Available online at http://www.journalcra.com

International Journal of Current Research
Vol. 8, Issue, 01, pp.25209-25212, January, 2016

INTERNATIONAL JOURNAL OF CURRENT RESEARCH

ISSN: 0975-833X

RESEARCH ARTICLE

INNOVATIVE CONCEPT IN DESIGNING OF MULTI-TUBULAR FIXED BED CATALYTIC REACTOR

*,1Shilpa Patel and 2Thakore, S. B.

1Government Polytechnic Gandhinagar
2L.D. College of Engineering

ARTICLE INFO

Article History:
Received 14th October, 2015
Received in revised form
19th November, 2015
Accepted 28th December, 2015
Published online 31st January, 2016

Key words:
Benzene & Ethylene.

ABSTRACT

If we use beds in series type fixed bed reactor, we need to take reactants/(Product +Reactants) out and exchange heat in heat exchanger & then to take back to the reactor & sent them to the next bed of catalyst. Here, we need to use heat exchanger for which capital & operating cost is required during designing of reactor for manufacturing of ethyl benzene from benzene & ethylene, we decided to use multi-tubular reactor in place of fixed beds in series. So, that capital & operating cost can be saved for heat exchanger. When we did calculations length of tubes required was much more than the height of the bed of catalyst. So, for better heat transfer we distributed total length of the bed of catalyst into three beds of catalyst in tubes only.

Citation: **Shilpa Patel and Thakore, S. B. 2016.** "Innovative concept in designing of multi-tubular fixed bed catalytic reactor", *International Journal of Current Research,* 8, (01), 25209-25212.

INTRODUCTION

It is well known that Shell and tube type reactor structure provides higher heat transfer co-efficient compared to the Jacketted tower type structure for chemical reaction. If heat transfer area is small in multi-tubular fixed bed catalytic reactor than reactants are taken outside where heat transfer is carried out in heat exchanger. This increases pressure drop and capital cost also increases. Alternative arrangement is possible which is explained here with example. Here, In multi-tubular fixed bed reactor reaction between benzene and Ethylene is carried out for the production of Ethylbenzene. Reaction is highly exothermic and steam is produced by the liberated heat

Solved problem

Reference for data to design reactor for Ethylbenzene is V.G. Jensai and G.V. Jeffreys mathematical methods in Chemical Engineering, 2nd Edition. Academic Press 1977 (Page 424)
Reactor type :MT (Multitabular)
Reactor Phase:Gas
Catalyst:Metal Oxides
Temp:600 – 650^{0}C

**Corresponding author: Shilpa Patel,*
Government Polytechnic Gandhinagar.

Pressure (atm):1 atm
Resistance time or0.2 Second
Space velocity:7500 GHSV (gas) (space velocity hourly)

Space velocity=7500 m/hr
$= 2.08 \frac{m}{Sec}$

100 Kmol /hr Ethylene C_2H_4=100 x 28 = $2800 \frac{kg}{hr.}$

125 Kmol /hr Benzene C_6H_6=100 x 80 = $8000 \frac{kg}{hr.}$

Mixture of reactants $= 10800 \frac{kg}{hr.}$

Avg. molecular weight= $\sum m_i y_i$
=0.444 x 28 + 0.555 x 80
=56.872

Density of gas mixture= $\frac{PM}{RT}$

$$= \frac{1 \times 56.872}{673} \times \frac{T_s(273)}{P_S V_S(1 \times 22.4)}$$
=1.029 Kg/m^3

Volumetric flow rate of reactants

$=10800\ kg/hr \div 1.029\ kg/m^3$

$= 2.91\ m^3/Sec$

Area required of reactor =

$$\frac{\text{Volumetric flowrate of reactans}\left(m^3/Sec\right)}{\text{Velocity of reactants}\ \left(m^3/Sec\right)}$$

$$= \frac{2.9}{2.08}$$

Cross sectional area of tubes in multi-tubular reactor = $1.394\ m^2$

Select 50 mm NB tubes of 14 BWG (3.76 mm) thickness
MOC of tubes = SS 316
Tube OD=50.8 mm
Tube ID=43.28 mm

