Flow boiling heat transfer of carbon dioxide with polyalkylene glycol–type lubricating oil in pre-dryout region inside horizontal tube

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ABSTRACT

A flow boiling heat transfer model for horizontal tubes is proposed for CO₂ with entrained polyalkylene glycol (PAG oil) in the pre-dryout region. A general power law–type model with a power number of 3 is used together with the average thermodynamic properties of the CO₂–oil mixture. A convective enhancement factor (F) is recommended according to the relationship between the Lockhart–Martinelli parameter and the ratio $\alpha_{tp}/\alpha_l$, which was obtained based on previous experimental results for CO₂ and oil. A new suppression factor (S) is introduced that comprises a suppression term for forced convection and oil concentration term for bubble generation. A comparison of six correlations showed that the proposed correlation can depict the influence of the mass and heat fluxes on both nucleate and convection boiling reasonably well.

Keywords: Carbon dioxide; heat transfer coefficient; correlation; lubricating oil; boiling heat transfer

Nomenclature

$Bo$ boiling number, $Bo = q/(Gh_\text{lv})$
$c_p$ specific heat (J kg⁻¹ k⁻¹)
$E_{\text{mix}}$ two-phase heat transfer multiplier
$F$ convective enhancement factor
$g$ gravitational acceleration (m s⁻²)
$h_\text{lv}$ latent heat of vaporization (J kg⁻¹)
$m$ mass flow rate (kg s⁻¹)
$P_c$ critical pressure (Pa)

$Bd$ bond number, $Bd = g\rho D^2/\sigma$
$D,d$ hydraulic diameter (m)
$EF$ enhancement factor
$Fr$ Froude number, $Fr = v^2/(gd)$
$G$ mass flux (kg m⁻² s⁻¹)
$M$ molecular weight (kg kmol⁻¹)
$n$ total number of data points
$Pr$ Prandtl number $Pr = \mu/(\rho a)$
1. INTRODUCTION

Carbon dioxide (CO₂) was rediscovered by Lorentzen and Pettersen (1993, 1994, 1995) to be a promising alternative refrigerant for heat pump, water heater, and mobile air-conditioning applications because of its favorable properties such as zero ODP (ozone depletion potential), negligible GWP (global warming potential), non-toxicity, and non-flammability. CO₂ also has excellent thermodynamic properties such as low viscosity and surface tension because of its low critical temperature and high operating pressure, which cause it to exhibit unique characteristics in flow and heat transfer. Many previous studies on flow boiling heat transfer of CO₂ (e.g., Pettersen et al., 2002, Tanaka and Hihara, 2000, Dang et al., 2010) showed that nucleate boiling is the dominant mechanism in CO₂ flow boiling because of the high density ratio of vapor to liquid.
In an actual heat pump cycle, the lubricating oil in the compressor is carried to the other components of the system along with the circulating refrigerant. The entrainment of a small amount of lubricating oil in the refrigerant can cause a profound deterioration in the heat transfer performance of the working fluids. Despite the importance of understanding the behavior of oil inside the CO₂ heat pump cycle, only limited information is available regarding the characteristics of flow boiling heat transfer of CO₂ in the presence of oil. Katsuta et al. (2003) measured the boiling heat transfer coefficient of CO₂ in a 3-mm I.D. tube at heat fluxes ranging from 5 to 15 kW m⁻² and mass fluxes ranging from 200 to 600 kg m⁻² s⁻¹. A sharp decrease in the heat transfer coefficient was reported when the oil concentration exceeded 0.3%. Gao and Honda (2004a, 2004b) visually observed the flow pattern of lubricating oil flowing with CO₂ at heat fluxes of 5–40 kW m⁻² and mass fluxes of 230–1200 kg m⁻² s⁻¹. The formation of an oil-rich layer near the heat transfer surface was reported to be the main cause of the nucleate boiling suppression. Dang et al. (2013) presented a systemic experimental study of CO₂ with a polyalkylene glycol (PAG)-type lubricating oil entrained from 0% to 5% in horizontal smooth tubes with inner diameters of 2–6 mm at mass fluxes of 360–1440 kg m⁻² s⁻¹ and heat fluxes of 4.5–36 kW m⁻². Experimental results showed that the presence of oil causes the mass flux to significantly influence the heat transfer coefficient at a low heat flux until dryout, while the mass flux was not observed to have a significant influence at a high heat flux.

Shen and Groll (2003) reviewed numerous methods for predicting the heat transfer of a refrigerant in the presence of lubricating oil. However, available literature on predicting the flow boiling heat transfer of CO₂ and oil is limited. Zhao et al. (2009) attempted to use two methods to predict the flow boiling coefficient: they chose the correlations of Kandlikar (1990), Liu and Winterton (1991), and Kattan et al. (1998) and compared their predictions with the CO₂–lubricant mixture data by replacing the viscosity of pure CO₂ with the viscosity of the mixture. The enhancement factor (EF) proposed by Eckels et al. (1994, 1998), correlation for R134a/369-SUS and two-phase heat transfer multiplier (E_{l,mix}) introduced by Tichy et al. (1986), and correlation for R12/300-SUS were applied in these predictions. Zhao et al. concluded that an overall prediction based on different correlations is inaccurate, especially under low vapor quality and low oil concentration conditions.

This report presents a prediction model of the flow boiling heat transfer of CO₂–PAG oil mixtures that considers the local properties of a mixture based on the previous experimental work of Dang et al. (2013). The proposed correlation and five other available correlations were compared using an experimental database that was set up according to the experimental results of four independent research groups.

2. EXPERIMENTAL APPARATUS AND DATA REDUCTION

The theoretical model is mainly based on previously published experimental results.
(Dang et al., 2013); a brief description is given here to explain how the data were obtained. Figure 1 shows a schematic diagram of the experimental apparatus. The main devices include a magnetic gear pump to circulate the refrigerant and oil, a mass flow meter (±0.5% accuracy), a heater, an oil sampling section, a test section, an oil separator, and a heat exchanger for cooling the refrigerant. The test section was a stainless steel tube heated electrically by a DC power supply. The outside wall temperature of the test section was measured using T-type thermocouples with an uncertainty of ±0.1 °C. Pressures of the working fluid at the inlet and outlet of the test section were monitored using a precision pressure sensor with an uncertainty of ±0.1% F.S. The vapor quality at the inlet of the test section was controlled by a heater. The oil concentration was measured by the oil sampling section and controlled by the opening of a needle valve at the exit of the oil separator. The oil separator was kept at 60 °C to reduce dissolution of CO₂ into the oil. The sampling was usually conducted three times at the beginning, middle, and end of each run. The uncertainty of the oil concentration was estimated to be ±0.1%.

