SURVEY OF LONG TIME BEHAVIOR AND COSTS OF INDUSTRIAL FLUIDIZED BED HEAT EXCHANGERS

R. Rautenbach and T. Katz
Institut für Verfahrenstechnik der RWTH Aachen, Turmstraße 46, Aachen, Germany

Keywords : Fluidized Bed, Heat Transfer, Commercial Units, Erosion, fluidized bed

Contents

Summary

In many cases of severe fouling and/or scaling of conventional heat exchangers fluidized bed heat exchangers can be expected to operate continuously with clean tube walls. Particularly advantageous are circulating fluidized beds due to their wide range of process stability. However, internal particle circulation is inevitably connected to backmixing of the fluid and, as a consequence, losses of mean driving force. The losses are negligibly small in cases of forced convection evaporators but have to be considered in cases of countercurrent liquid-liquid heat exchangers. Tube erosion by the fluidized particles is surprisingly low as long as the heat exchangers are operated within the range of design conditions. The paper discusses the main design criteria and, even more important, costs and long-term operational experiences of 15 industrial fluidized bed heat exchangers, some of them logging more than 40 000 hours of operation.

1. Introduction

The technology of fluidized bed heat exchange has been developed during the last 30 years in the USA, The Netherlands, and Germany (Allen and Grimmett 1978; Hatch and Weth 1970; Klaren, 1975, 1983; Kollbach 1987; Erdmann 1993; Rautenbach, Erdmann and Kollbach 1991b). The experiences gained with this technology can be summarized as follows.

(a) In most applications fouling/scaling of the heat transfer surfaces could be prevented.
(b) Heat transfer coefficients at low superficial velocities ($v \leq 0.5$ m s^{-1}) are excellent.
(c) Pressure losses at the recommended superficial velocity of ~0.5 m s^{-1} are comparable to conventional forced convection tube bundle heat exchangers since conventional heat exchangers are operating at the required velocity of about 2 m s^{-1}.
(d) Erosion of the tubes is negligible.

2. Design

The application of the fluidized bed technique is presently limited to vertically placed tube bundle heat exchangers. The scaling/fouling liquid and, as a consequence, the fluidized particles are flowing in the tubes. All our heat exchangers are operated with a circulating fluidized bed with about 90 per cent of the tubes in upward flow operation and 10 per cent acting as downcomers. A design program is available at the Institut für Verfahrenstechnik of the RWTH Aachen.