Total numbers of tubes required

$$nt = \frac{\text{Total area of tubes}}{\text{Area of a Single tube}} = \frac{1.394}{(\Pi/4)(0.04328)^2}$$

Height of bed of catalyst in tubes

$$L = \frac{\text{Net Volume of Catalyst}}{nt.(\Pi/4)di^2}$$

Net volume of Catalyst = Resistance time x Volumetric flow rate of feed

$$= 2.91\frac{m^3}{Sec} \times 0.2\ Sec$$

$$=0.532\ m^3$$

$$\therefore \text{ Height of bed of Catalyst} = \frac{0.532}{1323\times(\Pi/4)(0.04328)^2}$$

$$=0.25\ m$$

Area of tubes $= N_t\ \Pi\ do\ L$
$=1.323 \times \Pi \times 0.0508 \times 0.25$
$=52.484\ m^2$

Heat duty, $\phi t = \Delta H_R \times \frac{Kmol}{hr}$ of Ethylene
=6920.18 KW

Calculation of fixed bed side firm coefficient *hi*
$G=0.147\ Kg/(m^2.S)$

K=0.04 W/ (M.k)

$\mu=0.015\ mpa.S = 0.015 \times 10^{-3}\ pa.S$

$$\frac{hi\ dp}{K} = 3.6\left(\frac{dp}{\mu}\frac{G}{C}\right)^{0.365}$$

$$= \frac{hi\times 0.002565}{0.04} = 3.6\left[\frac{0.002565\times 0.047}{0.015\times 10^{-3}\times 0.3}\right]^{0.365}$$

$\therefore hi = 161.61\ w/m^2\ {}^0C$

hi varies from system to system therefore the given equation of *hi* is not reliable for each system. So, *hi* should be found from pilot plant for real design of reactor. Sketch is provided to understand the reactor.

Shell side heat transfer coefficient *ho* :

Steam is produced at shell side as reaction is exothermic and utilized for production of styrene for energy integrity

To calculate shell side boiling coefficient *ho,* Mostinki equation is used. Nucleate boiling is considered here

$$ho = hnB = 0.104\ P_C^{0.69}\left[\frac{\phi B}{A}\right]^{0.7}\left(1.8\left[\frac{p}{P_C}\right]^{0.17} 4\left[\frac{p}{P_C}\right]^{1.2} + 10\left[\frac{p}{P_C}\right]^{10}\right)$$

$$\left[\frac{\phi B}{A}\right] = \frac{6920.18KW}{52.484m^2} = \text{Heat Flux of reboiler in w/m}^2$$

$$= 131813.4\ w/m^2$$

P_C = for water 22.076 MPa
= 220.76 bar

p = Operating pr. of reboiler. Temp of steam produced is taken as 180^0C. At that temperature pressure is 1 MPa = 10 bar

$h_0 = 0.104 \times (220.76)^{0.69} \times (3247.4)^{0.7}$

$$\left[1.8\left(\frac{10}{220.76}\right)^{017} + 4\left(\frac{10}{220.76}\right)^{1.2} + 10\left(\frac{10}{220.76}\right)^{10}\right]$$

$$=12962.84\ W/m^2\ {}^0C$$

Now Overall heat transfer co-efficient Uo :

$$\frac{1}{Uo} = \frac{1}{ho} + \frac{1}{hod} + \frac{do\ln(do/di)}{2KW} + \frac{do\ l}{di\ hid} + \frac{do\ l}{di\ hid}$$

