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Review on Heat Transfer in Gas Solid Fluidized Beds

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Abstract. This review covers those aspects of heat transfer in fluidized beds related to bed/immersed tubes and bed/vessel walls heat transfer coefficients correlations. These correlations generally incorporated the effects of operating conditions, bed and gas thermophysical properties and bed size and geometry. Experimental as well as empirical dimensionless models are presented. The general tendency in the literature is towards building up more realistic hydrodynamic models which incorporate particle and gas convective heat transfer components, solids residence time and local bubble behaviour at immersed surfaces. It is also generally recognized in the literature that heat transfer cannot be satisfactorily explained on the basis of stationery models but rather on the basis of unsteady state heat conduction models of solids movement induced by rising bubbles. The aspects of research mostly overlooked in the surveyed investigations were those related to interphase heat transfer and the effect of operation under elevated temperatures and pressures.

Nomenclature

B	a consant.
C	specific heat, cal/gm °C.
C_{pg}	specific heat of gas, cal/gm °C.
C_{ps}	specific heat of solids, cal/gm °C.
D_B	the diameter of a sphere having the same bubble volume, cm.
d_p	particle diameter, cm.
d_t	bed or tube diameter, cm.
ε_f	void fraction in a bubbling bed, dimensionless.
ε_{mf}	void fraction in the bed at minimum fluidizing velocity, dimentionless.
h_w	heat transfer coefficient between bed and surface, cal/sec. cm^2 °C
k_g	thermal conducitivity, cal/cm. sec. °C.

L_f	height of bubbling fluidized bed, cm.
L_{mf}	the bed height at minimum fluidizing conditions, cm.
U_A	the absolute rise velocity of bubbles, cm/sec.
U_o	superficial fluid velocity (measured on an empty tube basis) through a bed of solids, cm/sec.
U_{mf}	the superficial gas velocity at minimum fluidization, cm/sec.
ϱ_g	fluid density gm/cm^2.
ϱ_s	density of solids, gm/cm^3.
μ	viscosity of gas, gm/cm. sec.
τ_o	the time of contact between the emulsion phase and surface, sec.
λ_g	the gas thermal conductivity, cal/sec. cm. °C.
λ_p	the particle thermal conducitivity, cal/sec. cm. °C.

Introduction

By far the greatest share of research effort on heat transfer studies in fluidized beds has been devoted to the study of small particle systems [1-11]. This significance of fine particles (as opposed to larger ones) is due to the heavy reliance of the chemical industry on them in fluidized processes to enhance high rates of heat and mass transfer or high rates of solid transport [12].

The pioneering work of Kunii and Levenspiel [13] represented one of the early systematic presentations of the topic. They presented gas/solid contacting in fluidized beds by the rather simple description of what they termed "The bubbling bed model." In their clasic book "Fluidization Engineering" [13], the earlier experimental findings on heat transfer between fluidized beds and surfaces were presented. An attempt to explain these findings in the light of the then available knowledge (and using the bubbling bed model) was attempted. Different correlations describing heat transfer at the walls of containing vessels were presented [14-17] and these included the Dow and Jacob correlation

$$\frac{h_w d_p}{k_g} = 0.55 \left(\frac{d_p \varrho_g U_o}{\mu}\right)^{0.8} \left(\frac{\varrho_s C_{ps}}{\varrho_g C_{pg}}\right)^{0.25} \left(\frac{d_t}{d_p}\right)^{0.03} \left(\frac{L_f}{d_t}\right)^{-0.65} \left(\frac{1-\varepsilon_f}{\varepsilon_f}\right)^{0.25} \tag{1}$$

The Van Heerden *et al.* correlation [16]:

$$h_w \frac{d_p}{k_g} = 0.58 \left(C_{pg} \frac{\mu}{k_g} \right)^{0.5} \left(B \frac{d_p \varrho_g u_o}{\mu} \right)^{0.45} \left(\frac{C_{ps}}{C_{pg}} \right)^{0.36} \left(\frac{\varrho_s (1 - \varepsilon_{mf})}{\varrho_g} \right)^{0.18} \tag{2}$$

And the Levenspiel and Walton [15] correlation:

$$\frac{h_w d_p}{k_g} = 0.6 \left(\frac{C_{pg} \mu}{k_g} \right) \left(\frac{d_p \varrho_g u_o}{\mu} \right)^{0.3} \tag{3}$$

