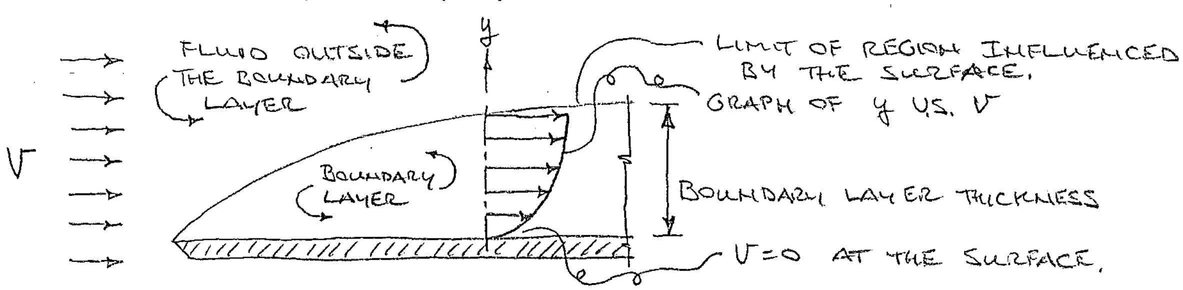
Heat Transfer

Calculation of the Convective Heat Transfer Coefficient

Fluid Velocity Boundary Layer

Imagine that a fluid (gas or liquid) is flowing along. If you look around inside the fluid you will find places where the fluid comes in contact with solid objects. The solid objects might be the walls of a room, your skin, the inside surface of a coffee cup – really any solid at all. If you were to take a really close look at the place where the solid and fluid meet you will find that the fluid atoms are stuck to the solid. The fluid is so stuck that right at the solid surface the fluid is not moving. Just a few atoms away from the surface the fluid atoms are moving. They slide past the fluid atoms stuck to the wall but are slowed by the attractive forces that exist between themselves and the stationary atoms stuck to the wall. Those slower fluid atoms act to slow the fluid atoms next to them and so on. At some small distance away from the solid surface you will find that the fluid is moving along at the bulk fluid velocity, as though the solid surface is not there at all.

The layer of fluid atoms slowed, or as they say "influenced", by the presence of the solid wall is called the Velocity Boundary Layer.



Within the Velocity Boundary Layer there are two competing forces that tend to control how the fluid moves across the solid surface. Inertial forces arise from the fluids velocity and tend to amplify any disturbances in the flow. On the other hand, Viscous forces arise from the attractive forces between fluid atoms and tend to dampen the attempt by the inertial forces to make the flow unsteady.

At low velocities the inertial forces are small and the viscous forces are relatively large so the fluid flows smoothly in a well organized pattern called *Laminar Flow*. At higher velocities the inertial forces are large and tend to dominate the relatively smaller viscous forces so the flow has a tendency to move in a highly disorganized way known as *Turbulent Flow*.

It is possible to numerically predict whether the flow will be laminar or turbulent by calculating the *Reynolds Number*. The Reynolds Number is simply a calculated ratio of the Inertial Forces over the Viscous Forces. The equation looks like this.

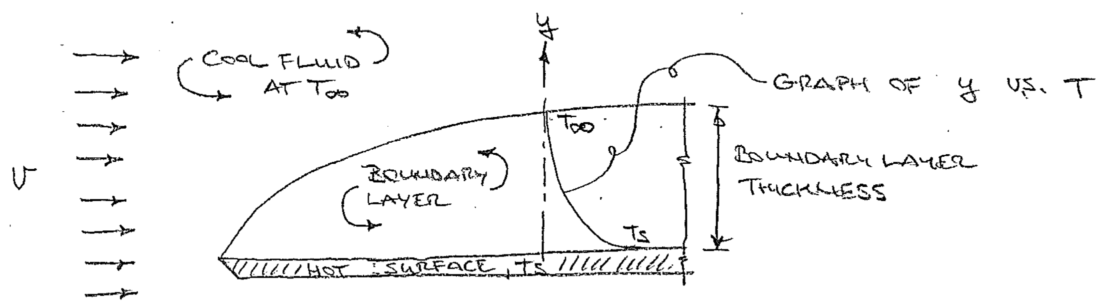
$$\mathrm{Re} = \frac{VL\rho}{\mu} \qquad \textit{also} \qquad \mathrm{Re} = \frac{VL}{\nu}$$

Where:
- V = Flow Velocity outside the Boundary Layer, m/s
- L = Dimensional Parameter, m
- ρ = Fluid density, kg/m^3
- μ = Dynamic viscosity, kg/m-s
- ν = Kinematic viscosity, m^2/s $= (\rho/\mu)$

You will notice that, in a way, Re is simply a dimensionless form of velocity. The size of the number that emerges from the equation is an indicator of the type of flow that will exist. For flow moving inside a pipe, for example, laminar flow occurs when Re is less than 2000 and turbulent flow occurs when Re is greater than 4000. Between 2000 and 4000 the flow can be either laminar or turbulent – its unpredictable.

