP, temperatura vrha atmosferske kolone, TC7.SP, protok gornjeg kružnog pretoka, FC14.SP, protok donjeg kružnog pretoka, FC18.SP i protok lakog plinskog ulja, FC3.SP. Time je viševeličinskoj jedinici za vođenje omogućeno 5 stupnjeva slobode, no stupnjevi slobode će biti manji ako se u kriterij optimalnosti uključe sljedeća ograničenja: konstantan protok sirovine, konstantna temperatura sirovine na izlazu iz peći i konstantan protok lakog plinskog ulja.

Na osnovi tako definiranog kriterija optimalnosti, uključenjem viševeličinske jedinice za vođenje počinje djelovanje viševeličinske jedinice za vođenje na ulazne veličine. Slikom 3 prikazane su aktivnosti viševeličinske jedinice za vođenje od trenutka njezinog pokretanja, pa u idućih nekoliko sati.

Prednost viševeličinske jedinice za vođenje u usporedbi s klasičnim vođenjem je u mogućnosti viševeličinske jedinice da svake minute pronalazi optimalnu radnu točku s obzirom na trenutačne radne uvjete i ograničenja u procesu.

Pokretanjem viševeličinske jedinice za vođenje počinje utjecaj na ulazne veličine, pa se mijenjaju vrijednosti sljedećih ulaznih veličina: temperature vrha atmosferske kolone, protok gornjeg kružnog pretoka i protok donjeg kružnog pretoka. Postizanjem maksimalne temperature predgrijavanja u trenutačnim radnim uvjetima prestaje daljnje djelovanje na ulazne veličine s ciljem povišenja temperature predgrijavanja, a proces se vodi na način da se zadrže optimalni radni uvjeti.

Primjenom viševeličinske jedinice za vođenje postiže se povišenje temperature predgrijavanja za 30 °C. S pretpostavkom da sustav naprednog vođenja radi konstantno i da se kao rezultat rada postiže povišenje temperature za 3 °C, procjene godišnjih ušteda mogu biti i više od 200 000 USD.

Zaključak

Postrojenje atmosferske destilacije kao veliki energetska potrošač unaprijeđeno je primjenom eksperimentalnog testa, identifikacijom dinamičkog matematičkog modela i razvijanjem viševeličinske jedinice za optimalno vođenje. Vođenje procesa primjenom viševeličinskog dinamičkog matematičkog modela i kriterija optimalnosti pokazalo se superiornijim načinom vođenja u usporedbi s klasičnim vođenjem u postizanju cilja maksimizacije predgrijavanja sirovine i veće stabilnosti procesa.

Tablica 1: Radni uvjeti procesa atmosferske destilacije

Protok sirovine, FC5_12.SP, t/h	415.0
Temperatura izlaza iz peći, TC5.SP, °C	368.0
Temperatura vrha atmosferske kolone, TC7.SP, °C	170.0
Protok gornjeg kružnog pretoka, FC14.SP, t/h	320.0
Protok donjeg kružnog pretoka, FC18.SP, t/h	40.0
Protok teškog benzina, FC2.SP, t/h	30.0
Protok lakog plinskog ulja, FC3.SP, t/h	114.0
Protok teškog plinskog ulja, FC27.SP, t/h	22.0
Tlak atmosferske kolone, PC19.SP, bar(g)	1.0
Temperatura ulaza u peć, TI14.PV, °C	248,3

Tablica 2: Osnovna jednadžba zbroja umnoška jediničnog impulsnog niza i koeficijenata odziva na jedinični impulsni niz

$$y_p(k+n) = \sum_{i=1}^N h_i u(k+n-i) + b(k)$$

$$b(k) = y_m(k) \sum_{i=1}^N h_i u(k-i)$$

Gdje je:

- k = trenutno vrijeme uzorkovanja,
- k+n = buduće vrijeme uzorkovanja time,
- u = vrijednost ulazne/poremećajne veličine,
- y_m = izmjerena vrijednost izlazne veličine,
- y_p = predviđena vrijednost izlazne veličine,
- h_i = koeficijent odziva na jedinični impulsni niz,
- b = odstupanje modela.