The local heat transfer coefficient $\alpha$ was determined from the measured heat flux $q$ and the temperature difference between the saturation temperature $T_{\text{sat}}$ and inside wall temperature $T_w$.

$$\alpha = \frac{q}{T_w - T_{\text{sat}}}$$

$T_{\text{sat}}$ was estimated from the measured pressure, and $T_w$ was calculated from the measured outside wall temperature using a one-dimensional heat conduction model.

The experimental conditions are summarized in Table 1. The thermodynamic properties of CO₂ were obtained from REFPROP 7.0. The measurement accuracies of some parameters are presented in Appendix 1.

3. THEORETICAL ANALYSIS

3.1 Available correlation for CO₂–oil mixture

Most of the CO₂–oil mixture correlations available in the literature were developed based on the properties of pure refrigerants. Katsuta et al. (2008a), Gao et al. (2008), and Aiyoshizawa et al. (2006) proposed correlations to predict the flow boiling heat transfer coefficient of CO₂ and PAG-type oil based on the properties of pure CO₂. The main characteristics of the above models are introduced below. More details on these models are presented in Appendix 2.

Katsuta et al. (2008a) and Gao et al. (2008) sought to depict the trend of the heat transfer characteristics of the CO₂–oil mixture by revising the boiling suppression factor correlation in the flow boiling heat transfer model for CO₂. Katsuta et al. (2008a) proposed their correlation by introducing a new suppression factor into the Stephan–AbdelSalam correlation (Stephan and AbdelSalam, 1980), which reflects the depression on nucleate boiling caused by the presence of oil.
\[ \alpha_{\text{nb}} = S\alpha_{\text{SA}} \times \phi \quad (2) \]

\[ S = \ln \left( 2.332 \frac{Bo^{0.518} Bd^{1.27} Fr^{0.964}}{Re_{\text{tp}}^{0.834}} \right) \quad (3) \]

where \( \phi \) is defined as

\[ \phi = \frac{1}{1 + OCR^{0.698} Bo^{0.207} Bd^{0.912}} \quad (4) \]

In addition, the factor \( \eta \), which is the function of the oil concentration, is employed to predict the dryout quality in the presence of oil.

\[ x_{\text{dry}} = 0.269 \times \frac{Re^{0.0571} Fr^{0.0697}}{Bo^{0.0519}} \eta \]

\[ OCR < 1 \quad \eta = 1 \]

\[ OCR \geq 1 \quad \eta = 1.169e^{-0.17 \times OCR} \quad (5) \]

The predicted results were reported to agree with 80% of their test results within a prediction deviation of ±20%.

Gao et al. (2008) employed Cheng’s correlations based on their own experimental data. In their analysis, they redefined suppression factor \( S' \) as a function of the vapor quality; the threshold of the vapor fraction was set to 0.7. The correlation was reported to capture 81% of their experimental data within a deviation of ±30%.

\[ S' = \begin{cases} S \times (-0.5x + 0.35) & (x \leq 0.7) \\ 0 & (x > 0.7) \end{cases} \quad (6) \]

Considering the suppression of nucleate boiling by flow boiling of CO\(_2\) in the presence of oil, Aiyoshizawa et al. (2006) adopted the superposition form of Chen and proposed new constants and exponents for convection enhancement and boiling suppression factor equations based on the previous experimental data of Dang et al. (2013). They used the properties of pure CO\(_2\) in their correlation.

\[ F = 0.8 + 0.5 \left( \frac{1}{X_0} \right)^{1.2} \quad (7) \]

\[ S = \frac{1}{6 + \left( Re_{\text{tp}} \times 10^{-4} \right)^3} \quad (8) \]

### 3.2 Proposed correlation

#### 3.2.1 Properties of CO\(_2\)–oil mixture

CO\(_2\) fluid is partly miscible with PAG oil according to Kawaguchi et al. (2000) and Kaneko et al. (2004). Figure 2 shows the two-phase separation temperature of PAG oil with CO\(_2\).
under subcritical conditions (Dang et al. 2012). CO₂ and PAG oil separate under the evaporation temperature from -50 °C to 30 °C when the oil concentration is less than 30%. When the oil separates from liquid CO₂, a certain amount of CO₂ may dissolve into the oil. In this study, the inlet oil concentration of the evaporator was below 5%, and the evaporation temperature was kept at 15 °C. Therefore, three phases exist during the evaporation process: a vapor CO₂ phase, liquid CO₂ phase (with negligible dissolved oil), and an oil layer with dissolved CO₂. Based on the experimental measurements of Youbi-Idrissi et al. (2003) and Garcia et al. (2008), the solubility of CO₂ in PAG oil layer is about 50% when the temperature is 15 °C and pressure is 5.0 MPa.

PAG oil flowing with bulk CO₂ affects both bubble generation on the surface of the tube and the wall superheating temperature. Thermodynamic properties of the oil, such as its viscosity, also influence the flow boiling heat transfer of CO₂. Using just the properties of CO₂ to predict the heat transfer coefficient of the CO₂–oil mixture in the model without considering the contribution of the oil’s properties seems unreasonable. In this study, the thermodynamic properties of the CO₂–oil mixture was introduced into the conventional flow boiling heat transfer correlation in order to predict the refrigerant–oil flow boiling coefficient.

The oil concentration in the refrigeration cycle $\omega_o$ is defined from the ratio of the mass flow rate of oil to the mass flow of the mixture.

$$\omega_o = \frac{m_{oil}}{m_{co} + m_{oil}}$$

(9)

Inside the evaporator, the local oil concentration $\omega_{ol}$ is the function of the vapor quality and increases with the progress of evaporation:

$$\omega_{ol} = \frac{\omega_o}{1-x}$$

(10)

where $x$ is the local vapor quality.

Figure 3 shows that the local oil concentration increases with vapor generation; at an inlet oil concentration $\omega_o$ of 1%, $\omega_{ol}$ may reach 5 % at a vapor quality of 0.8 and increase sharply at higher vapor quality. However, considering the relatively low dryout quality, which changes from 0.4 to 0.8 based on the mass flux, limiting the scope to low local oil concentrations seems a reasonable constraint.

The average thermal physical properties of CO₂–oil were used in this analysis based on the mixing rules shown in Appendix 3 (Eqs. (A-35)–(A-39)). The thermodynamic properties of PAG used in this study were provided by the manufacturer and are presented in Table 2. The data were correlated as a function of the temperature, as shown in Eqs. (A-40)–(A-43). Table 3 compares the properties of PAG oil, CO₂, and the calculated properties of the CO₂–oil mixture using the mixing rules. Even a tiny amount of entrained oil may lead to a dramatic increase in liquid viscosity and surface tension compared to pure CO₂. At a local oil concentration of 1%, the liquid viscosity and surface tension increase 9.9 and 2.1 times,
respectively; when the oil concentration is 5%, these increase 16.9 and 3.6 times, respectively. In contrast, the changes in liquid conductivity, liquid density, and specific heat compared to those of CO₂ are negligible when the oil concentration is small.