Fluid Thermal Boundary Layer

When a cool fluid moves across a warm surface it warms up. The fluid close to the warm surface becomes hotter than the fluid farther away from the surface. As a result of the observed fluid temperature distribution near the surface it is said that the fluid has a Thermal Boundary Layer. Like the Velocity Boundary Layer, the Thermal Boundary Layer is the layer of fluid close to a surface that is "influenced" by the surface. Its shape and thickness are usually different than the Velocity Boundary Layer but the way heat is transferred through the Thermal Boundary Layer is definitely affected by the nature of flow near the surface (laminar or turbulent).



Convective Heat Transfer

Recall that we have discussed the calculation of two major types of heat transfer: Conductive and Convective.

One can determine the rate of heat transfer in each of these forms as follows:

Conductive Heat Transfer Equation: $\dot{Q}_{COND} = \frac{kA}{L}(T_{HOT} - T_{COLD})$

Convective Heat Transfer Equation: $\dot{Q}_{CONV} = hA(T_S - T_\infty)$

To find "k", the Conductive Heat Transfer Coefficient, you just have to look up its value in the tables in the appendix of this document.

To find "h", the Convective Heat Transfer Coefficient, is a bit more work. The next few pages hold a step-by-step method of calculating "h". As part of the calculation process you will need to look up certain fluid properties in the fluid tables in the appendix this document.

(I'm sure you are aware that when looking up properties you simply need to determine the material through which the heat is passing, find that material in the tables and look in the appropriate column for the property. A trick to make sure you've found the right property is to check your units. If the units of the property seem to match the units required in an equation you probably are looking in the correct column.)

Calculating "h" – The Convective Heat Transfer Coefficient

Lots of experiments have been done relating "h" with various fluid properties, different flow types and a wide variety of surface shapes. Several important relationships have immerged from this work.

Nusselt Number: $$Nu = \frac{hL}{k}$$

Where:
h = Convective heat transfer coefficient, W/m^2-C
L = Dimensional Parameter, m
k = Thermal Conductivity, W/m-C

Your goal is to find "h" ... so ... just solve for it. If you have Nu, L and k, you will have "h".

$$h = Nu\frac{k}{L}$$

The trick is to find Nu, the Nusselt Number. Experiments have shown that Nu is related to two or three fluid properties: the Reynolds Number, Re (that we just looked at), the Prandtl Number, Pr and (sometimes) the Grashof Number, Gr.

Prandtl Number: $$\Pr = \frac{\mu C_P}{k}$$

Where:
C_P = Specific Heat of the fluid, J/kg-C

Often the Prandtl number is given directly in the tables of properties so you usually don't have to calculate it. You may have already noticed that it's another dimensionless number.

Grashof Number: $$Gr = \frac{g\beta(T_{SURFACE} - T_{\infty})L^3\rho^2}{\mu^2}$$

Where:
g = gravitational acceleration constant = 9.81 m/s^2
β = Isobaric Compressibility, 1/C (See note below.)

Note: See the section entitled "A Few Notes About Fluids and Heat" for more information on how to determine Cp, β and ρ.

Steps-by-Step Method finding "h"

1. Look at the shape and size of the convective heat transfer situation. Based on the different flow types and the illustrations presented in Table C-8 below, determine the geometric parameter, L. (Notice in some situation the geometric parameter is shown using the variable D for diameter. D and L are interchangeable.)
2. Determine if a fan or pump forces the flow to move (Forced Convection: Flow situations A, B and C on Table C-8 below) or if the flow moves naturally as a result of buoyancy forces (Natural convection: Flow situation D on Table C-8 below).
3. Determine the type of fluid being used to convect away the heat. Look-up the thermo-physical properties of that fluid in the tables provided.

4. Calculate the Reynolds Number, Re and the Prandtl Number (if its not given in the tables).
   If the flow is of the Natural Convection Type you will also have to calculate the Grashof Number, Gr.
5. Calculate Nu.
6. Calculate h.
7. Do what ever the problem asks you to do.