The correlation presented by Wen and Leva [16] correlated the data of four groups of investigators with the following general equation which covers a wide range of materials:

$$\frac{h_w d_p}{k_g} = 0.16 \left(\frac{C_{pg} \mu}{k_g} \right)^{0.4} \left(\frac{d_p \varrho_g \mu_o}{\mu} \right)^{0.76} \left(\frac{\varrho_s C_{ps}}{\varrho_g C_{pg}} \right)^{0.4} \left(\frac{u_o^2}{g d_p} \right)^{0.2} \left(\frac{\eta L_{mf}}{L_f} \right)^{0.36} \tag{4}$$

Where:

$$\eta = u_o - \frac{\text{(superficial velocity for uniform expansion of bed)}}{u_o} \tag{5}$$

The superficial velocity for uniform expansion of bed refers to the superficial fluid velocity at minimum fluidizing conditions (U_{mf}, cm/sec.).

L_f is the bed height at minimum fluidizing conditions. The other symbols were already defined in the preceding sections. Amshtein and Gepperin [18] made a good analysis of a comprehensive set of correlations that predict the heat transfer coefficients beween fluidized beds and container walls.

Heat Transfer Between Fluidized Beds and Immersed Tubes

A number of experimental studies have dealt with heat transfer between fluidized beds and immersed tubes. Wender and Cooper [19] correlated the data for vertical tubes reported by several investigators [16, 20-22] with the expression:

$$\frac{h_w d_p}{k_g} = 0.01844\, C_R\, (1-\varepsilon_f) \left(\frac{C_{pg}\varrho_g}{k_g} \right)^{0.43} \left(\frac{d_p \varrho_g u_o}{\mu} \right)^{0.23} \left(\frac{C_{ps}}{C_{pg}} \right)^{0.8} \left(\frac{\varrho_s}{\varrho_g} \right)^{0.66} \tag{6}$$

for

$$\frac{d_p \varrho_g u_o}{\mu} = 10^{-2} - 10^{2}$$

Wender and Cooper [19] found the above general correlation gave a deviation of ± 20% for the set of 323 data points used in the correlation.

The study of heat removal rates by boiler tubes immersed in fluidized beds, was another area of research which received considerable attention over the years [23-27]. Interest in this area was mostly focused on dependence of h_w for vertical tubes on operating conditions, the effects of distributor design and surface roughness.

Vreedenberg [28] investigated the effect of tube diameter, particle size, shape, density and gas velocity on the heat transfer coefficient on immersed horizontal tubes. His recommended correlations were:

$$\frac{h_w h_{ti}}{k_g} = 0.66 \left(\frac{C_{pg}\mu}{k_g} \right)^{0.3} \left(\frac{d_{ti} \varrho_g u_o}{\mu} \right) \left(\frac{\varrho_s}{\varrho_g} \right) \left(\frac{1-\varepsilon_f}{\varepsilon_f} \right)^{0.44} \tag{7}$$

for

$$\frac{d_{ti} \varrho_g u_o}{\mu} < 2000$$

and

$$\frac{h_w h_{ti}}{k_g} = 420 \left(\frac{C_{pg}\mu}{k_g} \right)^{0.3} \left[\left(\frac{d_{ti} \varrho_g u_o}{\mu} \right) \left(\frac{\varrho_s}{\varrho_g} \right) \left(\frac{\mu^2}{d_p^3 \varrho_s g} \right) \right]^{0.3} \tag{8}$$

for

$$\frac{d_{ti} \varrho_g u_o}{\mu} > 2500$$

Where d_{ti} is the outer diameter of the horizontal immersed tube.

Catipovic *et al.* [29] claimed that correlations based on experimentally based deterministic models have more predictive power than empirical dimensionless expressions. The correlation put forward by the authors was based on this principle and was in fact a hydrodynamic correlation which predicts heat transfer to horizontal tubes in a fluidized bed:

$$1 - \beta = 0.45 + \frac{0.061}{(U - U_{mf}) + 0.125} \tag{9}$$

Where:

- β is the time fraction that the tube is in contact with bubbles.
- U is the superficial gas velocity.
- U_{mf} is the superficial gas velocity at minimum fluidization.