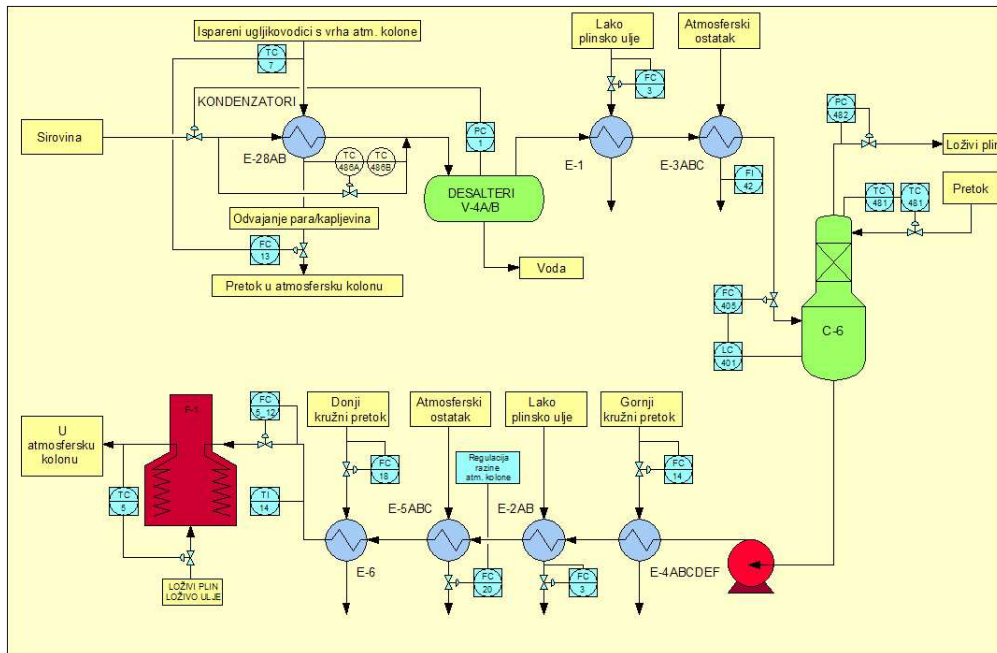
Tablica 3: Ulazne, poremećajne i izlazne veličine procesa

Ulazne veličine, MV	
FC5_12.SP	Regulacija protoka sirovine
TC5.SP	Regulacija temperature na izlazu iz peći
TC7.SP	Regulacija temperature vrha atmosferske kolone
FC14.SP	Regulacija protoka gornjeg kružnog pretoka
FC18.SP	Regulacija protoka donjeg kružnog pretoka
FC2.SP	Regulacija protoka teškog benzina
FC3.SP	Regulacija protoka lakog plinskog ulja
FC27.SP	Regulacija protoka teškog plinskog ulja
PC19.SP	Regulacija tlaka u atmosferskoj koloni
Poremećajne veličine, DV	
FC32.PV	Protok atm. ostatka u isparivač kolone za stabilizaciju
FC34.PV	Protok atm. ostatka u isparivač kolone za splitiranje
Izlazne veličine, CV	
TI14.PV	Temperatura ulaza u peć
FI42.PV	Protok atmosferskog ostatka

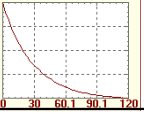
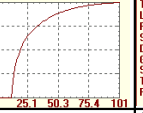
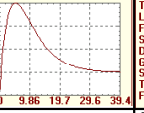
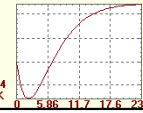
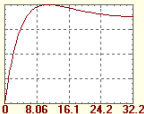
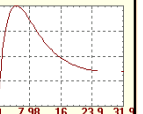
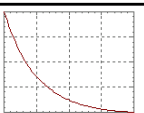
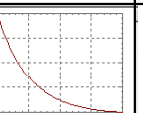
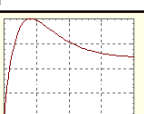
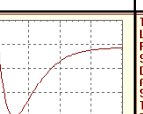
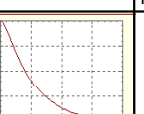
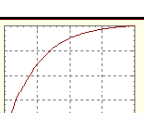
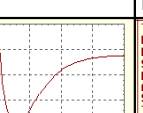
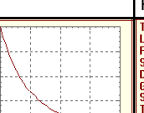
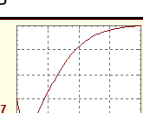
Tablica 4: Osnovni podaci pri provedbi eksperimentalnog istraživanja