A few Notes About Fluids and Heat

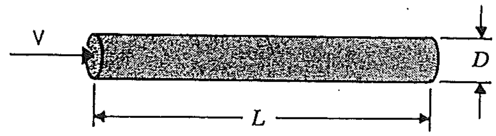
Thing 1: The tables give very complete property information for liquids. Just look up the properties – no big deal.

Thing 2: If you are dealing with a gas, like air, you will find the tables lacking when it comes to finding β. For most simple (ideal) gasses finding it is easy.

$\beta = \frac{1}{T}$ (For gasses only!) Where: T = Absolute temperature, K
OF THE GAS.

Table C·8 Convection heat transfer correlations

A. Flow in circular tubes



$$\mathrm{Re} \equiv \frac{VD\rho}{\mu}$$

$$\mathrm{Nu} \equiv \frac{hD}{k}$$

$$h \equiv \frac{q''}{\Delta T}$$

$$\Delta T = T_{\text{wall}} - T_{\text{mean}}$$

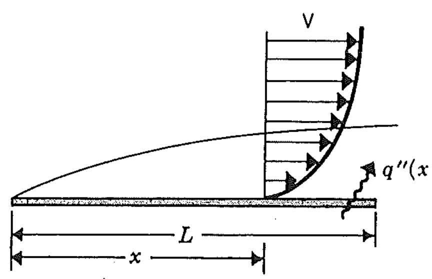
1. *Laminar flow,* Re < *2000*

   Fully developed flow $\dfrac{L/D}{\mathrm{RePr}} > 0.05$

   Nu = 4.364 *uniform wall heat flux*
   Nu = 3.66 *uniform wall temperature*

2. *Turbulent flow,* Re > *2000*
   - Pr < *0.1* (*liquid metals*) $10^2 < (\mathrm{RePr}) < 10^4$
     - $\mathrm{Nu} = 4.82 + 0.0185\,(\mathrm{RePr})^{0.827}$ *uniform wall heat flux*
   - *0.5* < Pr < *1.0* (*gases*)
     - $\mathrm{Nu} = 0.022\,\mathrm{Pr}^{0.6}\mathrm{Re}^{0.8}$
   - *1.0* < Pr < *20* *water and light oils*
     - $\mathrm{Nu} = 0.0155\,\mathrm{Pr}^{0.5}\mathrm{Re}^{0.83}$
   - Pr > *20*
     - $\mathrm{Nu} = 0.0118\,\mathrm{Pr}^{0.3}\mathrm{Re}^{0.9}$

B. Boundary layer on a flat plate



$$q''_{\text{av}} \equiv \frac{1}{L}\int_0^L q''(x)\,dX \qquad \Delta T \equiv T_{\text{wall}} - T_{\text{flow}}$$

$$h_x \equiv \frac{q''}{\Delta T} \qquad h_L \equiv \frac{q''_{\text{av}}}{\Delta T}$$

$$\mathrm{Nu}_x \equiv \frac{h_x x}{k} \qquad \mathrm{Nu}_L \equiv \frac{h_L L}{k} \qquad \mathrm{Re}_L = \frac{V\rho L}{\mu} \qquad \mathrm{Re}_x = \frac{V\rho x}{\mu}$$

1. *Laminar flow,* $\mathrm{Re}_L <$ 5×10^5
   0.5 < Pr < *15*

   $$\left.\begin{aligned}\mathrm{Nu}_x &= 0.332\,\mathrm{Re}_x^{1/2}\mathrm{Pr}^{1/3}\\ \mathrm{Nu}_L &= 0.664\,\mathrm{Re}_L^{1/2}\mathrm{Pr}^{1/3}\end{aligned}\right\}\ \textit{uniform wall temperature}$$

   $$\left.\begin{aligned}\mathrm{Nu}_x &= 0.453\,\mathrm{Re}_x^{1/2}\mathrm{Pr}^{1/3}\\ \mathrm{Nu}_L &= 0.906\,\mathrm{Re}_L^{1/2}\mathrm{Pr}^{1/3}\end{aligned}\right\}\ \textit{uniform wall heat flux}$$

2. *Turbulent flow,* $\mathrm{Re}_L >$ 5×10^5
   0.5 < Pr < 60
   $\mathrm{Nu}_x = 0.0295\,\mathrm{Pr}^{0.6}\mathrm{Re}^{0.8}$

