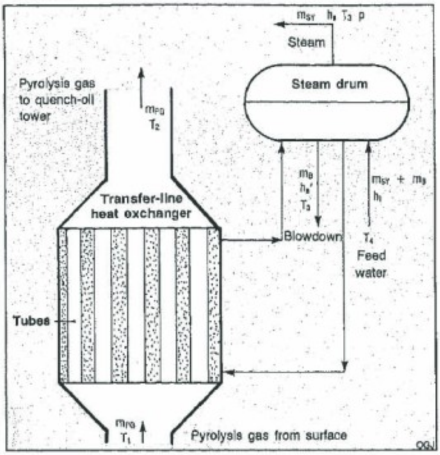
Method improves pyrolysis TLE operation

Jan Barton
Chemopetrol
Labem, Czechoslovakia

Fig. 1

Transfer-line heat exchanger



An objective cost function was derived for optimizing the duration of an operating cycle of the transfer line heat exchanger, or waste-heat boiler, on ethylene furnaces for generation of high-pressure steam.

The objective function was developed to minimize both the cost of cleaning and the heat loss associated with the higher pyrolysis gas exit temperatures required by coke deposition on the heat transfer area of the transfer line heat exchanger.

The objective function is based on the amount of heat transfer surface area of the exchanger and the cost to clean the exchanger because these are the main parameters affecting the energy capacity and heat losses in transfer line heat exchangers (TLE's).

The TLE, using the heat of the pyrolysis gas leaving the pyrolysis tube reactor at roughly 800° C. for generating high-pressure steam at 12 MPa, represents an important part of an ethylene unit. TLE's are conventionally designed as tubular heat exchangers, where the pyrolysis gas from the pyrolysis tube reactor flows inside the tubes and the water boils in the shell side at 12 MPa.

Heat remaining in the pyrolysis gas, after the TLE, is quenched by oil.

The oil is used to generate low-pressure steam at 0.7 MPa that is mixed with the hydrocarbon feed entering the furnace.

The tube bundle of the TLE is usually situated vertically, closely connected to the pyrolysis reactor (Fig. 1).

Basis of theory

Assume that for clean heat transfer areas of the TLE: $K = K(i)$ and $T_2 = T_2(i)$, initially. After a period of operation, τ, the following relationships apply: $_\tau K = K(\tau) < K(i)$ and $T_2(\tau) > T_2(i)$. See nomenclature box for definitions of variables and symbols. The heat loss is defined as the heat that cannot be used to generate steam, and is assumed to be associated with the heat lost through heat exchanger blowdown.

The heat transfer in the TLE is defined by:

$$Q_L = m_{PG}C_p[T_2(\tau) - T_2(i)] \quad (1)$$

It is assumed that a fraction of the heat, Q_L, which is not utilized for generating high-pressure steam in the exchanger due to the deposition of coke, is characterized by the parameter, $P\epsilon(0,1)$, where 0 represents clean conditions and 1 represents coked conditions.

The cost per day, C_1, resulting from the formation of coke and, hence, from the lower rate of high-pressure steam generation in the transfer line heat exchanger can be written as:

$$C_1 = [\sum_{i=1}^{\tau} \{(T_1 - T_3)[\exp(-K(0)S/m_{PG}C_p) - \exp(-K(1)S/m_{PG}C_p)] \times C_p m_{PG}\, 86{,}400(W_1 - W_2) \times 10^{-6}\}]\tau \quad (2)$$

If the feed of a pyrolysis furnace that is out of service due to transfer line heat echanger cleaning is sent to another furnace, and thus no losses arise in the overall rate of the pyrolysis plant production, the cost per day

Fig. 2

Reduction of heat transfer coefficient

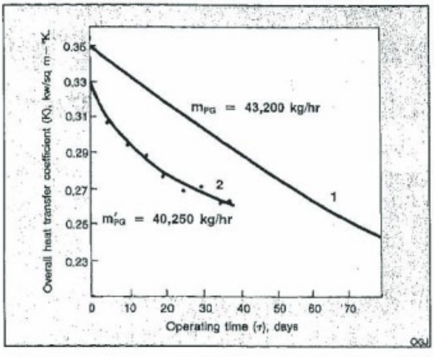
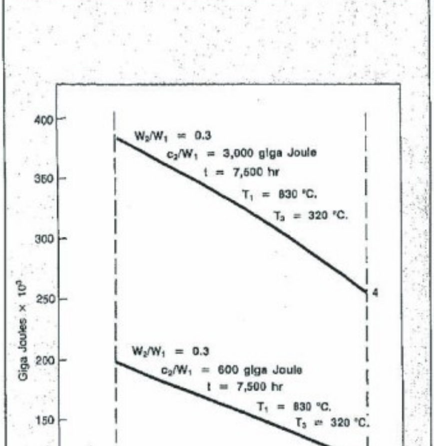