Vrijeme uzorkovanja, min	1
Broj uzoraka	9256
Maksimalno trajanje poremećaja, min	120
Minimalno trajanje poremećaja, min	30
Maksimalna amplituda promjene protoka sirovine, t/h	60.0
Minimalna amplituda promjene protoka sirovine, t/h	20.0
Maksimalna amplituda promjene temperature izlaza iz peći, °C	5.0
Minimalna amplituda promjene temperature izlaza iz peći, °C	2.0
Maksimalna amplituda promjene temperature vrha atm. kolone, °C	10.0
Minimalna amplituda promjene temperature vrha atm. kolone, °C	2.0
Maksimalna amplituda promjene protoka gornjeg kružnog pretoka, m ³ /h	150.0
Minimalna amplituda promjene protoka gornjeg kružnog pretoka, m ³ /h	20.0
Maksimalna amplituda promjene donjeg gornjeg kružnog pretoka, m ³ /h	10.0
Minimalna amplituda promjene protoka donjeg kružnog pretoka, m ³ /h	5.0
Maksimalna amplituda promjene protoka teškog benzina, m ³ /h	5.0
Minimalna amplituda promjene protoka teškog benzina, m ³ /h	2.0
Maksimalna amplituda promjene protoka lakog plinskog ulja, m ³ /h	15.0
Minimalna amplituda promjene protoka lakog plinskog ulja, m ³ /h	5.0
Maksimalna amplituda promjene protoka teškog plinskog ulja, m ³ /h	19.0
Minimalna amplituda promjene protoka teškog plinskog ulja, m ³ /h	3.0
Maksimalna amplituda promjene tlaka atmosferske kolone, bar (a)	0.4
Minimalna amplituda promjene tlaka atmosferske kolone, bar (a)	0.1

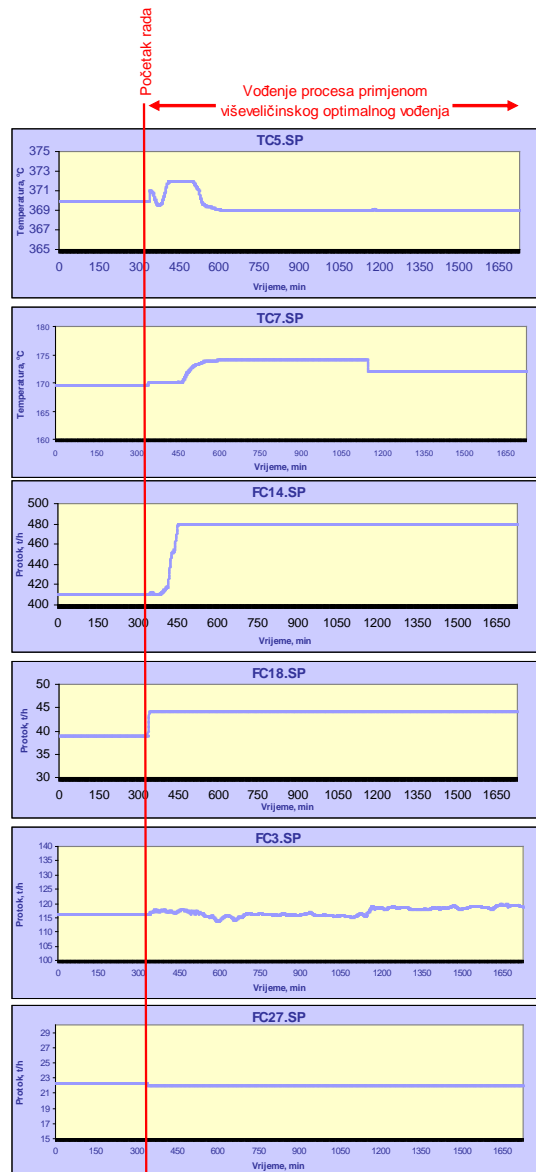
Slika 1: Procesna shema predgrijavanja atmosfenske destilacije



Slika 2: Dinamički matematički modeli zavisnosti ulaznih, poremećajnih I izlaznih veličina.