   $\mathrm{Nu}_L = 0.037\,\mathrm{Pr}^{1/3}\mathrm{Re}_L^{0.8}$

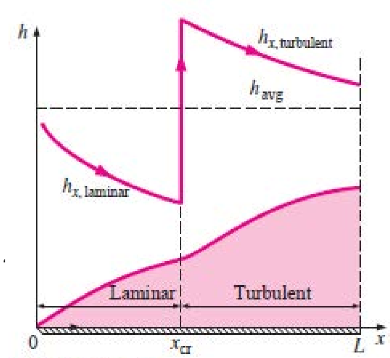
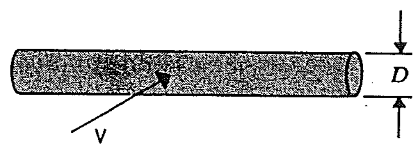


FIGURE 12–29
Graphical representation of the average heat transfer coefficient for a flat plate with combined laminar and turbulent flow.

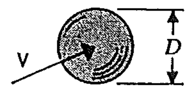
Table C·8 (*continued*)

C. Single circular cylinder or sphere in cross flow



$$\mathrm{Re} \equiv \frac{VD\rho}{\mu} \qquad \mathrm{Nu} \equiv \frac{\mathrm{h}D}{k}$$

$$\mathrm{h} \equiv \frac{q''_{av}}{\Delta T} \qquad \Delta T = T_{\mathrm{wall}} - T_{\mathrm{fluid}}$$



1. *Cylinder*
   $\mathrm{Nu} = C(\mathrm{Re})^n$ [NOTE: Re_{CR} = 200,000]

Re	*C*, gases	*C*, liquids	*n*
0.4–4	0.891	0.989 $\mathrm{Pr}^{1/3}$	0.330
4–40	0.821	0.911 $\mathrm{Pr}^{1/3}$	0.385
40–4000	0.615	0.683 $\mathrm{Pr}^{1/3}$	0.466
4000–40,000	0.174	0.193 $\mathrm{Pr}^{1/3}$	0.618
40,000–400,000	0.0239	0.0266 $\mathrm{Pr}^{1/3}$	0.805

2. *Sphere* [NOTE: Re_{CR} = 200,000]
   $\mathrm{Nu} = 0.37\ \mathrm{Re}^{0.6}$
   $17 < \mathrm{Re} < 70{,}000$, gases
   $\mathrm{Nu} = (1.2 + 0.53\ \mathrm{Re}_D^{0.54})\ \mathrm{Pr}^{0.3}$
   $1 < \mathrm{Re} < 200{,}000$, $\mathrm{Pr} > 3$

D. *Natural convection from a horizontal cylinder* (Nu, Gr *based on diameter D*)
$\mathrm{Nu} = C(\mathrm{GrPr})^n$

GrPr	*C*	*n*
10^3–10^9	0.53	$\frac{1}{4}$
10^9–10^{12}	0.13	$\frac{1}{3}$

E. Natural convection from vertical surfaces (Nu, Gr *based on height L*)
$\mathrm{Nu} = C(\mathrm{GrPr})^n$

GrPr	*C*	*n*
10^5–10^9	0.555	0.25
$> 10^9$	0.021	0.4

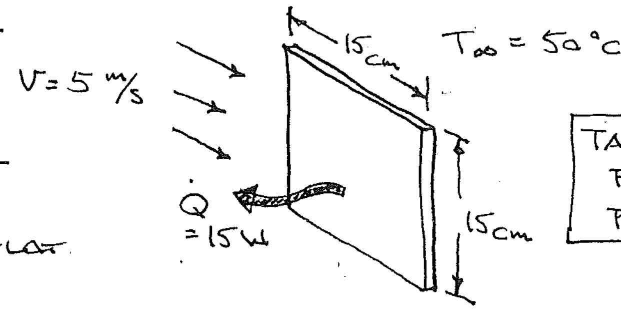
Compiled from W. M. Kays, *Convective Heat and Mass Transfer*, McGraw-Hill Book Company, New York, 1966; and W. H. McAdams, *Heat Transmission*, 3d ed., McGraw-Hill Book Company, New York, 1954.