Fig. 3

Exchanger combined costs*



* linear function of surface area

Nomenclature

B	=	Blowdown subscript
b, d, and e	=	Parameters from Table 2
B_0, B_1, B_2, B_3	=	Parameters from Equation 9
C	=	Objective cost function defined by Equation 4
C_1	=	Cost per day defined by Equation 2, monetary units/day
C_2	=	Cost per day of cleaning, monetary units/day
C_y	=	Cost of cleaning & heat losses over service life, monetary units
c_2	=	Cost of cleaning, monetary units
C_p	=	Mean specific heat capacity, kJ/kg-°K.
D_0, D_1, D_2, D_3	=	Parameters of Equation 10
E_0, E_1, E_2, E_3	=	Parameters of Equation 11
h	=	Enthalpy, kJ/kg
K	=	Overall heat transfer coefficient, kw/sq m-°K.
m	=	Flow rate, kg/hr or kg/sec
P	=	Parameters of Equation 2
PG	=	Pyrolysis gas
p	=	Pressure, MPa
Q	=	Exchanger capacity, kw
Q_L	=	Heat loss defined by Equation 1, kw
r	=	Correlation coefficient of multiple linear regression
S	=	Heat transfer area, sq m
SY	=	High-pressure steam
T	=	Temperature, °C. or °K.
t	=	Available time of operation, hr/year
t_L	=	Total service life of waste heat exchanger, years
W_1	=	Price of high-pressure steam, monetary units/giga Joule
W_2	=	Price of process steam, monetary units/giga Joule
W_3	=	Price of heat transfer area, monetary units/sq m
x	=	Mass fraction
τ	=	Time of operation, days

of cleaning is determined by:

$$C_2 = c_2/\tau \qquad (3)$$

The form of the objective function is:

$$C = C_1 + C_2 \qquad (4)$$

and it holds that the rate of change of the cost with respect to the operation is zero, or:

$$dC/d\tau = 0, \text{ for: } C = C_{min}, \text{ and } \tau = \tau_{opt} \qquad (5)$$

Blowdown rate

The transfer line exchanger blowdown rate can be calculated from an enthalpy balance by the following equation:

$$m_B = [m_{PG}C_p(T_1 - T_2) - m_{SY}(h_e - h_i)]/(h_e - h_i) \qquad (6)$$

where h_e is the exit steam enthalpy and h_i is the entering water enthalpy.

The accuracy of the value of m_B determined by Equation 6 can be assessed with the aid of the KOMAT program in which the error-spreading theory is used to calculate a nonmeasured quantity's error on the basis of errors recorded in measured quantities.[1-3]

The error of the calculation becomes tolerable if, in any case, the maximum relative error of the measured quantities did not exceed ±0.5% relative.

This, however, is very difficult to achieve in terms of industrial practice.

Considerably more accurate results of the determination of the transfer line heat exchanger blowdown flow rate could be arrived at only by measuring the blowdown stream directly. In the case that the temperature, T_4, of the water fed into the transfer line heat exchanger approaches T_3 in the exchanger, the heat losses are then concentrated in the preheating of the feed water to the transfer line heat exchanger.

Coke formation

Coke formation is an undesirable chemical reaction taking place in the course of the pyrolysis of all hydrocarbons. Due to the high value of the activation energy of coking during the pyrolysis of naphtha, coke formation is heaviest at the inlet of the transfer line heat exchanger tubes, on surfaces of highest temperature.

Line 2 of Fig. 2 shows the dependence of the overall heat transfer coefficient, K, on the overall operating time of the transfer line heat exchanger.

Considering fluctuations in the pyrolysis gas flow rate, the calculated value of the heat transfer coefficient was corrected according to:

$$K' = Km_{PG}'/m_{PG} \qquad (7)$$

where the quantities with the apostrophe correspond to $m_{PG}' = 40{,}250$ kg/hr, chosen as the reference flow rate.

Line 1 in Fig. 2 shows the model dependence of K on τ for $m_{PG} = 43{,}200$ kg/hr (12 kg/sec).

It holds for the time dependence of K according to Line 1 that:

$$K = 0.35 - 1.5 \times 10^{-3}\tau + 2 \times 10^{6}\tau^{2} \qquad (8)$$

and it is assumed that K is independent of S, T_1, and T_3.