	MV1 FC5_12.SP	MV2 TC5.SP	MV3 TC7.SP	MV4 FC14.SP
CV1 TI14.PV	 Trial 1 Lap Order 1 FIR Order 0 Stable Dead T = 0 Gain = -0.884 Settle T = 0 TISettle = 120 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 12.0 Gain = 626 Settle T = 0 TISettle = 101 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = 230 Settle T = 0 TISettle = 39.4 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = 0.170 Settle T = 0 TISettle = 23.4 FIR Form = UK
	MV5 FC18.SP	MV6 FC3.SP	MV7 FC27.SP	MV8 PC19.SP
CV1 TI14.PV	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = 1.44 Settle T = 0 TISettle = 32.2 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = 0.514 Settle T = 0 TISettle = 31.9 FIR Form = UK	-	-
	DV1 FC32.PV	DV2 FC34.PV		
CV1 TI14.PV	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -0.139 Settle T = 0 TISettle = 40.1 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -0.139 Settle T = 0 TISettle = 40.1 FIR Form = UK		
	MV1 FC5_12.SP	MV2 TC5.SP	MV3 TC7.SP	MV4 FC14.SP
CV2 FI42.PV	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = 510 Settle T = 0 TISettle = 126 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -460 Settle T = 0 TISettle = 110 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -1.30 Settle T = 0 TISettle = 64.8 FIR Form = UK	-
	MV5 FC18.SP	MV6 FC3.SP	MV7 FC27.SP	MV8 PC19.SP
CV1 TI14.PV	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = 1.60 Settle T = 0 TISettle = 30.0 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -120 Settle T = 0 TISettle = 153 FIR Form = UK	 Trial 1 Lap Order 1 FIR Order 0 Stable Dead T = 0 Gain = -140 Settle T = 0 TISettle = 53.3 FIR Form = UK	 Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = 10.8 Settle T = 0 TISettle = 19.7 FIR Form = UK
	DV1 FC32.PV	DV2 FC34.PV		
CV2 FI42.PV	-	-		

Slika 3: Djelovanje viševeličinske jedinice za vođenje pri maksimizaciji temperature predgrijavanja





CRUDE UNIT FEED PREHEAT MAXIMIZATION IN PRACTICE

Abstract

A specific dynamic mathematical model of crude unit is developed on the basis of plant dynamic test experimentation and mathematical model identification. The valid mathematical model has been applied in multivariable controller design. The practical cases of control strategies have been studied and used in practice to provide superior dynamic control performance, reduce variability in controlled variables that enables model predictive on line closed loop feed preheat maximization.

Introduction

Crude unit is basic refinery process processing almost ever more then 2 million t/y of crude. It has a preheat section, a furnace section, a crude column section and a light ends section. Higher crude capacities require higher energy consumption. New crude units are always designed to use more heat in preheat section whilst the existing units are controlled and optimized to do the same during the crude processing.

On the other hand, the advanced process control (APC) has had a major impact on the refining and petrochemical industry since its inception 25 years ago (1, 2).

With huge number of industrial installations, model predictive control (MPC) is currently the most widely implemented advanced process control technology for process plants.

MPC loops are computer applications that manipulate setpoints of regulatory controllers to control high-level process measurements, e.g., measurements representative of a unit's overall performance, such as product compositions, vessel temperatures, column differential pressures and regulatory controller outputs (3).

Some of the benefits of MPC system are better control performance, less down time, reduced maintenance requirements and improved flexibility and agility.

Crude unit, besides FCC unit, is one of the most complex processes in the refining industry for applying model predictive control. It produces 8 products that must satisfy quality requirements and therefore is designing of MPC on crude unit every time a challenge.

While applying model predictive control to crude unit, one of goals is minimizing energy consumption. Minimizing fuel gas and oil consumption is done through better preheat of crude unit feed. Therefore, control strategy is maximum furnace inlet temperature achieved with better heat exchange through downstream heat exchangers.

Improvement approach

The object of the study, crude unit preheat section in Rijeka Refinery, is considered. The process is described and summarized in Table 1 and Figure 1.

Cold feed is entered to preheat in condensers E-28A/B with column top vapor, prior feeding to desalters. Then, the feed passes through heat exchangers E-1 and E-3ABC where exchanges heat with LGO and crude residue before it's feeding into C-6 preflash column. In preflash column vapor hydrocarbons are vented to condensers E-29A/B while the bottom is pumped through heat exchangers E-4ABCDEF, E-2A/B, E-5ABC and E-6 where is preheated with middle pump around, side product LGO, crude residue and bottom pump around before it is fed into the furnace F-1.

Feed flow rate is controlled by flow controller FC5_12. Furnace outlet temperature is controlled by TC5 temperature control. Top crude column temperature is controlled by temperature controller TC7, while the middle and bottom pump around flows are controlled by FC14 and FC18. Heavy naphtha, LGO and HGO side products flows are controlled by FC2, FC3 and FC27. Crude residue flow is controlled by cascade crude column level control LC3. Inlet furnace temperature is measured by T114.