EXAMPLE

A VERTICALLY ORIENTED 15cm × 15cm CIRCUIT BOARD UNIFORMLY DISSIPATES 15 WATTS OF HEAT FROM ITS COMPONENT SIDE ONLY. IT IS COOLED BY AIR AT 50 °C

FOR THE 3 SITUATIONS OUTLINED BELOW FIND THE SURFACE TEMPERATURE OF THE BOARD.

a) FORCED CONVECTION WITH AN AIR VELOCITY OF 5 m/s.
b) FORCED CONVECTION WITH AN AIR VELOCITY OF 5 m/s AND ASSUMING A TURBULENT FLOW AS THE COMPONENTS ACT AS TURBULATORS.
c) NATURAL CONVECTION



TAKE $T_f = 60$ °C FOR THIS PROBLEM.

a & b

STEP ①: SHAPE, SIZE & L
$L = 0.15$ m
GEOMETRY Ⓑ FLAT PLATE, C-8.

STEP ②: FORCED CONVECTION

STEP ③: FLUID PROPERTIES → 60 °C = 333 °K ≈ 330 K.
FROM TABLE A-19 FOR AIR
$\rho = 1.079$ kg/m³, $k = 0.0283$ W/m·°C, 1.99×10^{-5} kg/m·s
$\nu = 1.86 \times 10^{-5}$ m²/s, $Pr = 0.708$

STEP ④: $Re = \frac{VL}{\nu} = \frac{(5\ m/s)(0.15\ m)}{1.86 \times 10^{-5}\ m^2/s} = 4.03 \times 10^4 = 40{,}300$

a) $Re < 500{,}000$ ∴ LAMINAR FLOW.

b) TURBULENT FLOW DUE TO COMPONENTS.

STEP ⑤: $Nu = ?$

a) $Nu = 0.906\, Re^{1/2} Pr^{1/3}$ (UNIFORM WALL HEAT FLUX).
$= 0.906\,(40{,}300)^{1/2}(0.708)^{1/3}$
$= 169$

b) $Nu = 0.037\, Re^{0.8} Pr^{1/3}$ (TURBULENT).
$= 0.037\,(40{,}300)^{0.8}(0.708)^{1/3}$
$= 167$

STEP ⑥ : $h = ?$

a) $h = Nu \frac{k}{L} = (169) \frac{(0.0283)}{(0.15)}$

$= 31.9 \ W/m^2 \cdot °C$

b) $h = Nu \frac{k}{L} = (167) \frac{(0.0283)}{(0.15)}$

$= 31.5 \ W/m^2 \cdot °C$

STEP ⑦ : $T_s = ?$

$\dot{Q} = hA (T_s - T_\infty)$

$\therefore T_s = T_\infty + \frac{\dot{Q}}{hA}$

a) $T_s = 50 + \frac{15}{(31.9)(0.15 \times 0.15)}$

$= 70.9 \ °C$

b) $T_s = 50 + \frac{15}{(31.5)(0.15 \times 0.15)}$

$= 71.2 \ °C$

T_f CHECK...

$T_f = \frac{T_s + T_\infty}{2}$

$= \frac{71 + 50}{2}$

$= 60.5 \ °C$

(CLOSE TO OUR GUESS).

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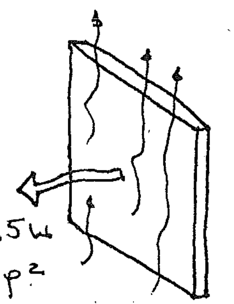
c) NATURAL CONVECTION



$T_\infty = 50 °C$

$T_f = 60 °C$ (ASSUMED)

$T_s = 70 °C$ (ASSUMED).

STEP ① : $L = 0.15 \ m$

GEOMETRY Ⓔ, VERTICAL SURFACE.

STEP ② : NATURAL CONVECTION

STEP ③ : DONE

STEP ④ : $Gr = \frac{g\beta (T_s - T_\infty) L^3 \rho^2}{\mu^2}$

$= \frac{(9.81)(0.003)(70-50)(0.15)^3(1.029)^2}{(1.99 \times 10^{-5})^2}$

$\beta = \frac{1}{60+273} = 0.003$

$= 5.84 \times 10^6$

STEP ⑤ : $Gr\,Pr = (5.84 \times 10^6)(0.708) = 4.135 \times 10^6$

$\therefore \quad Nu = C (Gr\,Pr)^n$

$= 0.555 (4.135 \times 10^6)^{0.25}$

$= 25.03$

STEP ⑥ : $h = Nu \frac{k}{L} = (25.03)\left(\frac{0.0283}{0.15}\right)$

$= 4.72 \ W/m^2 \cdot °C$

STEP ⑦ : $T_s = T_\infty + \frac{\dot{Q}}{hA} = 50 + \frac{15}{(4.72)(0.15 \times 0.15)}$