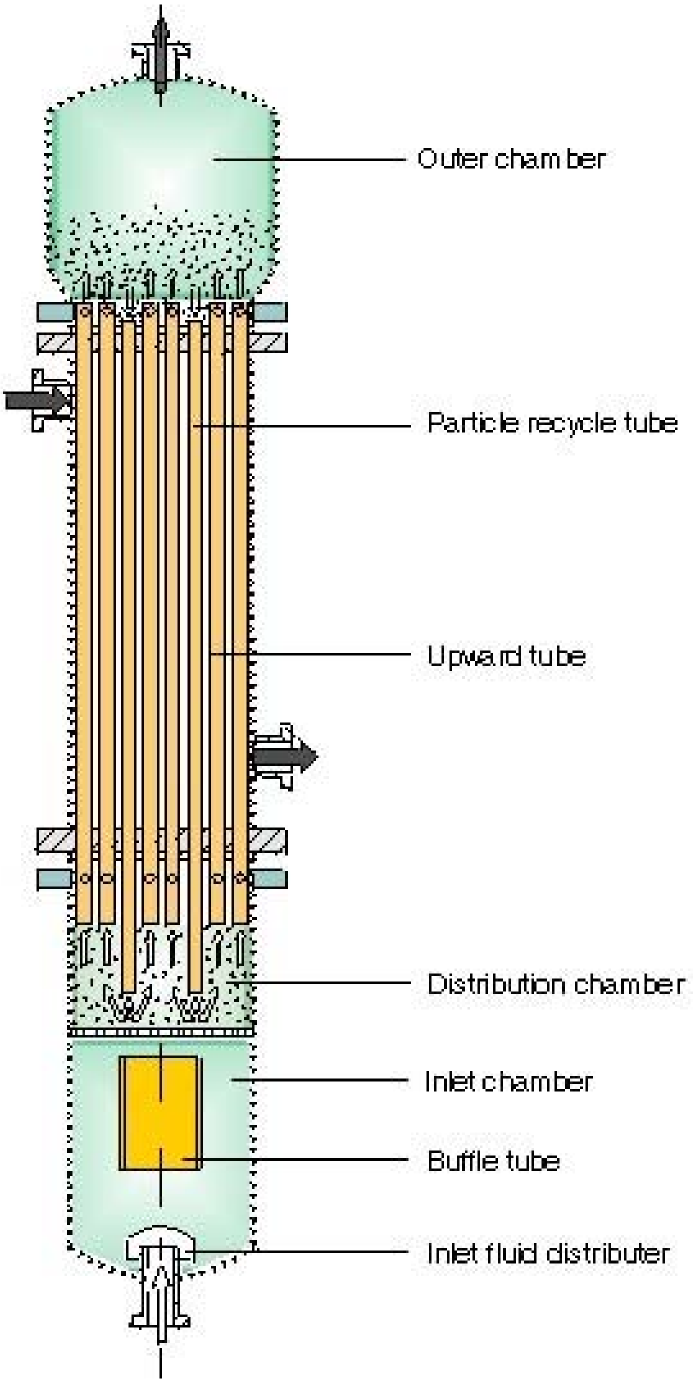

Figure 1. The fluidized bed heat exchanger and its major elements (circulating fluidized bed).

Figure 1 shows the relatively simple design which was a major aim of our development. According to Figure 1, almost any single pass shell and tube heat exchanger can be converted into a fluidized bed heat exchanger by a few and quite simple additions: tube exit nozzles, a large outlet chamber for the separation of the particles, and tube extensions at the lower end - short extensions for the upward flow and longer ones for the downcomers.

The short baffle tube above the fluid inlet should be particularly noticed. It induces a lift effect for the particles, resulting in a significant reduction of the required amount of particles.

Figure 2 shows the tube inlet and outlet elements and the sieve plate required for an even distribution of the particles to the tube bundle. Furthermore, it shows the situation in the tubes during start-up and in steady-state operation.