$hod = hid = 5000w/(m^2\ {}^0C)$

$$\frac{1}{Uo} = \frac{1}{12962} + \frac{1}{500} + \frac{0.0502\ln(50.8/43.28)}{2\times 16} + \frac{50.8}{43.28}\frac{1}{5000} + \frac{50.8}{43.28}\times\frac{1}{161.6}$$

$$\frac{1}{Uo} = 0.7\times 10^{-4} + 2\times 10^{-4} + 2.5\times 10^{-4} + 2.317\times 10 + 0.00726$$

$$\frac{1}{Uo} = 0.00741$$

$$\therefore Uo = 134.95$$

Here, value of hi is less there far, Uo (overall heat transfer co-efficient) is less. So, the actual area (hence the length of the tubes) required for heat transfer is more than height of the bed of catalyst for reaction. Here, length of the tubes for heat transfer is calculated

$$\text{Area required} = \frac{\phi t}{Uo \Delta T_{\ln}}$$

Here, Temperature reactor = 400^0C

Temperature of Steam = 180^0C

$$= \frac{6920.18 \times 10^3}{134.95 \times (400 - 180)}$$

$= 160m^2$

$$\text{Length of tubes required} = \frac{160}{n_t \, \Pi \, Do}$$

$$= \frac{160}{1323 \times \Pi \times 0.0508}$$

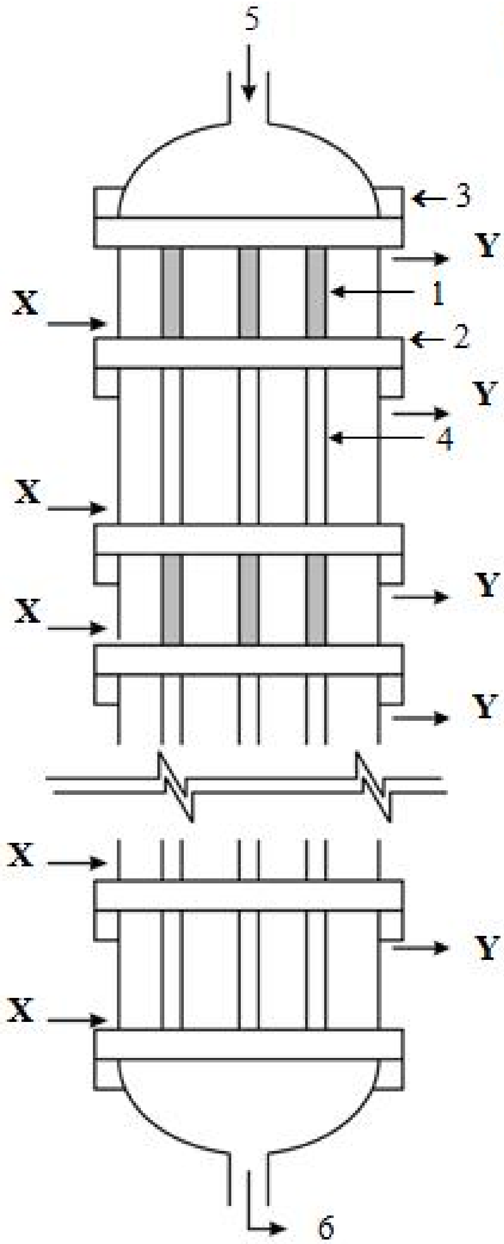
= 7.55 m

$\therefore$ Length of the tube must be kept empthy

=7.55 – 0.25 = 7.30 m

Total length of tube in which catalyst is = 0.25m which should divided into three section each of around 0.08 m

$\therefore$ The way of the tube is 0.08 m catalyst + 2.433 m empty
+ 0.08 m Catalyst + 2.433 m empty
+ 0.08 m Catalyst + 2.433 m empty



SKETCH OF MULTI-TUBULAR FIXED BED REACTOR WITH EMPTY TUBES

1. Bed of Catalyst in tubes
2. Mesh for support of Catalyst which is sandwiched between the flange joints
3. Tube sheet
4. Empty tubes
5. Reactants inlet
6. Product mixture outlet

X Nozzle for D.M. water

Y Nozzle for saturated steam outlet

Conclusion

Here, bed of catalyst is divided into three sections in place of single bed of catalyst. Total length of the tube required for heat transfer area is more than height of the bed of catalyst. Empty space is there in between beds of catalyst which offers heat transfer area. So, we need not take reactants outside the reactor and separate heat exchanger is also not required. So, capital cost is less here. In this design pressure drop is also less.