The average thermal properties of the mixture defined in this study were applied to calculating the nucleate boiling and force convective heat transfer. For example, the force convection component is given by the following formula:

\[
\alpha_i = 0.023 \text{Re}_m^{0.8} \text{Pr}_m^{0.4} \left( \frac{\lambda_m}{d} \right) \]

where \( \text{Re}_m = \frac{G(1-x)D}{\mu_m} \) and subscript “m” denotes the average properties of the mixture being used.

### 3.2.2 Proposed correlation

Based on the experimental data from Dang et al. (2013), a new correlation is proposed. The correlation of Steiner and Taborek (1992) was used in this study; this is a general power law–type model with a power number of 3. The terms for nucleate boiling and convective heat transfer are shown in Eqs. (13) and (14), respectively.

\[
\alpha_n = \left[ (S\alpha_{nb})^3 + (\alpha_{cb})^3 \right]^{1/3}
\]

\[
\alpha_{nb} = 0.00122 \left( \frac{\lambda_m^{0.79}c_{pm}^{0.45} \mu_m^{0.49}}{\sigma_m^{0.5} \rho_m^{0.29} \rho_v^{0.24} \rho_l^{0.24}} \right) t_{lw}^{0.24} \left( \frac{t_{lw} - t_{sat}(p_l)}{\rho_{sat}(t_{lw}) - p_l} \right)^{0.75} \]

\[
\alpha_{cb} = F \alpha_i
\]

\( F \) and \( S \) are defined as

\[
F = \begin{cases} 
1.0, & X_n \geq 10 \\
2.55(0.413 + 1/X_n)^{1.05}, & X_n < 10 
\end{cases}
\]

\[
S = S_{oil} \frac{1}{0.9 + 0.4 \left( \text{Re}_{qoil} \times 10^{-1} \right)^{0.1} \left( \text{Bo} \times 10^4 \right)^{0.25}}
\]

where \( \text{Re}_{qoil} = \frac{G(1-x)D}{\mu_m} \) and \( S_{oil} \) is introduced to reflect the suppression of nucleate boiling due to the presence of oil:

\[
S_{oil} = \frac{-\omega_{oil}}{2\ln(1-\omega_{oil})}, 0.0005 \leq \omega_{oil} < 1
\]

### 3.2.2.1 Discussion of conventional heat transfer component

In general, flow boiling data can be correlated using the form \( (\alpha_{tp}/\alpha_i) \sim (1/X_n)^n \), where \( \alpha_{tp} \) and \( \alpha_i \) are the flow boiling heat transfer coefficient and heat transfer coefficient, respectively, for liquid flowing alone and \( X_n \) is the Lockhart–Martinelli parameter. Figure 4 shows that the
\( \alpha_{tp}/\alpha_t \) regulations of the CO\(_2\)–oil mixture and pure CO\(_2\) are similar and not linear to \( 1/X_{tt} \) in a logarithmic coordinate system. The data of \( \alpha_{tp} \) were obtained from Dang et al. (2013). When \( 1/X_{tt} \) and the oil concentration are increased, \( \alpha_{tp}/\alpha_t \) of the CO\(_2\)–oil mixture deviates from that of pure CO\(_2\). However, the \( \alpha_{tp}/\alpha_t \) values of CO\(_2\)–oil mixtures with the same oil concentrations under different conditions tend to be identical at high \( 1/X_{tt} \) values, as shown in Fig. 5.

Figure 6 compares the forced convective heat transfer results from using the mixture properties and those of pure CO\(_2\). The liquid-phase forced convective heat transfer of the mixture is much smaller than that of the pure CO\(_2\) due to the reduction in \( \text{Re}_m \) with the addition of oil to the refrigerant. The forced heat transfer of the two-phase flow is also affected by the velocity of the vapor and increases with vapor generation. Chen introduced a two-phase multiplier \( F \) that relates the ratio of the two-phase Reynolds number to that of the liquid Reynolds number: \( F \propto \text{Re}_{tp}/\text{Re}_l \). The Lockhart–Martinelli number is used to calculate \( F \), as shown in Eqs. (18) and (19).

\[
F = \begin{cases} 
1.0, & X_{tt} \geq 10 \\
2.0(0.213 + 1/X_{tt})^{0.756}, & X_{tt} < 10
\end{cases}
\]  

(18)

\[
X_{tt} = \left( \frac{1-x}{x} \right)^{0.9} \left( \frac{\rho_v}{\rho_m} \right)^{0.5} \left( \frac{\mu_m}{\mu_v} \right)^{0.1}
\]  

(19)

Similar to the \( F \) factor proposed by Chen (1966), a new definition for the \( F \) factor is introduced in Eq. (15) based on the experimental data shown in Fig. 5. In addition, \( F \) proposed by Yoshida et al. (1994) is shown in Eq. (20) for comparison.

\[
F = \begin{cases} 
1.0, & X_{tt} \geq 10 \\
1 + 2.0(1/X_{tt})^{0.88}, & X_{tt} < 10
\end{cases}
\]  

(20)

Figure 7 compares the \( F \) factors from four correlations (Chen, 1966, Yoshida et al., 1994, Aiyoshizawa et al., 2006, and the present study). The values of the \( F \) factor proposed in this study as calculated by using the mixture’s properties were similar to those of the correlations by Yoshida et al. and Chen, which were calculated by using the properties of pure CO\(_2\). The \( F \) factor of Aiyoshizawa et al. (2006) was much lower than that of Chen (1966). Aiyoshizawa et al. (2006) apparently tried to decrease the value of \( F \) to reflect the suppression of oil. Thus, \( F \) was less than 1.0 when the vapor quality was very low, which is physically impossible.

3.2.2.2 Discussion of nucleate boiling component

The nucleate boiling component was constructed using Forster and Zuber’s (1955) correlation as modified by Chen (1966), which is shown in Eq. (13).

As noted in Section 3.2.1, three phases exist during the evaporation process: a vapor CO\(_2\) phase, liquid CO\(_2\) phase (with negligible dissolved oil), and oil layer with dissolved CO\(_2\). The
proportion of PAG oil dissolved into CO₂ is usually less than 0.2%; its effect on the change in saturation pressure of CO₂ was negligible in our preliminary experiment. Thus, as long as a separate liquid CO₂ phase exists, the saturated temperature of the CO₂–oil mixture is assumed to be equivalent to that of pure CO₂ and can be determined by the measured local pressure. In addition, the vapor properties used in Eq. (13) are the same as those of pure CO₂. The latent heat of pure CO₂ was used without considering the effect of the oil because the oil concentration was small.

Figure 6 compares the \( \alpha_{nb} \) values calculated by Eq. (13) for both pure CO₂ and the CO₂–oil mixture against the vapor quality. \( \alpha_{nb} \) of the CO₂–oil mixture decreased with increasing vapor quality due to the change in mixture properties caused by increases in the local oil concentration.