The investigators [29] claimed that the above hydrodynamic model gave reliable predictions of heat transfer coefficients for a wide range of particle sizes, gas velocities, different total pressures and quite different physical properties.

Assuming a bubble velocity equal to U_{mf} the following heat transfer coefficient was predicted by equation (9) above:

$$\frac{hd_p}{\lambda_g} = 6(1-\beta) + (0.0175\, A_r^{0.46}\, P_r^{0.33})(1-\beta) + \frac{d_p}{D}\,(0.88\, Re_{mf}^{0.5} + 0.0042 Re_{mf})^{0.33}\,(P_r)\,\beta \tag{10}$$

Catipovic postulated that the correlation of Baskakov and Suprun [30] for vertical tubes;

$$\frac{h_{gc}\, d_p}{\lambda_g} = 0.0175\, A_r^{0.46}\, P_r^{0.33} \tag{11}$$

gave an excellent fit to the available h_{mf} data for horizontal tubes.

The 1982 PhD work referred to earlier [12], proposed (on the basis of experimental data) a theoretical model for the overall convective heat transfer coefficient for such a system. The author postulated that the overall heat tansfer coefficient is a function of "gas" and "particle" heat transfer coefficients, and that, to a first order of approximation, the particle and gas convective components are in fact additive. The parameters examined in that research work were: The effective thermal conductivity of the emulsion phase in the bed, the gas film thickness around the tube, the emulsion phase contact time and replacement frequency and the bubble voidage. These parameters were utilized in examining the differences between the heat transfer characteristics of beds of large and small particle sizes.

The author pointed out that the major difficulty associated with the prediction of the particle heat transfer coefficient using the pocket renewal model approach was the determination of the time of contact between the emulsion phase and the immersed surface. In the study, this contact time was measured experimentally for various particle sizes and fluidization velocities. The author concluded that his technique was reliable and gave consistent results. In addition the author proposed the following semi-empirical equation (based on bubble hydrodynamics) for estimating the emulsion phase residence time:

$$\tau_e = \frac{D_B}{U_A}\left[\frac{3\pi}{8}\left(\frac{U_A}{U-U_{mf}}\right)-1\right] \tag{12}$$

Several studies have been concerned with the measurement of h_w between the fluidized beds and various immersed objects such as spheres, cylinders, wires as well as horizontal and vertical tubes [31-38].

In general experimental investigations of h_w for vertical tubes are relatively more scarce than for horizontal tubes. It is generally believed that h_w values are dependent on tube orientation [25].

Experimental data of h_w for vertical tubes available in the literature which can provide a good basis for establishing a correlation for Nusselt number $N_{u,max}$ corresponding to:

$$N_{u,\,max} = h_{w,\,max}\frac{\bar{d}_p}{k_g}$$

are presented in Table 1.

Table 1. System and material's properties for some heat transfer investigations in fluidized beds [39]

Investigator(s)	Materials	$\overline{d}_{p}, \mu_m$	$\rho_s, K_g/m^3$	$C_{ps}, J/K_gK$	F_t,m
Antonishin [40]	Sand	268	2700	800	14,21,28
Borodulya *et al.* [41]	Sand	800	2480	800	13
Chen & Withers [42]	Glass beeds	127-610	2500	754	12.7,22
Mickley *et al.* [43]	Glass beeds	104-430	2464	754	6.35
Ozkaynak & Chen [44]	Glass beeds	136-610	2470	754	12.7
Wunder [45]	Glass beeds	100 - 770	2500	754	30,40
	Bronze	108, 150	8620	343	40
Mathur *et al.* [25]	Sand	500-803	2623	800	12.7
	Glass beeds	275 - 405	2364	754	28.6 , 50.8

Where d_p is the harmonic mean diameter of bed particles, M, and D_t is the diameter of heat transfer tube immersed in bed particles, M.

Correlations for Nusselt Nuber (max) to Archimedes Number, for vertical tubes immersed if fluidized beds as proposed by different experimental investigators are given in Table 2.

Models Incorporating Bed Hydrodynamics

Mathur *et al.* [25] presented an explanation of the effect of bed hydrodynamics (solids and bubble movement) on the solids residence time at the immersed heat transfer surface. This residence time sensitively controls the value of h_w. The investigators [25] postulated that the bed hydrodynamics itself is dependent on various system parameters such as bed size (diameter and height), distributor design, particle size, internals and on operating conditions such as pressure, temperature and fluidizing velocity.