One of the important objectives of the process control is continuous energy consumption minimization, processing the crude with minimizing fuel gas and oil consumption, through maximization of feed preheat. To achieve defined objective the refinery management made decision to improve process control with MPC (Model Predictive Control) using Honeywell Robust Model Predictive Control Technology (RMPCT). Therefore, control strategy is defined and MPC controller is designed so that process control is enabled to continuously achieve maximum possible furnace inlet temperature.

One of the primary incentives for the implementation of multivariable control is to avoid the complexity and inflexibility of single loop schemes and advanced dynamic process performance and process optimization. The control strategy of process improvement in practice: energy consumption minimization and disturbance feedforward control is established. Although distillation processes are inherently nonlinear, if operated over a sufficiently small region, process control systems could be based on linear input output dynamic models. So, the multivariable model predictive control of the crude unit is developed using discrete convolution summation equation (4,5,6), shown in Table 2.

From the preceding equations, a set of future predictions is based on past moves. Therefore, to find the convolution summation equation, the experimentation investigation had to be done on commercial plant using plant tests and multivariable model identification.

The experimental plan is established and performed as follows. First, the multivariable controller is designed with the input, disturbance and output control variables shown in Table 3.

Prior to plant tests, every independent variable has been checked, using pretests phase, to make sure that each input variable can be varied. The number of input variables and the time to steady state determine the actual plant test. The protocol is consisted of pulses of differing duration, long, short and intermediate duration and different step magnitude. The step magnitude has been set high enough to avoid process noise magnitude.

During the tests the experimental data was collected using Process Historical Data (PHD) computer and Honeywell software packages. The basics of experimental plant test are shown in Table 4.

The identification of crude unit has many features that are common to most chemical engineering systems, that finite impulse response (FIR) identification method could be applied (7,8,9). Accordingly, the model identification FIR method has been made and for each input output pair one discrete impulse model has been performed. All data analysis, model structure determination, parameter estimation and model validation has been performed using Honeywell RMPCT software package.

Step response models relating all input, disturbance and output control variables have been obtained as follows in Figure 2.

From the step response models, uncertainty frequency response characteristics and model-predicted output vs. measured data comparison the models have been judged on three important factors: does the model make physical sense, does the model fit the data and is the model uncertainty reasonable? The answers on these questions are that: models make physical sense, models fit the data and models' uncertainties are reasonable.

Preheat maximization

To do only preheat maximization multivariable controller can manipulate six manipulated variables: feed flow, FC5_12.SP, furnace outlet temperature, TC5.SP, crude column top temperature, TC7.SP, middle pump around flow, FC14.SP, bottom pump around flow, FC18.SP and LGO side product flow, FC3.SP. It allows the multivariable controller five degrees of freedom. The degree of freedom is going to decrease when the following objectives are included in the control strategy: feed flow can not be increased or decreased, furnace outlet temperature has to be in narrow limits, LGO side product flow have to be maintained constant.

On the basis of defined control strategy the multivariable controller has been started. The Figure 3 shows multivariable controller control action from the moment controller starts acting and during next few hours. Before multivariable controller started running the process was in basic control.

Against the basic control the multivariable controller is searching the optimum control strategy every minute. After the multivariable controller is switched on, the controller starts moving the following manipulated variables: crude column top temperature, middle and bottom pump around flows. When preheat maximum possible temperature has been reached the multivariable controller stops all moves

of manipulated variables and maintain the preheat temperature, T114.PV, on its maximum value during its run. The result of using multivariable controller is increase in preheat temperature for 3 °C. With the assumption that multivariable controller is working constantly and that is superior than basic control in preheat temperature maximization for about 3 °C, the saving per year could be more than 200.000 USD.

Conclusion

The crude unit as a major energy consumer, has been improved in practice on commercial plant by experimental plant tests, mathematical model identification and multivariable controller design. The applied control strategies based on the cases derived from practice provided superior dynamic control performance in preheat maximization and reduced variability in controlled variables. The designed multivariable controller provides multi-objective control handling based on objective hierarchy.