Chen (1966) introduced the factor \( S \) to express the suppression of nucleate boiling by convective heat transfer and related it to the two-phase flow Reynolds number as follows:

\[
S = \frac{1}{1 + (2.56 \times 10^{-6})(Re, F^{1.25})^{1.17}}
\]

As shown in Fig. 5, scattering of \( \alpha_{tp}/\alpha_t \) at low \( 1/X_t \) (i.e., low vapor quality) at different heat fluxes showed the contribution of nucleate boiling to heat transfer. At a small mass flux of 360 kg m\(^{-2}\) s\(^{-1}\), \( \alpha_{tp}/\alpha_t \) at low vapor quality increased with the heat flux. In addition, \( \alpha_{tp}/\alpha_t \) took similar values at \( G = 360 \) kg m\(^{-2}\) s\(^{-1}\) and \( q = 9 \) kW m\(^{-2}\) compared to at \( G = 720 \) kg m\(^{-2}\) s\(^{-1}\) and \( q = 18 \) kW m\(^{-2}\). The relative significance of nucleate boiling against convective heat transfer in flow boiling heat transfer can be expressed by the boiling number \( (Bo = q/(h_v G)) \).

Yoshida et al. (1994) introduced \( Bo \) in their \( S \) model as follows:

\[
S = \frac{1}{1 + 0.9\left(\frac{Re, G^{10^{-4}}}{(Bo \times 10^4) \times X_t^{0.5}}\right)}
\]

In this study, a similar expression was proposed based on the experimental data in the presence of oil. The expression of \( S \) shown in Eq. (16) considers the suppression of nucleate boiling due to both convective heat transfer and the presence of oil. Since the effect of oil on the nucleate boiling is sophisticated—involving bubble generation suppression, retarded bubble departure, increased surface tension, separation of the oil layer from liquid refrigerant, and the corresponding heat transfer resistance of the oil film—a new factor \( S_{oil} \) was introduced into Eq. (16) to reflect all of the other effects that could not be represented by the mixture properties. According to Gao et al. (2007), the heat transfer coefficient of a mixture with an oil concentration of less than 0.05% is similar to that of pure CO₂. Therefore, the lower limit for the oil concentration was set to 0.05% in Eq. (17).

Figure 8 compares the four \( S \) values. The regulation of \( S \) calculated by Eq. (16) and the correlation proposed by Chen (1966) decreased quickly at the beginning and then slowly decline after the vapor reached 0.5. The comparison shows that the proposed \( S \) had similar
values as the original Chen’s model at low vapor qualities. When the vapor quality was increased, the decrease in $S$ of the proposed model was more smooth than that of Chen’s model. In contrast, the $S$ values of Aiyoshizawa et al.’s model (2006) were much lower than those of the other models because they represented the suppression of nucleate boiling due to the presence of oil by reducing the $S$ value. In addition, the $S$ of Yoshida et al.’s model (1994) in Eq. (22) was much higher than that of the other models.

Figure 9 shows the tendencies of the predicted heat transfer coefficient based on the contributions of different constituent terms to the variation in the overall heat transfer coefficient, against the vapor quality. The contribution of nucleate boiling decreased and that of convective heat transfer increased with increasing vapor quality before the onset of dryout. The total heat transfer coefficient ($\alpha_{tp}$) increased with the mass flux, heat flux, and vapor quality. All of these behaviors were consistent with the regulation based on the experimental results of Dang et al. (2013). According to the experimental results, the nucleate boiling and convective heat transfer alternated dominance over the flow boiling at heat transfer under various conditions. At a mass flux of 360 kW m$^{-2}$ s$^{-1}$, the nucleate boiling term ($S\alpha_{nb}$) increased significantly with the heat flux. At a heat flux of 18 kW m$^{-2}$, the nucleate boiling and convective heat transfer were dominant in the lower and higher vapor quality regions, respectively. The overall heat transfer coefficient ($\alpha_{tp}$) was almost constant over the vapor quality range of 0.2–0.8. At a high heat flux of 36 kW m$^{-2}$, the nucleate boiling term was large and decreased linearly with increasing vapor quality until the vapor quality reached 0.8. At a small heat flux of 4.5 kW m$^{-2}$, the convective part of the heat transfer coefficient ($F\alpha_{l}$) dominated almost the entire pre-dryout flow boiling region.

4. COMPARISON of PREDICTED VALUE with EXPERIMENTAL RESULTS

A database was set up based on data from four independent research groups, as listed in Table 4. The database covers experimental data obtained using a mini-channel of 0.86 mm by Zhao et al. (2002) to data by Gao et al. (2008) using a tube diameter of 3.76 mm. Because Zhao et al. (2002) did not explain the type of oil they used, PAG-type oil was considered for this calculation. For the data from Katsuta et al. (2008a, 2008b), an average oil concentration was used in the calculation because they presented their groups of measured values for different oil ranges rather than exact oil concentrations. Since the regulation of flow boiling heat transfer observed in the post-dryout region was not similar to that in the pre-dryout region, only the predicted flow boiling heat transfer coefficient in the pre-dryout region was considered in this analysis.

This experimental database was used to compare six prediction methods: namely, the correlations presented by Katsuta et al. (2008), Gao et al. (2008), Aiyoshizawa et al. (2006), Hwang et al. (1997), Thome et al. (2004), and the newly proposed model. The correlations proposed by Hwang et al. (1997) and Thome et al. (2004) for pure CO$_2$ were used, and the
mixture properties of CO₂ and oil were applied in these correlations. In contrast, the properties of CO₂ were employed in the correlations of Katsuta et al. (2008), Gao et al. (2008), and Aiyoshizawa et al. (2006) following the original proposal.

In order to compare the accuracies of the prediction methods with the experimental database, the deviation was defined as shown in Eq. (23). Two characteristic parameters were used to provide a global quantitative measure of the performance of the new correlation; their definitions are shown in Eqs. (24) and (25). The relative mean absolute error (RMAE) and relative standard deviation (RSD) in Eqs. (24) and (25) indicate how close the predictions were to the eventual outcomes.

\[
ED = \frac{y_{\text{exp}} - y_{\text{cal}}}{y_{\text{exp}}} \quad (23)
\]

Relative mean absolute error

\[
\text{RMAE} = \frac{1}{n} \sum_{i=1}^{n} \left| \frac{y_{\text{exp},i} - y_{\text{cal},i}}{y_{\text{exp},i}} \right| \quad (24)
\]

Relative standard deviation

\[
\text{RSD} = \sqrt{\frac{1}{n-1} \sum_{i=1}^{n} \left( \frac{y_{\text{exp},i} - y_{\text{cal},i}}{y_{\text{exp},i}} \right)^2 - \text{RMAE}^2} \quad (25)
\]

Figure 10 shows the previous experimental results (Dang et al., 2013, Katsuta et al., 2008a, 2008b, Gao et al., 2007, 2008) for CO₂ with a small amount of lubricating oil and the prediction results of the new correlations. Figure 10(a) shows that the results predicted by the newly proposed correlation illustrated similar regulations at different oil concentrations with increasing vapor quality. However, the oil concentration affected the calculated heat transfer coefficient more significantly than that indicated by the experimental results. Aiyoshizawa et al.’s (2006) results were independent of the oil concentration and did not reflect the influence of oil on the heat transfer. Figures 10(b) and (c) show that the prediction by the newly proposed correlation tended to be similar to the experimental values of Katsuta et al. (2008a) and Gao et al. (2007), and the prediction results reflected the regulation of the effect of the oil concentration on the heat transfer coefficients.