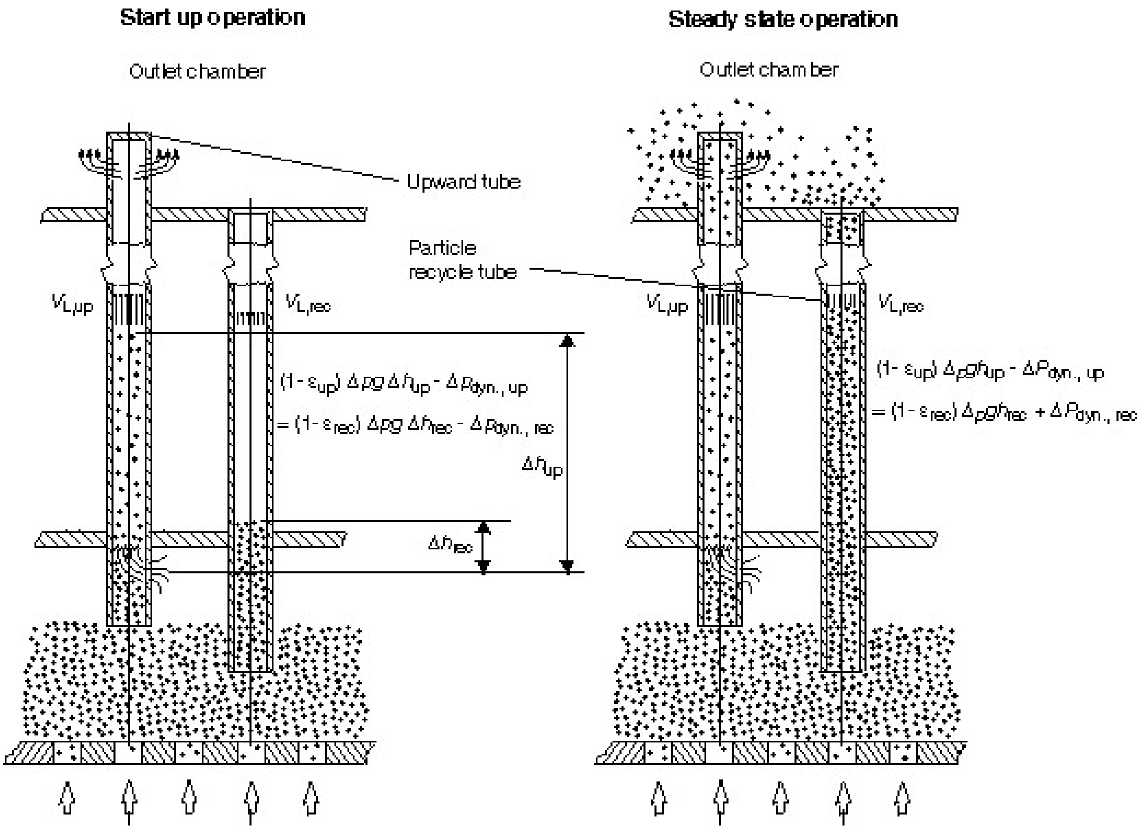

Figure 2. Start-up and steady-state operation of a circulating fluidized bed heat exchanger.

3. Heat Transfer of a Fluidized Bed

The increase of heat transfer between fluid and tube walls by fluidized particles is well-known (Richardson and Mitson 1958l; Wehrmann and Mersmann 1981; Jamialahmadi, Malayeri and Müller-Steinhagen 1995). Figure 3 shows the results of experiments with particles of different material compared with the well-known heat transfer in tubes for single phase flow.

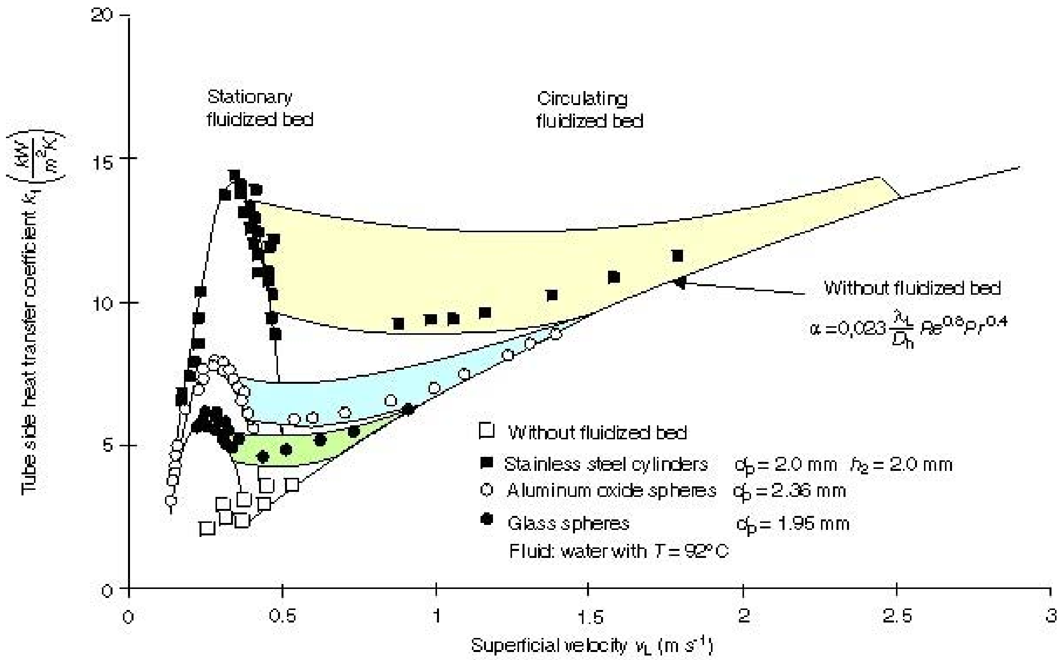

Figure 3. Influence of superficial velocity and particle material on tube side heat transfer. Comparison with single phase flow.

The overall heat transfer accounts for the following three resistances: resistance of the shell side k_2 (tube outside), resistance of the tube material, and resistance of the inner tube side k_1 (tube inside).

The reciprocal value of the heat transfer coefficient is given by

$$\frac{1}{k} = \left[\frac{1}{k_1} + R_{F,1}\right]\frac{d_2}{d_1} + \frac{d_2 \ln(d_2/d_1)}{2\lambda_{\text{tube}}} + \frac{1}{k_2} + R_{F,2}$$

In cases of a heat exchanger heated with condensing steam, the shell side heat transfer coefficient k_2 and the heat transfer in the tube material are well-defined. However, the tube side heat transfer coefficient k_1 depends on the mode of operation.