Table 1: Crude Unit Steady-state Process Data

Feed flow, FC5_12.SP, t/h	415.0
Furnace outlet temperature, TC5.SP, °C	368.0
Crude column top temperature, TC7.SP, °C	170.0
Middle pump around flow, FC14.SP, t/h	320.0
Bottom pump around flow, FC18.SP, t/h	40.0
Heavy naphtha side product flow, FC2.SP, t/h	30.0
LGO side product flow, FC3.SP, t/h	114.0
HGO side product flow, FC27.SP, t/h	22.0
Crude column pressure, PC19.SP, bar(g)	1.0
Furnace inlet temperature, T114.PV, °C	248,3

Table 2: Basic Discrete Convolution Summation Equation

$$y_p(k+n) = \sum_{i=1}^N h_{iu}(k+n-i) + b(k)$$

$$b(k) = y_m(k) \sum_{i=1}^N h_{iu}(k-i)$$

where: k = current sampling time,
 k+n = nth future sampling time,
 u = input variable (manipulated or disturbance),
 y_m = measured output, control variable,
 y_p = predicted output, control variable,
 h_i = impulse response coefficient,
 b = model bias.

Table 3: Proces Input, Disturbance and Output Control Variables

Input Variables, MV's	
FC5_12.SP	Feed flow control
TC5.SP	Furnace outlet temperature control
TC7.SP	Crude column top temperature control
FC14.SP	Middle pump around flow control
FC18.SP	Bottom pump around flow control
FC2.SP	Heavy naphtha side product flow control
FC3.SP	LGO side product flow control
FC27.SP	HGO side product flow control
PC19.SP	Crude column pressure control
Disturbance Variables, DV's	
FC32.PV	Crude residue flow to Stabilization column reboiler
FC34.PV	Crude residue flow to Splitter column reboiler
Output Control Variables, CV's	
TI14.PV	Furnace inlet temperature
FI42.PV	Crude residue flow

Table 4: Basics of Experimental Plan Plant Test Data

Sample period, min	1
Number of samples	9256
Long duration pulse time, min	120
Short duration pulse time, min	30
Maximum feed flow step magnitude, t/h	60.0
Minimum feed flow step magnitude, t/h	20.0
Maximum furnace outlet temperature step magnitude, °C	5.0
Minimum furnace outlet temperature step magnitude, °C	2.0
Maximum crude column top temperature step magnitude, °C	10.0
Minimum crude column top temperature step magnitude, °C	2.0
Maximum middle pump around flow step magnitude, m ³ /h	150.0
Minimum middle pump around flow step magnitude, m ³ /h	20.0
Maximum bottom pump around flow step magnitude, m ³ /h	10.0
Minimum bottom pump around flow step magnitude, m ³ /h	5.0
Maximum heavy naphtha side product flow step magnitude, m ³ /h	5.0
Minimum heavy naphtha side product flow step magnitude, m ³ /h	2.0
Maximum LGO side product flow step magnitude, m ³ /h	15.0
Minimum LGO side product flow step magnitude, m ³ /h	5.0
Maximum HGO side product flow step magnitude, m ³ /h	19.0
Minimum HGO side product flow step magnitude, m ³ /h	3.0
Maximum crude column pressure step magnitude, bar (a)	0.4
Minimum crude column pressure step magnitude, bar (a)	0.1