Despite the large number of detailed models available in the technical literature which describe the means by which surface to bed heat transfer takes place, these models are often of limited practical use because they require a relatively large number of parameters which are not usually readily available in practical situations [48,49]. Furthermore, many of these models are of questionable validity because they do not make adequate allowance for the dynamic behavior of gas fluidized systems. The work of Mathus *et al.* [25] is one of the earlier exceptions of this observation. Bock and Molerus [50] studied the influence of hydrodynamics on local bubble

Table 2. Nusselt number correlations for vertical tubes immersed in fluidized beds

Investigator	Correlation	
Renz *et al.* [28]	$N_{u,max} = 0.88\, A_r^{0.213}$ $150 < A_r < 2 \times 10^5$	(13)
Varygin and Martysushin [46]	$N_{u,max} = 0.86\, A_r^{0.2}$ $30 \leq A_r \leq 1.35 \times 10^5$	(14)
Grewal *et al.* [36]	$N_{u,max} = 0.9\ [A_r \frac{D_t (= 127\,mm)}{D_t}] \left(\frac{C_{ps}}{C_{pg}}\right)^{0.2}$ $75 < A_r \leq 20000$	
Denloye and Botteril [47]	$N_{u,max} = 0.843\, A_r^{0.15} + 0.86\, A_r^{0.39}\ \bar{d}_p^{\,0.5}$ $10^3 < A_r < 2 \times 10^6$	(16)
Mathur *et al.* [30]	$N_{u,max} = 209 \left(A_r \frac{C_{ps}}{C_{pg}}\right)^{0.145} \left(\frac{\bar{d}_p}{D_T}\right)^{0.065}$	(17)

Where A_r is Archimedes Number $= \dfrac{g\bar{d}_p^{\,3}\, \varrho_g\, (\varrho_s - \varrho_g)}{\mu^2}$, dimensionless

behavior and time-averaged local heat transfer at a vertical single tube and a vertical tube bundle in fluidized beds of different diameters. This time averaged local heat transfer coefficient was correlated by the expression:

$$\frac{1}{\alpha} = \frac{1}{1 - \varepsilon_B} \left[\frac{1}{\alpha_{max_p} (1 - \varepsilon_{mf})} + \frac{\sqrt{\pi}}{2} \sqrt{\frac{t}{(\lambda_p C)_E}} \right] \tag{18}$$

Where t is the contact time of the emulsion phase, and is defined by the equation:

$$t = (1 - \varepsilon_B) / f_B$$

and

α is the heat transfer coefficient.

ε_B is the local mean bubble volume traction.

α_{max_p} is defined by:

$$\alpha_{max_p} = (4\lambda_g / d_p) \{ ((2\varrho / d_p) + 1) \ln (d_p/2\varrho) + 1) - 1\} \tag{19}$$

Where

$$\varrho = 2\omega_g (2 - \gamma) / \gamma$$

ω_g is the mean free path of gas molecules

γ is an accommodation coefficient.

Bock and Molerus's correlation gave a good match between experimental and predicted heat transfer coefficients.

Models Incorporating the Effect of Pressure

The effect of pressure on heat transfer and hydrodynamics received relatively little attention in the literature. Ernst [51] reported measurements obtained in a small laboratory scale facility. His results (using catalyst particles in the size range of 20 to 100 μ_m and at pressures of 0.1 and 10 M Pa) showed no measurable pressure effect on heat transfer. The maximum heat transfer coefficient achieved reported from these experiments was of the order of 570 W/m^{2}K.