Figure 11(a) compares the predicted and measured heat transfer coefficients in a 2-mm I.D. tube with an oil concentration of 1% and mass flux of 360 kg m⁻² s⁻¹ to demonstrate the effect of the heat flux based on the data of Dang et al. (2013). The heat flux had a strong positive influence on the heat transfer coefficient for a majority of the vapor quality region when the heat flux was increased from 4.5 kW m⁻² to 36 kW m⁻², and the mass flux stayed at 360 kg m⁻².
s\(^{-1}\). The theoretical results predicted by the newly proposed correlation depicted the characteristics of the heat transfer coefficient well: they showed the tendency of the heat transfer coefficients against the vapor quality to change from an upward movement to a downward movement when the heat flux rose and the mass flux remained constant. At heat fluxes of 4.5 and 9 kW m\(^{-2}\), the theoretical heat transfer coefficients from the newly proposed correlation matched the measurements well at low vapor quality and gradually deviated from the experimental coefficients at high vapor quality. For both the heat fluxes, the predicted results tended to increase with the vapor quality in the pre-dryout region. This shows the strong influence of convective boiling on the heat transfer coefficients. At a heat flux of 18 kW m\(^{-2}\), both the measured pre-dryout heat transfer coefficients and the prediction by the new proposed correlation barely changed with variations in the vapor quality. The nucleate boiling was considered to be the dominant mechanism affecting the contribution of the heat transfer coefficient to the flow boiling of CO\(_2\). When the heat flux was increased to 36 kW m\(^{-2}\), the theoretical heat transfer coefficient decreased steadily with increasing vapor quality and matched the dynamics of the measured results well. The prediction results from the correlation of Aiyoshizawa et al. (2006) increased with the heat flux and vapor quality. However, the tendencies of some prediction data were different from those of the experimental results when the heat flux reached 18 kW m\(^{-2}\) and especially 36 kW m\(^{-2}\).

Figure 11(b) compares the measured and calculated results for a CO\(_2\)–oil mixture at a heat flux of 18 kW m\(^{-2}\), mass flux of 360–1440 kg m\(^{-2}\) s\(^{-1}\), and oil concentration of 1%. In Fig. 11(b), both the theoretical and experimental heat transfer coefficients showed that the mass flux had a positive influence on the heat transfer coefficient of the CO\(_2\)–oil mixture in the pre-dryout region; this differs from the case under oil-free conditions. Under oil-free conditions, the heat transfer coefficient with a large mass flux was slightly lower than that with a small mass flux (Dang et al., 2010). The difference was attributed to the suppression of nucleate boiling due to oil entrainment. In oil-free CO\(_2\), the density difference between the liquid and vapor phases is very small; thus, nucleate boiling is predominant for the overall heat transfer coefficient compared to convective heat transfer. However, the presence of oil retards bubble generation and bubble departure due to the increased surface tension and liquid viscosity, which suppresses nucleate boiling. The convective heat transfer of the CO\(_2\)–oil mixture thus becomes prominent. The values predicted by the new proposed correlation almost matched the trend of the measured data. When the mass flux was 360 kg m\(^{-2}\) s\(^{-1}\), the nucleate boiling and convective heat transfer were comparable, and the heat transfer coefficients showed a gentle tendency with an increase in vapor quality. When the mass flux was increased to 720 and 1440 kg m\(^{-2}\) s\(^{-1}\) at a heat flux of 18 kW m\(^{-2}\), the influence of the dominant convective heat transfer caused the heat transfer coefficient to increase with the vapor quality.

Figures 12(a)–(d) compare the measured and calculated flow boiling heat transfer...
coefficients of the CO₂–oil mixture using the above correlations. The comparisons showed that the best-fitting correlations were those of Aiyoshizawa et al. (2006) and the proposed model, which were based on almost the same experimental results. These correlations were in good agreement with approximately 95% of the data of Dang et al. (2013) and also corroborated about 90% of the data of Gao et al. (2007, 2008) and approximately 70% of the data of Katsuta et al. (2008a, 2008b) within a deviation of ±30%. Because of the smaller inner diameter of the test tube (0.86 mm) used by Zhao et al. (2002) and the unknown oil type, the prediction deviations of these two correlations are large. The correlations proposed by Katsuta et al. (2008) and Gao et al. (2008) partially corroborated the data from Dang et al. (2013), Gao et al. (2007, 2008), and Katsuta et al. (2008a, 2008b). Table 5 presents the RMAE and RSD of the six correlations. The correlations of Hwang et al. (1997) and Thome et al. (2004) for pure CO₂, when used with the properties of the mixture, could not obtain acceptable deviations. The correlations for pure CO₂ must therefore be corrected before they can be utilized to predict the heat transfer coefficient of a CO₂–oil mixture. The correlations of the proposed model and Aiyoshizawa et al. (2006) outperformed the other correlations in terms of predicting the flow boiling heat transfer of the CO₂–oil mixture. The main differences between these two correlations were that the proposed model considered the properties of the mixture during calculation and provided superior performance when based on the measured results of Dang et al. (2013).

5. CONCLUSIONS

This paper proposed a prediction model for the flow boiling heat transfer of CO₂ with a small amount of PAG-type lubricating oil. The main conclusions are as follows:

(1) A prediction model for the heat transfer coefficient is proposed that considers the properties of the CO₂–oil mixture using new F and S factors. The proposed F factor with the mixture properties is similar to that proposed for pure refrigerant. Factor S was revised by introducing factor S_{oil} to consider the influence of oil on nucleate boiling.

(2) The proposed correlation matched most (95%) of the experimental results of Dang et al. (2013) with a deviation of ±30%. It also matched 90% of the data from Gao et al. (2007, 2008) within a deviation of ±30%.

(3) Based on this model, increasing the oil concentration causes a gentle decrease in the predicted heat transfer coefficient of a CO₂–oil mixture.

(4) The proposed correlation can describe the tendency of the heat transfer coefficient of a CO₂–oil mixture at various heat and mass fluxes and agrees well with the experimental results.