Figure 1: Crude Unit Preaheat Section Flow Diagram

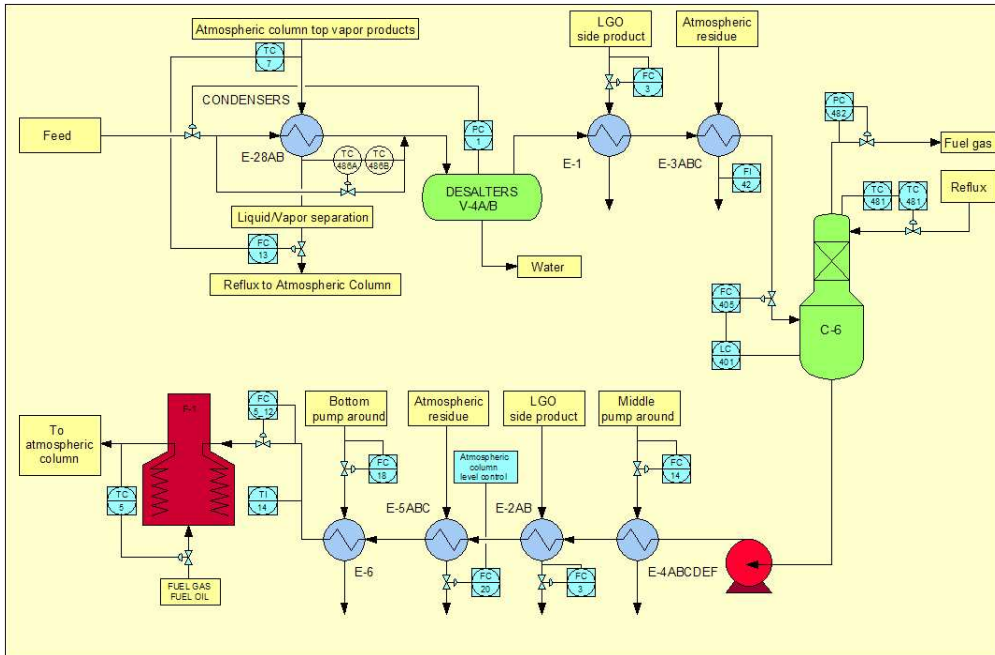


Figure 2: Step response models relating all input, disturbance and output control variables