Row [52] analyzed the work of Sobreiro and Monteiro on the effect of pressure on minimum fluidization velocity. It had been pointed earlier by Şobreiro and Monteiro that U_{mf} varies with gas pressure (density) which is in turn strongly dependent on particle size. This is because $R_{e_{mf}}$ (Reynold's number at minimum fluidization $d_p U_{mf} \varrho_g/\mu$) varies markedly with particle size, and the ratio of form drag to skin friction is much larger with larger particles. An expression for U_{mf} based on Ergun's equation [53] was found to fit data quite well up to pressures of 100 bar [54]. This expression is of the form:

$$U_{mf} = \frac{\mu}{\varrho_g d_p} 42.9 (1-\varepsilon_{mf}) \{ (1 + 3.112 \times 10^{-4} \frac{\varepsilon^3_{mf}}{(1-\varepsilon_{mf})^2} G_a)^{\frac{1}{2}} - 1 \} \tag{20}$$

where

$$G_a = \varrho_g (\varrho_s - \varrho_g) g d_p^3 / \mu^2$$

Since $\varrho_s \gg \varrho_g$ and viscosity is independent of pressure, G_a is linear in pressure (or almost so far permanent gases).

Row published a graphical representation of U_{mf} variations with pressure (see Fig. 1).

From this plot representation (of a power fluidized by nitrogen at room temperature), the marked change of U_{mf} with particle size is clearly seen. When $Re_{mf} < 2$ (*i.e.* $d_p < 100\,\mu$m), U_{mf} is virtually independent of pressure. For large particles where $Re_{mf} > 2$, *i.e.* $d_p > 10$ mm and when $\varrho_s \gg \varrho_g$, Ergun's equation reduces to:

$$U_{mf} = [(\varrho_s - \varrho_g) g \varepsilon^3_{mf} d_p / 1.75 \varrho_g]^{\frac{1}{2}} \tag{21}$$

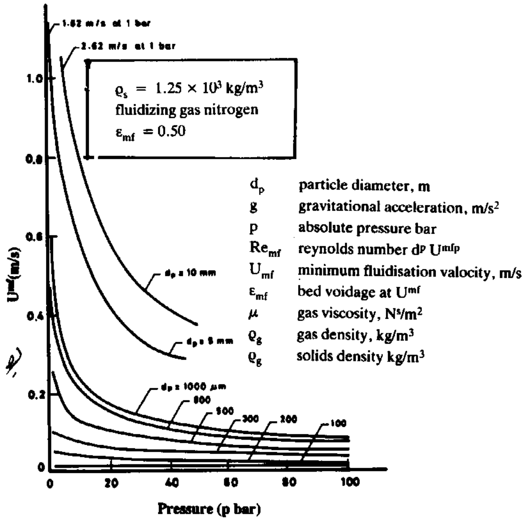

Fig. 1. Variation of U_{mf} with pressure in fluidized beds [52]

Xavier *et al.* [55] investigated heat transfer up to 2.5 M Pa in a small column, 0.1 m in diameter. In contrast to the results of Ernst, the measurements of Xavier *et al.* indicated a marked influence of pressure on the heat transfer coefficient. In fact a change in the maximum heat transfer coefficient from about 200 W/m^{2}K at 0.1 M Pa to 800 W/m^{2}K at 2.5 M Pa was observed.

Borodulya et al [56] reported measurements in a column, 0.105 m in diameter, in a pressure range up to 8.1 M Pa. It was observed that the influence of pressure on the maximum heat transfer coefficient increased with increasing particle size. When pressure was increased from 1.1 to 8.1 M Pa, the maximum heat transfer coefficient increased from 560 to 720 W/m^{2}K for particles of 126 μm, while for particles of d_p = 3100 μm an increase of heat transfer coefficient from 520 to 1280 W/m^{2}K was obtained. These significant effects were explained by the improvement in the bed structure as well as the increased contribution of the gas convective component.

Heat transfer measurements on a horizontal tube bank at pressures up to 1 M Pa were performed by Canada and McLaughlin. The investigators used a squire bed 0.305 m long equipped with a ten row tube bundle. Heat transfer coefficients were measured at five locations, and the fluctuations in the heat transfer coefficients around the arithmetic mean did not exceed ± 10%. The authors found in addition that the bed height exerted no measurable influence on the heat transfer coefficients. The investigators also claimed that, up to fivefold the minimum fluidizing velocity, gas velocity exerted no influence on heat transfer.

Bock and Schweinzer studied heat transfer to horizontal tube banks in a pressurized gas/solid fluidized bed. The investigators used a square bed, 0.4 m long and 1 m in diameter which was installed in a pressure vessel. Heat transfer coefficients were measured at different locations in the tube banks using five electric heating elements. The local bubble flow was determined by using capacitance tubes. It was concluded from the results of this study that an increase in heat transfer resulted from the increase in pressure. The results were particularly marked in the range of large particles. it was also observed that the gas convective component which predominated for larger particles and at high pressures, appeared to be independent of gas velocity in the range of the investigated experimental conditions.