Optimization

The calculations of the optimum operating time, τ_{opt}, of the transfer line heat exchanger are based on the following assumptions:

- The flow rate of pyrolysis gas through the transfer line heat exchanger is constant ($m_{PG} = 12$ kg/sec).
- The heat transfer area is within the range of 131-331 sq m.
- The pyrolysis temperature, T_1, lies within the range 1,080-1,120° K. (301-341° C.).
- The parameter, P = 0.5, or 50% of the Q_L is used for generating process steam.
- The price ratio, W_2/W_1 = 0.3 to 1.0.
- The cost ratio, C_2/W_1, corresponds to 600-6,000 giga Joules.

Calculation algorithm

Multiple linear regression yielded the following dependencies:

$$C_{min} = B_0 + B_1S + B_2T_1 + B_3T_3/W_1 \qquad (9)$$

$$\tau_{opt} = D_0 + D_1S + D_2T_1 + D_3T_3 \qquad (10)$$

$$\Delta T = E_0 + E_1S + E_2T_1 + E_3T_3 \qquad (11)$$

where the temperatures, T_1 and T_3, are expressed in degrees K.

Fig. 4

Surface area vs. outlet gas temp.

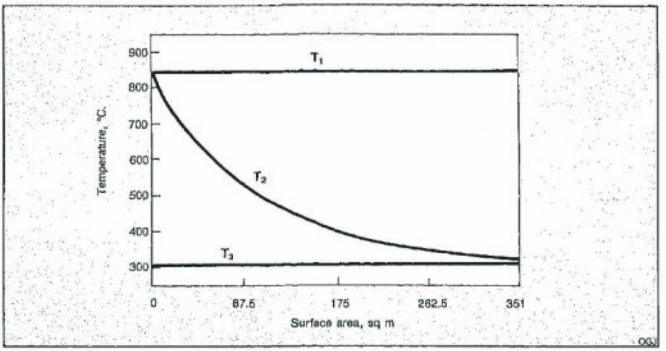


Table 1

Values of parameters B, D, and E

W_2/W_1	c_2/W_1	B_0	B_1	B_2	B_3	r
0.3	3,000	100.1988	−0.2285	0.1094	−0.1069	0.9981
1.0	3,000	74.4728	−0.1603	0.0808	−0.0789	0.9976
0.3	6,000	135.2149	−0.2272	0.1485	−0.1441	0.9975
0.3	600	49.8627	−0.1336	0.0592	−0.0572	0.9988
		D_0	D_1	D_2	D_3	r
0.3	3,000	66.0139	0.0966	−5.1972E-02	4.9466E-02	0.9891
1.0	3,000	86.8721	0.1041	−5.6999E-02	6.6111E-02	0.9834
0.3	6,000	102.7135	0.1055	−7.4520E-02	6.2502E-02	0.9813
0.3	800	28.4177	0.0601	−2.6059E-02	2.5701E-02	0.9806
		E_0	E_1	E_2	E_3	r
0.3	3,000	43.4614	−0.0876	4.8861E-02	−6.2497E-02	0.9864
1.0	3,000	50.4999	−0.1084	4.6226E-02	−4.1706E-02	0.9873
0.3	6,000	55.9326	−0.1111	5.4648E-02	−5.7383E-02	0.9961
0.3	600	9.8039	−0.0496	3.2164E-02	−2.4464E-02	0.9904

Table 2

Values of parameters $b_{i,j}$, $d_{i,j}$, and $e_{i,j}$

i,j	b	d	e
0,0	52.55	11.64	2.889
0,1	−24.99	34.14	22.52
0,2	1.56E-02	1.37E-02	8.36
r	= 0.9871	0.9977	0.9542
1,0	−0.157	5.33E-02	−3.57E-02
1,1	4.68E-02	2.63E-02	−3.70E-02
1,2	−1.66E-05	8.26E-06	−1.12E-05
r	= 0.8524	0.0336	0.9835
2,0	6.24E-02	−1.96E-02	3.14E-02
2,1	−3.08E-02	−2.56E-02	2.59E-03
2,2	1.64E-05	−8.91E-06	4.07E-06
r	0.9917	0.9954	0.9438
3,0	−6.059E-02	−1.51E-02	−3.18E-02
3,1	2.95E-02	3.08E-02	7.50E-03
3,2	−1.59E-05	6.71E-06	−5.77E-06
r	= 0.9903	0.9616	0.7669

The values of parameters B, D, and E are given in Table 1.

The dependencies of these parameters on the terms, W_2/W_1 and C_2/W_1, were approximated by linear relations of the type:

$$B_0 = b_{0,0} + b_{0,1}W_2/W_1 + b_{0,2}c_2/W_1 \qquad (12)$$

$$B_1 = b_{1,0} + b_{1,1}W_2/W_1 + b_{1,2}c_2/W_1, \text{ etc.} \qquad (13)$$

The value of the parameters, b, d, and e, are presented in Table 2. They were also obtained by multiple linear regression of data from Table 1.

The effect of heat transfer area S is on the combined costs of exchanger cleaning

The author. . .