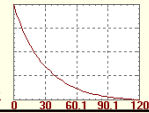
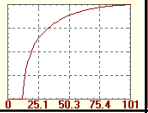
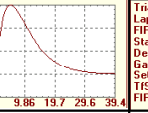
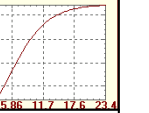
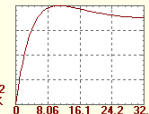
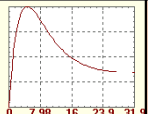
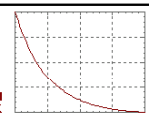
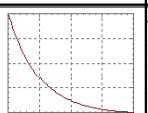
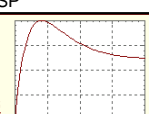
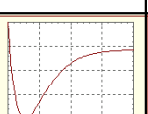
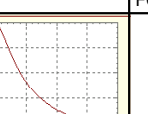
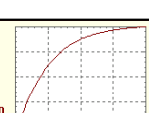
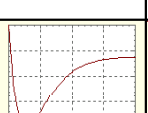
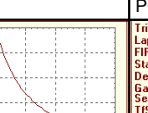
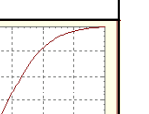
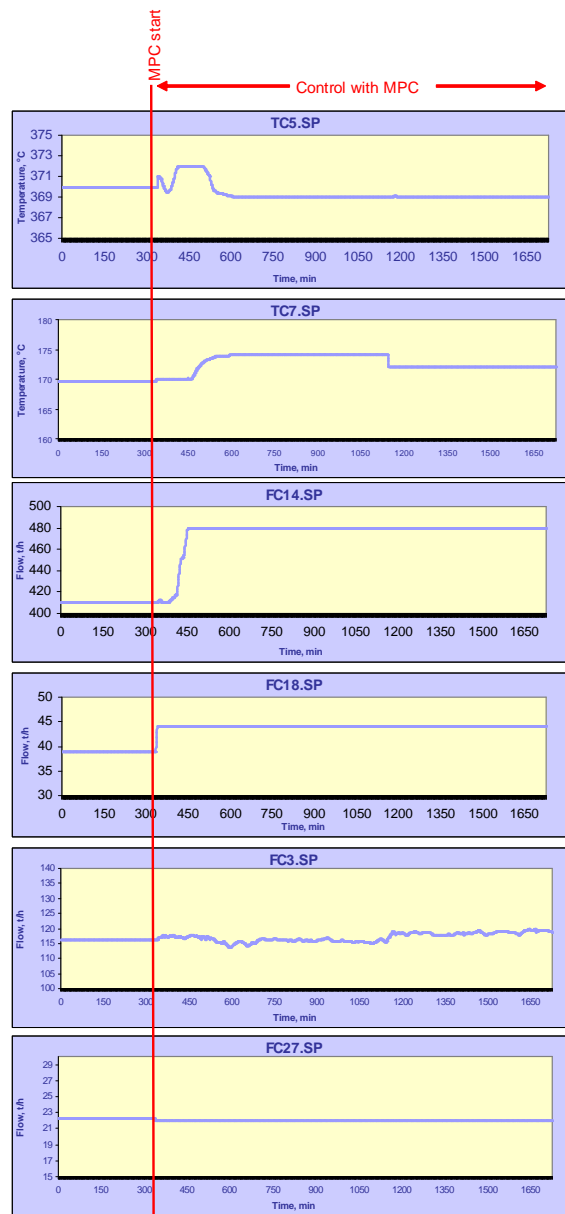
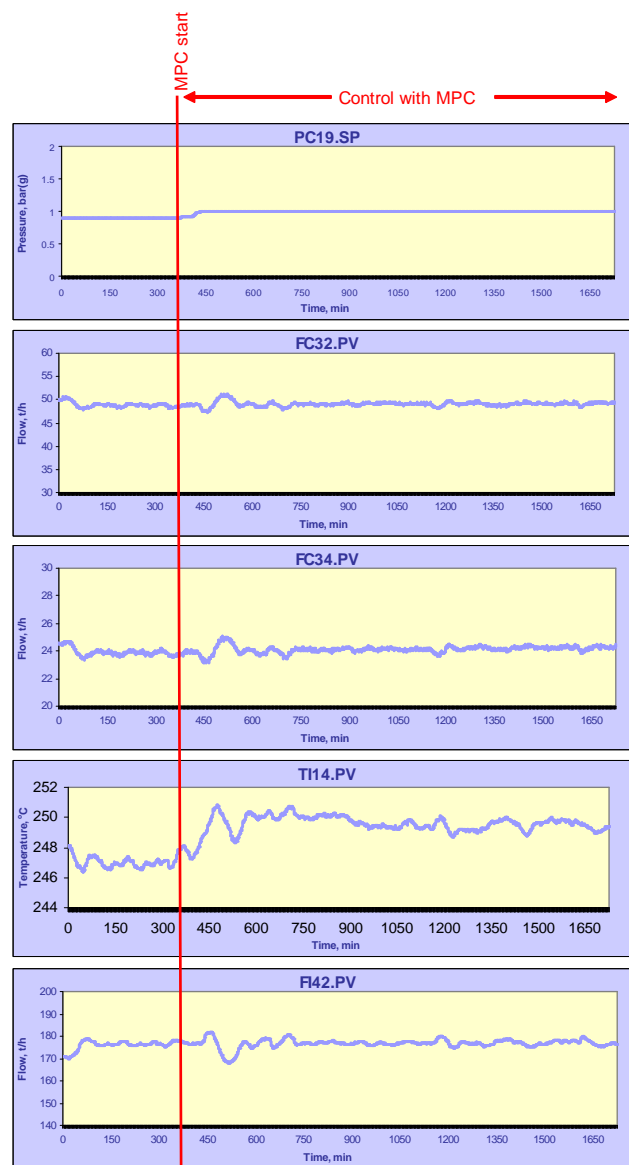
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	MV5 FC18.SP	MV6 FC3.SP	MV7 FC27.SP	MV8 PC19.SP
CV1 TI14.PV	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = .144 Settle T = 0 TfSettle = 32.2 FIR Form = UK 	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = .0514 Settle T = 0 TfSettle = 31.9 FIR Form = UK 	-	-
	DV1 FC32.PV	DV2 FC34.PV		
CV1 TI14.PV	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -.0139 Settle T = 0 TfSettle = 40.1 FIR Form = UK 	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -.0139 Settle T = 0 TfSettle = 40.1 FIR Form = UK 		
	MV1 FC5_12.SP	MV2 TC5.SP	MV3 TC7.SP	MV4 FC14.SP
CV2 FI42.PV	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = .510 Settle T = 0 TfSettle = 126 FIR Form = UK 	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -.460 Settle T = 0 TfSettle = 110 FIR Form = UK 	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -1.30 Settle T = 0 TfSettle = 64.8 FIR Form = UK 	-
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CV1 TI14.PV	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = .160 Settle T = 0 TfSettle = 30.0 FIR Form = UK 	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = -.120 Settle T = 0 TfSettle = 153 FIR Form = UK 	Trial 1 Lap Order 1 FIR Order 0 Stable Dead T = 0 Gain = -.140 Settle T = 0 TfSettle = 53.3 FIR Form = UK 	Trial 1 Lap Order 2 FIR Order 0 Stable Dead T = 0 Gain = 10.8 Settle T = 0 TfSettle = 19.7 FIR Form = UK 
	DV1 FC32.PV	DV2 FC34.PV		
CV2 FI42.PV	-	-		

Figure 3: Multivariable controller control action in preheat maximization





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UDK	Ključne riječi:	Key words:
665.6.011	analiza i optimizacija procesa	process analysis and optimization
66.012.7	više veličinsko optimalno prediktivno vođenje RMPCT	multivariable optimal predictive control RMPCT
665.622.048.3	atmosferska destilacija	topping
66.041.1	predgrijavanje	foreheating
66.012.3	utrošak energije u procesu	process energy consumption

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