Heat Transfer Mechanisms

One of research areas which attracted a lot of attention was the study of heat transfer mechanisms. Botterill gave a good presentation of this topic in his book "Fluid Bed Heat Transfer" [48]. In their wider context, studies in this area of research, were concerned mainly with experimental investigations and Modelling of gas/solid heat transfer, particle/particle heat transfer and bed/surface heat transfer. In the latter area the limiting laminar layer models [14] and the particle replacement models [57,58] received special attention.

Situations involving radiant heat transfer in fluidized beds were the subject of interest in some recent publications [59,60]. Such cases usually involve particle/particle radiative heat transfer as well as bed to surface radiative heat transfer.

At present it is generally recognized [61] that heat transfer to a fluidized bed cannot be satisfactorily explained on the basis of a film model, *i.e.* a stationery resistance to heat transfer, but more appropriately on the basis of an unsteady state heat conduction model. This heat conduction takes place in those parts of the bed which, following the solids movement, contact the heat transfer surface for some period. The solids movement, which is mainly induced by the rising bubbles, may have a more or less distinct pattern near the wall of the fluidized bed and the pattern will determine the heat transfer coefficient.

The solids movement in a fluidized bed is induced by the rising bubbles [61,62-67]. Bubbles are not merely gas pockets but may have a wake in which solids are carried upwards. Moreover the bubbles are surrounded by a cloud of a dense phase consisting of gas and particles. From the wake, an exchange of heat and mass with the surrounding dense phase takes place continuously. The modelling of solids circulation and its effect on heat transfer coefficients was discussed in some detail by Hoelen and Stemerdding [61].

Heat transfer coefficient between an immersed surface and a fluidized bed consists of three additive components: the particle convecive component, the gas convective component and the radiative component [68,69]. The most significant of these components, in general, is the particle convection [70,71]. For this reason, a relatively large number of heat transfer models have been put forward in the literature for the particle convective component [53,69,72,73]. Martin's model [55] makes use of some basic ideas adopted from the kinetic theory of gases in order to describe the mechanism of energy transfer through the moving particles, and is reported to give a good fit to experimental data [72]. The model of Bock and Molerus [69] describes adequately the dependence of heat transfer on particle diameter as a result of varying bubble flow conditions. However, the interphase heat transfer is generally overlooked in many of the heat transfer studies presented in the literature.

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استعراض عملية انتقال الحرارة المميعة المحتوية على الغازات والمواد الصلبة في المهـد

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ملخص البحث. إن هذا الاستعراض يشمل العلاقات المتبادلة لمعاملات انتقال الحرارة المميعة في المهد بين المهد والأنابيب المغمورة وبين المهد وجدران الوعاء. وقد اهتمت الترابطات المستخدمة لهذا الغرض من ناحية عامة بتأثيرات ظروف التشغيل والخصائص الحرارية الفيزيائية للمهد والغاز بالإضافة إلى حجم وشكل المهد نفسه. وقد استخدمت لهذا الغرض نماذج رياضية مختبرية ووضعية بدون وحدات أبعاد.

ولقد كان الاتجاه العام في الأدبيات نحو بناء نماذج رياضية هيدروديناميكية واقعية تأخذ بعين الاعتبار مكـونـات انتقال الحرارة الحركية ومدة مكوث المواد الصلبة بالإضافة إلى السلوك المحلي للفقاعات عند الأسطح المغمورة. كما أن هناك اتفاقا عاما في الأدبيات إنه ليس من الممكن فهم عملية انتقال الحرارة بصورة مرضية عن طريق النماذج الرياضية بافتراض الطبقات الثابتة ولكن عن طريق نماذج انتقال الحرارة بالاتصال بافتراض حالات غير مستقرة لحركة المواد الصلبة التي تسببها حركة الفقاعات المتصاعدة. أما النـواحي البحثية التي أغفلتهـا الأبحاث التي يشملها هذا الاستعراض بصورة عامة فهي تلك المتعلقة بانتقال الحرارة بين الأطوار وتأثير التشغيل تحت ظروف درجات الحرارة والضغط المرتفعين.