Barton

Jan Barton is currently manager of the central laboratory of the Research Institute of Utility Glass in Novy Bor, Czechoslovakia, where he has specialized in energy conservation, naphtha pyrolysis modeling, and systems-engineering methods since 1989.

From 1980 to 1989, Barton was with Chemopetrol Research Institute of Inorganic Chemistry in Labem, Czechoslovakia, where he served as a member of the systems-engineering group, and where he prepared this article. He has a PhD in organic technology from the Prague Institute of Chemical Technology.

and heat loss, Q_L.

The sum of the costs of cleaning the exchanger and heat losses over the whole service life, t_L, of the exchanger may be written:

$$C_y = (t/24\tau_{opt})C_{min}\tau_{opt}\tau_L \quad (14)$$

where the time required for cleaning the exchanger is excluded from the assumed yearly operating time.

The values of the Parameters B and D in Table 2 indicate that a change in S, T_1, and T_3, that brings about a decrease in C_{min} results, at the same time, in an increase in τ_{opt} and a decrese in delta T.

Fig. 3 shows the dependence of the transfer line heat exchanger comparative cost value as a linear function of S on conditions that the relative cost of 1 sq m of the heat transfer area is W_3/W_1 = 100 giga Joules/sq m (Curve 2), and W_3/W_1 = 300 giga Joules/sq m (Curve 3).

The effect of S ON C_y/W_1 at t_L = 10 years, t = 7,500 hr, W_2/W_1 = 0.3, c_2/W_1 = 3,000 giga Joules, T_1 = 830° C., and T_3 = 320° C., is demonstrated by Curve 4.

Curve 5, also illustrating the dependence of C_y/W_1 vs. S, was constructed with all conditions identical as for Curve 4, except for c_2/W_1 = 600 giga Joules. It is possible to estimate from Fig. 3, the recovery period for the investment cost of installing transfer line heat exchangers with a larger heat transfer area on assumption that W_3/W_1 = 100 giga Joules/sq m, i.e., approximately 2 to 5 years.

Fig. 4 shows the effect of S on the temperature of T_2, at the constant number of tubes having a constant internal diameter, when K = 0.36 kw/sq m − °K., T_1 = 847° C. (1,120 °K.), T_3 = 301° C. (574° K.), and m_{PG} = 12 kg/sec.

Recommendations

To reduce the undesirable losses of energy caused by excessive blowdown from the transfer line heat exchanger, which may be caused by leaks for instance, it is necessary to directly measure the blowndown flow rate.

When revamping a transfer line heat exchanger, it is advantageous to consider a larger heat transfer area (by as much as 50%) than is currently used in ethylene plants to obtain the temperature difference between the pyrolysis gas to the transfer line heat exchanger outlet and the point in the exchanger where steam is at roughly 25° C., taking the clean heat transfer area into account (if 12 MPa high-pressure steam is generated).

To reduce the costs of cleaning the heat transfer area in the transfer line heat exchanger, decoking of the pyrolysis tube reactor and transfer line heat exchanger should be done simultaneously. Reducing the cost of cleaning to one fifth of the original value enables the heat loss caused by coking in transfer line heat exchangers to be decreased by as much as 50%.

References

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Pipeline inspectior

D. L. Atherton
C. Jagadish
P. Laursen
V. Storm
F. Ham
B. Scharfenberger
Queen's University
Kingston, Ont.

Detailed measurements show that pipeline in-line inspection-tool speed can cause significant reduction in defect-induced magnetic flux leakage (MFL) signals.

This reduction must be taken into account in any attempt to measure defect penetration to 10% accuracy as opposed to simply detecting anomalies.

It is generally known that the defect-induced signals obtained from MFL tools used for in-line inspection of oil and gas pipelines are sensitive to the velocity of the inspection tool.

It is often thought that this is due to the use of simple inductive sensing coils in which, because the induced voltage is proportional to the rate of change of flux, signals should increase with increasing velocity.

Other experts in this field have reported that with proper design the effect can be virtually eliminated, and they have supported this claim with simple examples of metal-loss defect signals.[1]

Our experience with magnetic levitation systems[2,3] suggests that, for typical pipeline MFL tools, the time constant for the magnetic flux to diffuse through the pipe wall could be comparable with the transit time. Consequently we expected significant changes in the anomalous MFL patterns induced by defects, particularly for far-side corrosion pits.

We have therefore undertaken precision experimental measurements of defect-induced magnetic leakage flux distributions generated with an MFL anomaly detector representative of the advanced inspection tools used for pipeline monitoring.

Given here are some initial results from detailed maps of defect-induced MFL patterns made at different relative tool velocities for various defect

Fig. 1

MFL, 75% far-side pit, 15 mph