3.1. Without a Fluidized Bed (Laminar/Turbulent Pipe Flow)

Gnielinski (1975, 1989) and Dittus and Boelter (1930) proposed the following correlations for the calculation of the tube side heat transfer coefficient k_1:

$$2300 < Re < 10^6 \quad Nu = \frac{k_1 d_1}{\lambda} = \frac{\xi/8(Re-1000)Pr}{1+12.7\sqrt{\xi/8}(Pr^{2/3}-1)}\left[1+\left(\frac{d_i}{L}\right)^{2/3}\right]\left(\frac{Pr}{Pr_W}\right)^{0.11}$$

with $\xi = (1.82\log_{10} Re - 1.64)^{-2}$

$$600 < Re < 7000 \quad Nu = 0.023\, Pr^{0.4} Re^{0.8}\left(\frac{Pr}{Pr_W}\right)^{0.11}$$

$$Re < 600 \; Nu = \left[3.66^3 + 0.7^3 + \left(1.615 \sqrt[3]{Re \; Pr \left(\frac{d}{L} \right)} - 0.7 \right)^3 \right]^{1/3} \left(\frac{Pr}{Pr_W} \right)^{0.11}$$

3.2. Circulating Fluidized Bed in the Tubes

A regression analysis of more then 300 experiments resulted in the following correlation:

$$Nu_P = \frac{k_1 d_P}{\lambda} = 2.7 \, Pr^{0.33} Re^{0.75} (1-\varepsilon)^{0.12} \left(\frac{\Delta\rho}{\rho_1} \right)^{0.26} \left(\frac{d_P}{d_1} \right)^{0.20} Fr^{-0.35} \left(\frac{Pr}{Pr_W} \right)^{0.11}$$

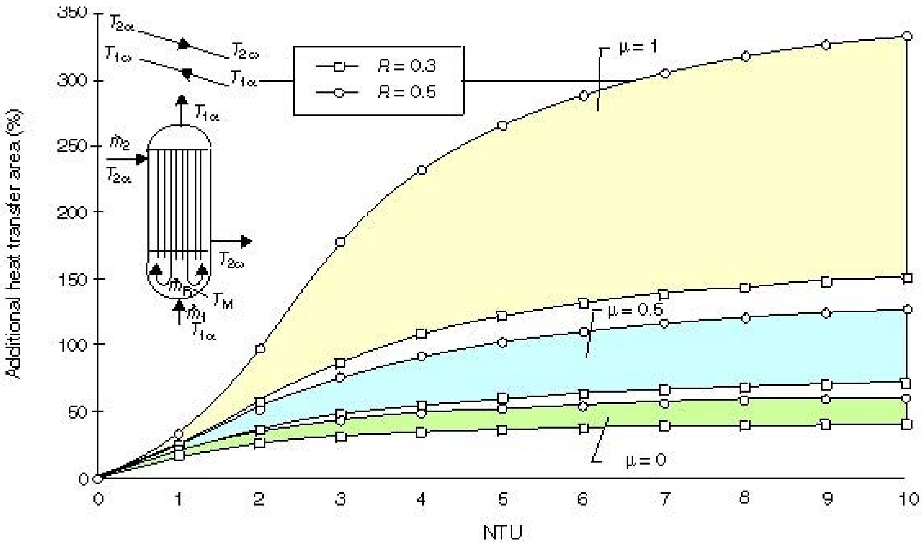

Figure 4. Influence of backmixing on heat transfer surface area (reference: zero backmixing).

The circulating fluidized bed has the advantage of very stable operation conditions in a wide range of flow rates but unfortunately a negative consequence regarding the driving force. As a consequence of the internal circulating the logarithmic mean temperature difference decreases and, consequently, the heat transfer area must be increased compared to a heat exchanger without backmixing. Figure 4 shows the influence of the recirculation rate *R*, the heat capacity ratio *my*, and the dimensionless axial temperature increase (or NTU) on the heat transfer surface increase (reference: zero backmixing).

