CHEMICAL REACTOR DESIGN FOR PROCESS PLANTS

Volume Two: Case Studies and Design Data

HOWARD F. RASE

W. A. Cunningham Professor of Chemical Engineering
The University of Texas at Austin

Original Illustrations by

JAMES R. HOLMES

Associate Professor of Engineering Graphics
The University of Texas at Austin

A WILEY-INTERSCIENCE PUBLICATION

JOHN WILEY & SONS, New York • London • Sydney • Toronto

CASE STUDY 103

Quench Cooler

THIS STUDY illustrates a quench design problem that is applicable to any very fast reaction in which reaction continues during the quenching process. In this particular case the reaction must be quenched as rapidly as possible, while at the same time minimizing pressure drop, since the ultimate destination of the effluent is a compressor suction.

Problem Statement

Design a quench cooler for quenching the effluent from two coils of the ethane cracking furnace of Case 102B.

Chemistry and Kinetics

The chemistry and kinetics are already documented in Case Study 102. For this quench cooler the homogeneous reaction continues to occur in the gases discharging from the furnace. If the model accurately described the coke-forming reaction both homogeneous and heterogeneous, it would be more valuable in the design of the quench cooler, for it would then provide a means for determining on stream time for the cooler by monitoring the changes in temperature profile and pressure as coke is deposited on the tube walls.

The model of Case 102B will be used. Although it will correctly predict the formation of coke at the inlet end of the exchanger for ethane cracking effluent, the quantitative data are not sufficiently accurate to use seriously in any operating time study. The heat effects, however, are sufficiently accurate for describing the declining reaction as required for determining heat transfer area.

Models for quench cooler fouling exist but details have not been presented (1).

Design Basis

For design purposes the exit conditions for Case 102B will be used with the total amount flowing equivalent to that from two 5-in coils. Outlet temperature should be safely above the dew point of the mixture which depends primarily upon the small amounts of heavy polymeric products not normally included in the reaction model. In ethane cracking the amount of 400°F plus material is so small that the dew point is very low, and the outlet temperature and thus the steam temperature is based solely on the steam economics in the particular plant (2). The decision would also be based in part on the steam turbines to be selected. Higher pressure steam produces better turbine efficiency (lower water rates). For illustrative purposes we select 1500 psia because of the number of large compressors in the operating area. If the main use for this system would be smaller turbines, 900 psia would be a better pressure level at this point in time since turbines at this level are cheaper and are produced by most manufacturers (5). The following design parameters will be used.

Inlet Conditions. 1581°F @ 31.2 psia. 1581 → 596

Composition. See Table CS-2.8 Case 102B, p. 31. 20 ft long

Steam Pressure. 1500 psia @ 596°F (prior to use, steam will be superheated.)

Double-Pipe Tube Properties. Inside tube: 0.97 in. ID and 1.25 in. OD. Outside tube and materials of construction: to be specified by manufacturer.

Maximum Allowable ΔP. $\approx$ 2 psi.

Maximum Tube Length. Based on given limitations of the correlation for the film coefficient a maximum L/D of approximately 240 will be set. This limits the tube length to approximately 20 ft.

Design Equations

The same heat and material balance and pressure-drop equations as those given for Case 102B are used. In Eq. 1.50 the heat-flux term becomes

$$q_O = U(T - T_j) \qquad \text{(CS-3.1)}$$

where U is the overall film coefficient, T is the gas temperature, °F, and T_j is the boiling water temperature, °F. For the double-pipe design, shown in Fig. 10.10,

$$U = \frac{D_O}{Dh_i} + \frac{D_O b_w}{D_m \lambda_w} + \frac{1}{h_o} \tag{CS-3.2}$$

where h_i is the gas film coefficient and h_o is the boiling water coefficient taken as 2040 BTU/(hr)(ft²)(°F).

The high flux encountered in such exchangers creates steep temperature gradients and distorted velocity profiles that demand special correlations (3). The following equation for h_i has been recommended based on a thorough review of the literature (4).

$$\left(\frac{h_i D}{\lambda_f}\right)_Z = \frac{0.021(N_{Re})_Z^{0.8}(N_{Pr})_Z^{0.4}}{(T_w/T)^{0.29+0.0019Z/D}} \tag{CS-3.3}$$

where T and T_w are the bulk fluid and the wall temperatures, respectively, and sub Z refers to the value of the parameter at Z distance from entrance.

This correlation gives results within 10% for most high temperature data (3, 4) in the following ranges:

$$10 < \frac{L}{D} < 240$$

$$200 < T < 2800°\text{R}$$

$$1.1 < \frac{T_w}{T} < 8.0$$

Design Procedure

Various values of mass velocity will be tried and results will be compared on the basis of heat recovered and pressure drop. Calculations are made for one tube, and then the exchanger is designed so that the total flow will be distributed equally among all tubes.

Results

In Table CS-3.1 the results of several calculations are given. It is clear that the case for $G_s = 12$ meets the design requirements. Although the additional conversion in the exchanger tube is small, it does increase with lower mass velocities. Conversion at the low average temperature of the exchanger does produce undesired side reaction products including some which are not very precisely documented by this model. It would seem wise, therefore, to use the maximum allowable ΔP and minimize thereby the extent of conversion.

Table CS-3.1 Comparison of Design Cases Quench Cooler

Mass Velocity lb/ft² sec	ΔP, psi	Outlet Temp. °F	Conversion
10	1.48	670	2.59×10^{-3}
12	2.1	679	2.59×10^{-3}
13	2.44	683	2.17×10^{-3}

residency time

From Table CS-2.7, p. 29, the total output for two coils is

$$(8586)(2)(1.25) = 21465 \text{ lb/hr}$$

Number of exchanger tubes = $21465/[(3600)(12)(\pi/4)(0.97/12)^2] = 96.8$ or 97 tubes.

REFERENCES

1. J. Chen and M. J. Maddock, *Hydrocarbon Process.*, **52** (5), 147 (1971).
2. J. P. Fanaritis and H. J. Streich, *Chem. Eng.* (*N.Y.*), p. 80 (May 28, 1973).
3. D. W. Sundstrom and R. L. DeMichell, *Ind. Eng. Chem. Process Des. & Develop.*, **10**, 114 (1971).
4. W. R. Gambil, *Chem. Eng.* (*N.Y.*), **74** (18), 147 (1967).
5. J. F. Farrow, *Hydrocarbon Process.*, **50** (3), 71 (1971).

TO ≈ 1"

$A_I = \frac{\pi D^2}{4} = \frac{3}{4} \text{ in}^2 \cdot \frac{\text{ft}^2}{144 \text{ in}^2}$

one tube = .005 ft²

.06 lb/sec per tube

Re # ?