ACKNOWLEDGMENTS

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REFERENCES


Garcia, J., Youbi-Idrissi, M., Bonjourc, J., Fernandez, J. 2008, Experimental and PC-SAFT volumetric and phase behavior of carbon dioxide + PAG or POE lubricant systems, J. of Supercritical Fluids, 47, 8–16.


Refrigerant System Symposium, Scottsdale, AZ.
Appendix 1

Measured accuracy of the parameters in the experiment of Dang et al. (2013)

<table>
<thead>
<tr>
<th>Item</th>
<th>Uncertainties</th>
</tr>
</thead>
<tbody>
<tr>
<td>oil concentration ($\omega_o$)</td>
<td>0.1%</td>
</tr>
<tr>
<td>Heat flux ($q_w$)</td>
<td>±3%</td>
</tr>
<tr>
<td>$T_w$</td>
<td>±0.1 °C</td>
</tr>
<tr>
<td>$T_{sat}$</td>
<td>±0.03 °C</td>
</tr>
<tr>
<td>Vapor fraction</td>
<td>±3.3%</td>
</tr>
<tr>
<td>Heat transfer coefficient ($\alpha$)</td>
<td>From ±8.9% to ±13%</td>
</tr>
</tbody>
</table>

Appendix 2

1) Correlation of Hwang et al. (1997) for pure CO2

\[
\alpha_{np} = \alpha_{nb} + \alpha_{cb}
\]  

\[
\alpha_{nb} = 0.00122S \left( \frac{\lambda_{lp}^{0.79} c_{pl}^{0.5} \rho_{pl}^{0.49}}{\sigma^{0.6} \mu_{li}^{0.29} h_{lw}^{0.24} \rho_{li}^{0.24}} \right) \left[ T_w - T_{sat} (p_i) \right]^{0.4} \left[ p_{sat} (T_w) - p_i \right]^{0.75}
\]  

\[
S = 1 - \exp \left( -F \alpha_1 X / \lambda_i \right), X = 0.05 \left( \frac{\sigma}{g(\rho_{li} - \rho_v)} \right)^{0.5}
\]  

\[
\alpha_{cb} = \alpha_i F \Pr_i^{0.6}, \quad \alpha_i = 0.023 \Re_i^{0.8} \Pr_i^{0.4} (\lambda_i / d)
\]  

\[
F = 1.0, X_\eta \geq 10
\]  

\[
F = 2.0 (0.213 + 1 / X_\eta)^{0.736}, X_\eta < 10
\]

\[
X_\eta = \left( \frac{1-x}{x} \right)^{0.9} \left( \frac{\rho_v}{\rho_i} \right)^{0.5} \left( \frac{\mu_i}{\mu_v} \right)^{0.1}
\]

2) Correlation of Thome et al. (2004) for pure CO2

\[
\alpha_{np} = \frac{\theta_{dry} \alpha_v + d \left( 2\pi - \theta_{dry} \right) \alpha_{wet}}{2\pi d}
\]  

\[
\alpha_{wet} = \left[ (S \alpha_{nh,CO2})^3 + \alpha_{cb}^{1/3} \right]^{1/3}
\]
\[ \alpha_{nb} = 55 p_r^{0.12} \left( - \log_{10} p_r \right)^{-0.55} M^{-0.5} q^{0.67} \quad (A-10) \]

\[ \alpha_{cb} = 0.0133 \left[ \frac{4G(1-x)\delta}{(1-\varepsilon)\mu_1} \right]^{0.69} \left( \frac{c_{nl} \mu_1}{\lambda_1} \right)^{0.4} \frac{\lambda_1}{\delta} \quad (A-11) \]

\[ \alpha_v = 0.023 \left( \frac{Gxd}{\varepsilon \mu_v} \right)^{0.8} \left( \frac{c_p \mu_v}{\lambda_v} \right)^{0.4} \frac{\lambda_v}{d} \quad (A-12) \]

\[ \alpha_{nb,CO_2} = 0.71\alpha_{nb} + 3790 \quad (A-13) \]

\[ S = \frac{(1-x)^{1/2}}{0.121Re_t^{0.225}} \quad (A-14) \]

\[ \varepsilon = \frac{x}{\rho_v} \left[ (1+0.12(1-x)) \left( \frac{x}{\rho_v} + \frac{1-x}{\rho_1} \right) + \frac{1.18}{G} \left( \frac{g\sigma(\rho_1 - \rho_v)}{\rho_v^2} \right)^{1/4} \right] (1-x)^{-1} \quad (A-15) \]

3) The correlations of Gao et al. (2008) are the same as those of Cheng et al. (2006) except for additional suppression factors for the CO₂–oil mixture.

\[ \alpha_{tp} = \left[ (S\alpha_{nb})^3 + \alpha_{cb}^3 \right]^{1/3} \quad (A-16) \]

\[ S' = \begin{cases} S \times (-0.5x + 0.35) & (x \leq 0.7) \\ 0 & (x > 0.7) \end{cases} \quad (A-17) \]

4) Correlation of Katsuta et al. (2008b) for CO₂–oil mixture

\[ \alpha_{tp} = \alpha_{nb} + \alpha_{cb} \quad (A-18) \]

\[ \alpha_{cb} = F \times 0.023 \left( \frac{\lambda_1}{D} \frac{G(1-x)D}{\mu_1} \right)^{0.8} Pr^{0.4} \quad (A-19) \]

\[ F = 1 + 0.258 \left( \frac{1}{X_n} \right)^{0.886} + 92.32 \left( \frac{\rho_v}{\rho_1} \right)^3 \left( \frac{1}{X_n} \right)^{0.9} \quad (A-20) \]

\[ \alpha_{nb} = Sa_{SA} \times \phi \quad (A-21) \]
\[ \alpha_{SA} = 207 \left( \frac{\lambda_t}{D_b} \right)^{0.745} \left( \frac{qD_b}{\lambda_t T_{sat}} \right)^{0.581} \left( \frac{\rho_v}{\rho_i} \right) \Pr_i^{0.533} \]  
\[ (A-22) \]

\[ S = \ln \left( 2.332 \frac{Bo^{0.518} Bd^{1.27} Fr^{0.964}}{Re_{ip}^{0.834}} \right) \]  
\[ (A-23) \]

\[ \phi = \frac{1}{(1 + OCR^{0.698} Bo^{0.207} Bd^{0.912})} \]  
\[ (A-24) \]

\[ x_{dry} = 0.269 \times \frac{Re^{0.0571} Fr^{0.0697}}{Bo^{0.0519}} \]  
\[ OCR < 1 \quad \eta = 1 \]  
\[ OCR \geq 1 \quad \eta = 1.169 e^{-0.17xOCR} \]  
\[ (A-25) \]