$$\mathrm{NTU} = \frac{k\,A}{(\dot{m}\,c_P)_1} = \frac{\Delta T_{ax}}{\Delta \vartheta_{ln}} \quad R = \frac{w_{RF}}{w_{WAT}} = \frac{\dot{m}_{1,RF}\, c_{p,1} + \dot{m}_{P,RF}\, c_{p,P}}{\dot{m}_{WAT}\, c_{p,1}} \quad u = \frac{(\dot{m}\,c_P)_1}{(\dot{m}\,c_P)_2}$$

Normally, the recirculating rate is between 0.3 and 0.5. According to Figure 4 the effect of backmixing is negligible in cases of steam heated forced convection heat exchangers ($u = 0$) as for example installed in multiple-effect evaporators. However, in cases of liquid/liquid countercurrent a significantly required heat transfer surface area is

Bibliography and Suggestions for further study

Allen C A and Grimmett E S (1978) *Liquid-fluidized-bed Heat Exchanger Design Parameters*. Department of Energy, Idaho Operations Office.

Dittus F W and Boelter L M (1930) Heat transfer in automobile radiators of the tubular pipe. *Univ. of California Public. in Eng.* 2.

Erdmann C (1993) *Wärmeaustauscher mit zirkulierender Wirbelschicht zur Verhinderung von Belagbildung*. Düsseldorf: VDI-Verlag GmbH.

Gnielinski V (1975) Neue Gleichungen für den Wärme- und Stoffübergang in turbulent durchströmten Rohren und Kanälen. *Forsch. Ing.-Wes.* 41(1), 8-16.

Gnielinski V (1989) Zur Wärmeübertragung bei laminarer Rohrströmung und konstanter Wandtemperatur. *Chem.-Ing.-Tech.* 61, 160-161.

Hatch L P and Weth, G G (1970) Scale Control in High Temperature Distillation Utilizing Fluidized Bed Heat Exchangers, Research and Development Progress Rep. No. 571.

Jamialahmadi M, Malayeri M R and Müller-Steinhagen H (1995) Prediction of heat transfer to liquid-solid fluidized beds. *Can. Journal of Chem. Engn.* 73, 444-455.

Katz T (1995) *Bestimmung der internen Zirkulationsrate eines Wirbelschicht-Wärmeaustauschers IVT Information 25*. Institut für Verfahrenstechnik der RWTH Aachen.

Klaren D G (1975) *Development of a Vertical Flash Evaporation*. Ph. D. Thesis, Delft University of Technology, The Netherlands

Klaren D G (1983) The fluidized bed heat exchanger: principles and modes of operation and heat transfer results under severe fouling conditions. *Fouling Prev. Res. Dig.* 5(1).

Klaren D G (1995) *Verbesserte Version des selbstreinigenden Wirbelschicht-Wärmeaustauschers für stark verschmutzende Flüssigkeiten*. Ludwigshafen, Deutschland: Vortrag der Klarex Technology BV, Nijkerk, Niederlande bei der BASF.

Kollbach J S (1987) *Entwicklung eines Verdampfungsverfahrens mit Wirbelschicht-Wärmeaustauscher zum Eindampfen krustenbildender Abwässer*. Dissertation, RWTH Aachen.

Rautenbach R and Katz T (1995) *WSWAT-Design 2.2. Program and Manual*. Institut für

Verfahrenstechnik, RWTH Aachen.

Rautenbach R, Erdmann C and Kollbach J S (1991a) Selbstreinigender Wirbelschicht-Wärmeaustauscher - hydraulische und thermische Apparatecharakteristik, Betriebserfahrungen. *Chem.-Ing.-Tech.* 63, 1005-1007.

Rautenbach R, Erdmann C and Kollbach J S (1991b) The fluidized bed technique in the evaporation of wastewaters with severe fouling/scaling potential - latest developments, applications, limitations. *Desalination* 81, 285-298.

REFA (1991) *Methodenlehre der Betriebsorganisation, Plan und Gestaltung. 1. Aufl*. München.

Richardson J F and Mitson A E (1958) Sedimentation and fluidisation part II - heat transfer from a tube wall to a liquid-fluidized bed. *Trans. Instn. Chem. Engrs.* 36, 270-282.

VDI-Wärmeatlas (1994) *7. erw. Auflage*. VDI - VERLAG Düsseldorf.

Wehrmann M and Mersmann A (1981) Wärmeübertragung in Flüssigkeit durchströmten Fest- und Fließbetten. *Chem.-Ing.-Tech.* 53, 804.