5) Correlation of Aiyoshizawa et al. (2005)

\[ \alpha_{tp} = F \alpha_{ch} + S \alpha_{nb} \]  
\[ (A-28) \]

\[ \alpha_{ch} = 0.023 \left( \frac{\lambda_t}{d} \right)^{0.8} \left( \frac{G(1-x)d}{\mu_i} \right)^{0.4} \Pr^{0.4} \]  
\[ (A-29) \]

\[ \alpha_{nb} = 207 \left( \frac{\lambda_t}{D_b} \right)^{0.745} \left( \frac{qD_b}{\lambda_t T_b} \right)^{0.581} \left( \frac{\rho_v}{\rho_i} \right) \Pr_i^{0.533} \]  
\[ (A-30) \]

\[ D_b = 0.51 \left( \frac{2\sigma}{g(\rho_i - \rho_v)} \right)^{0.5} \]  
\[ (A-31) \]

\[ Re_{ip} = \left( \frac{G(1-x)D^2}{\mu_i} \right)^{1.25} \]  
\[ (A-32) \]

\[ F = 0.8 + 0.5 \left( \frac{1}{X_{ti}} \right)^{1.2} \]  
\[ (A-33) \]

\[ S = \frac{1}{6 + \left( Re_{ip} \times 10^{-4} \right)^3} \]  
\[ (A-34) \]
Appendix 3

\begin{equation}
\rho_m = \rho_o \cdot \omega_{oi} + \rho_r \cdot (1 - \omega_{oi}) \tag{A-35}
\end{equation}

\begin{equation}
\ln \nu_m = \omega_{oi} \ln \nu_o + (1 - \omega_{oi}) \ln \nu_r \tag{A-36}
\end{equation}

\begin{equation}
c_{pm} = (1 - \omega_{oi})c_{pr} + \omega_{oi}c_{po} \tag{A-37}
\end{equation}

\begin{equation}
\sigma_m = \sigma_r + (\sigma_o - \sigma_r)\sqrt{1 - \omega_i} \tag{A-38}
\end{equation}

\begin{equation}
\lambda_m = \lambda_r \omega_r + \lambda_o (1 - \omega_r) - 0.72 (1 - \omega_r) \omega_r (\lambda_o - \lambda_r) \tag{A-39}
\end{equation}

Conde (1996) presented an equation to predict the density of oil against temperature:

\begin{equation}
\rho(T) = \rho_0(T) - A(T_0 - T) \tag{A-40}
\end{equation}

where \( T_0 \) is a reference temperature and was taken as 15 °C here; \( \rho_0(T_0) \) is the density at the reference temperature (kg m\(^{-3}\)); and \( A = 0.8 \) for the oil proposed by Conde.

Liley et al. (1973) proposed a formula to calculate the specific heat:

\begin{equation}
C_p = 4.1861 \sqrt{S_0} \left[ 0.388 + 0.00045 \left( \frac{9}{5} t + 32 \right) \right] \tag{A-41}
\end{equation}

where \( S_0 \) is the relative density. \( S_0 = \rho_o/\rho_w \), which is the ratio of the liquid density and pure water density (999.0 kg m\(^{-3}\) at a temperature of 15 °C).

Robbins and Kingrea (1962) proposed a correlation to calculate the conductivity of a fluid as a function of the molecular weight, density, specific heat, and critical temperature.

\begin{equation}
\lambda = \frac{\Delta_1}{M^{1/2}} \left( \frac{3 + 20 \left( 1 - \frac{T}{T_c} \right)^{2/3}}{3 + 20 \left( 1 - \frac{T_{sb}}{T_c} \right)^{2/3}} \right)^{2/3} \tag{A-42}
\end{equation}

where \( \Delta_1 = 1.5 \) for lubricants. \( T_{sb} \) and \( T_c \) are the standard boiling point and critical temperature, respectively, of PAG oil (K) as shown in Table 2. Based on the data in Table 2, the maximum deviation between the properties calculated by Eq. (A-42) and those provided by the manufacturer of the oil is less than ±5%.

The viscosity of PAG oil is correlated to the temperature in the following function:

\begin{equation}
\lg(\lg(\nu + A)) = B \lg T + C \tag{A-43}
\end{equation}

where \( A, B, \) and \( C \) have recommended values of 0.6, 2.468, and 6.46, respectively, to best fit the viscosity data given in Table 2.
Figure 1 Schematic diagram of experimental apparatus.
Figure 2 Two-phase separation temperature.
Figure 3 Variation in local oil concentration against vapor quality.
Figure 4 Variation in boiling heat transfer coefficient ratio at various oil concentrations against Lockhart–Martinelli parameter.
Figure 5 Variation in boiling heat transfer coefficient ratio against Lockhart–Martinelli parameter at $\omega_o = 1\%$. 
Figure 6 Comparison of $\alpha_l$ and $\alpha_{nb}$ calculated for pure CO$_2$ and for CO$_2$–oil.

\[
G = 360 \text{ kgm}^{-2}\text{s}^{-1}
\]

$\alpha_{nb}$ pure CO$_2$

$\alpha_{nb}$ CO$_2$–oil $\left(\omega_0 = 1\%\right)$
Figure 7 Variations in factor $F$ values of four correlations at 15 °C against vapor quality.
Figure 8 Variations in factor $S$ values of four correlations at different mass and heat fluxes against vapor quality.
Figure 9 Contributions of different constituent terms to variation in total heat transfer coefficient ($\alpha_{tp}$) from nucleate boiling ($S\alpha_{nb}$) and convective heat transfer ($F\alpha_l$) with the vapor quality.
Fig. 10 (a) Data of Dang et al. (2013)

Fig. 10 (b) Data of Katsuta et al. (2008a)
Figure 10 Comparison of measured heat transfer coefficients and those calculated using proposed correlation at different oil concentrations.
Figure 11 Comparison of theoretical heat transfer coefficients based on correlations of authors and Aiyoshizawa et al. (2006) and measured data in 2-mm I.D. tube at different mass and heat fluxes based on data of Dang et al. (2013).
Fig. 12 (a) Correlations proposed by Gao et al. (2008)

Fig. 12 (b) Correlations proposed by Katsuta et al. (2008)
Figure 12 Comparison of heat transfer coefficients predicted using six correlations and database.

Fig. 12 (c) Correlations proposed by Aiyoshizawa et al. (2006)

Fig. 12 (d) New correlations proposed by authors

Figure 12 Comparison of heat transfer coefficients predicted using six correlations and database.
Table 1 Experimental conditions

<table>
<thead>
<tr>
<th>Items</th>
<th>CO₂–oil mixture</th>
<th>CO₂–oil mixture</th>
<th>oil</th>
<th>Pure CO₂</th>
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</thead>
<tbody>
<tr>
<td></td>
<td>1 % oil</td>
<td>5% oil</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Tube Material</td>
<td>SUS316</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Tube I.D. [mm]</td>
<td>2.0</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Oil type</td>
<td>PAG100</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Oil concentration [%]</td>
<td>0.5–5.0</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Mass velocity [kg m⁻² s⁻¹]</td>
<td>360, 720, 1440</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Heat flux [kW m⁻²]</td>
<td>4.5, 9, 18, 36</td>
<td></td>
<td></td>
<td></td>
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<tr>
<td>Evaporating temperature [°C]</td>
<td>15</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Quality</td>
<td>Approximately 0–1.0</td>
<td></td>
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Table 2 Properties of PAG-type oil

<table>
<thead>
<tr>
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<th>100</th>
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<tbody>
<tr>
<td>Temperature [°C]</td>
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<td></td>
<td></td>
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</tr>
<tr>
<td>Molecular weight [kg kmol⁻¹]</td>
<td>1600</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Density [kg m⁻³]</td>
<td>---</td>
<td>1000.6</td>
<td>---</td>
<td>1028.6</td>
<td>---</td>
</tr>
<tr>
<td>Specific heat [kJ kg⁻¹ K⁻¹]</td>
<td>1.7577</td>
<td>---</td>
<td>---</td>
<td>1.9251</td>
<td>2.0925</td>
</tr>
<tr>
<td>Conductivity [W m⁻¹K⁻¹]</td>
<td>0.1478</td>
<td>---</td>
<td>---</td>
<td>0.1465</td>
<td></td>
</tr>
<tr>
<td>Kinematic viscosity [mm² s⁻¹]</td>
<td>---</td>
<td>---</td>
<td>106.1</td>
<td>---</td>
<td>20.29</td>
</tr>
<tr>
<td>Standard boiling point [K]</td>
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<td></td>
<td></td>
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<td>845.41</td>
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<tr>
<td>Critical temperature [K]</td>
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<td></td>
<td></td>
<td></td>
<td>968.57</td>
</tr>
<tr>
<td>Eccentricity factor</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td>1.352</td>
</tr>
</tbody>
</table>

Table 3 Comparison of the properties of the oil–CO₂ mixture and pure CO₂

<table>
<thead>
<tr>
<th>Items</th>
<th>CO₂–oil mixture 1 % oil</th>
<th>CO₂–oil mixture 5% oil</th>
<th>oil</th>
<th>Pure CO₂</th>
</tr>
</thead>
<tbody>
<tr>
<td>Liquid conductivity [W m⁻¹ K⁻¹]</td>
<td>0.09195</td>
<td>0.09355</td>
<td>0.1232</td>
<td>0.09186</td>
</tr>
<tr>
<td>Liquid viscosity [Pa s⁻¹]</td>
<td>7.35 × 10⁻⁴</td>
<td>12.56 × 10⁻⁴</td>
<td>3.41 × 10⁻³</td>
<td>7.44 × 10⁻⁵</td>
</tr>
<tr>
<td>Liquid density [kg m⁻³]</td>
<td>822.73</td>
<td>828.85</td>
<td>1006.84</td>
<td>821.21</td>
</tr>
<tr>
<td>Specific heat [J kg⁻¹ K⁻¹]</td>
<td>3425.16</td>
<td>3356.61</td>
<td>1728.58</td>
<td>3442.3</td>
</tr>
<tr>
<td>Surface tension [N m⁻¹]</td>
<td>0.00422</td>
<td>0.00701</td>
<td>0.0245</td>
<td>0.00196</td>
</tr>
</tbody>
</table>

The saturation temperature of pure CO₂ is 15 °C.
Table 4 Experimental studies on flow heat transfer of CO$_2$–oil mixtures

<table>
<thead>
<tr>
<th>References</th>
<th>Tube Inner</th>
<th>Heat flux</th>
<th>Mass flux</th>
<th>Oil concentration</th>
<th>Oil test methods</th>
<th>Saturated temperature</th>
<th>Number of data items</th>
</tr>
</thead>
<tbody>
<tr>
<td>Katsuta et al. (2008a)</td>
<td>Smooth stainless D = 3 mm, L = 0.5 m</td>
<td>5, 10, 15</td>
<td>400</td>
<td>PAG 100, 0%, 0.05%–0.5%, 0.6%–1.2%, 1.8%–2.5%, 3.0%–4.0%</td>
<td>Oil supply system, oil mass flow meter, and oil sampling tank</td>
<td>–10, 0, 10</td>
<td>55</td>
</tr>
<tr>
<td>Katsuta et al. (2008b)</td>
<td>Smooth stainless D = 3 mm, L = 0.5 m</td>
<td>10</td>
<td>200, 400, 600</td>
<td>PAG 0%–0.05%, 0.5%–1.0%, 1.0%–2.0%, 2.0%–3.0%, 3.0%–5.0%, 7.0%–9.0%</td>
<td>Oil sampling tank</td>
<td>0</td>
<td>34</td>
</tr>
<tr>
<td>Gao et al. (2008)</td>
<td>Smooth inner stainless D = 3.76 mm</td>
<td>10, 20, 30</td>
<td>100, 190, 380</td>
<td>PAG 1.01% PAG &lt; 0.01%, 0.09%, 0.17%, 0.48%, 0.72%</td>
<td>Measured from the size and time interval of an oil droplet falling into the sight glass chamber</td>
<td>10</td>
<td>51</td>
</tr>
<tr>
<td>Gao et al. (2007)</td>
<td>Smooth stainless D = 3 mm, L = 2.185 m; Micro-fin D = 3.04 mm, Fin height = 0.11 mm</td>
<td>5, 10, 20, 30</td>
<td>190, 380, 770, 390, 780, 1180</td>
<td>PAG 0.01%, 0.11%, 0.30%, 0.57%</td>
<td>Oil sampling tube</td>
<td>10</td>
<td>53</td>
</tr>
<tr>
<td>Zhao et al. (2002)</td>
<td>Mini tube D = 0.86 mm</td>
<td>11</td>
<td>300</td>
<td>Oil type N/A 0%, 1%, 3%, 4%, 5%, 7%</td>
<td>N/A</td>
<td>10</td>
<td>54</td>
</tr>
<tr>
<td>Dang et al. (2013)</td>
<td>Smooth stainless D = 2 mm</td>
<td>4.5, 9, 18, 36</td>
<td>360, 720, 1440</td>
<td>PAG 100, 0%, 1%, 3%, 5%</td>
<td>Oil sampling tube</td>
<td>15</td>
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Table 5 Deviations of methods when predicting heat transfer data at \( \omega_0 > 0.05\% \)

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<td>Model</td>
<td>Mean absolute error</td>
<td>Standard deviation</td>
<td>Mean absolute error</td>
<td>Standard deviation</td>
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<tr>
<td>Thome et al. (2004)</td>
<td>–146%</td>
<td>290%</td>
<td>–141%</td>
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<td>Hwang et al. (1997)</td>
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<td>Katsuta et al. (2008a)</td>
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<td>24%</td>
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<td>Gao et al. (2008)</td>
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<td>112%</td>
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<td>Aiyoshizawa et al. (2006)</td>
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