$\underline{T_s = 191.2 °C}$

ESTIMATES A BIT LOW I'D SAY

TABLE A–15

Properties of saturated water

Temp. T, °C	Saturation Pressure P_{sat}, kPa	Density ρ, kg/m³ Liquid	Density Vapor	Enthalpy of Vaporization h_{fg}, kJ/kg	Specific Heat c_p, J/kg · K Liquid	Specific Heat Vapor	Thermal Conductivity k, W/m · k Liquid	Thermal Conductivity Vapor	Dynamic Viscosity μ, kg/m · s Liquid	Dynamic Viscosity Vapor	Prandtl Number Pr Liquid	Prandtl Number Vapor	Volume Expansion Coefficient β, 1/K Liquid
0.01	0.6113	999.8	0.0048	2501	4217	1854	0.561	0.0171	1.792×10^{-3}	0.922×10^{-5}	13.5	1.00	-0.068×10^{-3}
5	0.8721	999.9	0.0068	2490	4205	1857	0.571	0.0173	1.519×10^{-3}	0.934×10^{-5}	11.2	1.00	0.015×10^{-3}
10	1.2276	999.7	0.0094	2478	4194	1862	0.580	0.0176	1.307×10^{-3}	0.946×10^{-5}	9.45	1.00	0.733×10^{-3}
15	1.7051	999.1	0.0128	2466	4185	1863	0.589	0.0179	1.138×10^{-3}	0.959×10^{-5}	8.09	1.00	0.138×10^{-3}
20	2.339	998.0	0.0173	2454	4182	1867	0.598	0.0182	1.002×10^{-3}	0.973×10^{-5}	7.01	1.00	0.195×10^{-3}
25	3.169	997.0	0.0231	2442	4180	1870	0.607	0.0186	0.891×10^{-3}	0.987×10^{-5}	6.14	1.00	0.247×10^{-3}
30	4.246	996.0	0.0304	2431	4178	1875	0.615	0.0189	0.798×10^{-3}	1.001×10^{-5}	5.42	1.00	0.294×10^{-3}
35	5.628	994.0	0.0397	2419	4178	1880	0.623	0.0192	0.720×10^{-3}	1.016×10^{-5}	4.83	1.00	0.337×10^{-3}
40	7.384	992.1	0.0512	2407	4179	1885	0.631	0.0196	0.653×10^{-3}	1.031×10^{-5}	4.32	1.00	0.377×10^{-3}
45	9.593	990.1	0.0655	2395	4180	1892	0.637	0.0200	0.596×10^{-3}	1.046×10^{-5}	3.91	1.00	0.415×10^{-3}
50	12.35	988.1	0.0831	2383	4181	1900	0.644	0.0204	0.547×10^{-3}	1.062×10^{-5}	3.55	1.00	0.451×10^{-3}
55	15.76	985.2	0.1045	2371	4183	1908	0.649	0.0208	0.504×10^{-3}	1.077×10^{-5}	3.25	1.00	0.484×10^{-3}
60	19.94	983.3	0.1304	2359	4185	1916	0.654	0.0212	0.467×10^{-3}	1.093×10^{-5}	2.99	1.00	0.517×10^{-3}
65	25.03	980.4	0.1614	2346	4187	1926	0.659	0.0216	0.433×10^{-3}	1.110×10^{-5}	2.75	1.00	0.548×10^{-3}
70	31.19	977.5	0.1983	2334	4190	1936	0.663	0.0221	0.404×10^{-3}	1.126×10^{-5}	2.55	1.00	0.578×10^{-3}
75	38.58	974.7	0.2421	2321	4193	1948	0.667	0.0225	0.378×10^{-3}	1.142×10^{-5}	2.38	1.00	0.607×10^{-3}
80	47.39	971.8	0.2935	2309	4197	1962	0.670	0.0230	0.355×10^{-3}	1.159×10^{-5}	2.22	1.00	0.653×10^{-3}
85	57.83	968.1	0.3536	2296	4201	1977	0.673	0.0235	0.333×10^{-3}	1.176×10^{-5}	2.08	1.00	0.670×10^{-3}
90	70.14	965.3	0.4235	2283	4206	1993	0.675	0.0240	0.315×10^{-3}	1.193×10^{-5}	1.96	1.00	0.702×10^{-3}
95	84.55	961.5	0.5045	2270	4212	2010	0.677	0.0246	0.297×10^{-3}	1.210×10^{-5}	1.85	1.00	0.716×10^{-3}
100	101.33	957.9	0.5978	2257	4217	2029	0.679	0.0251	0.282×10^{-3}	1.227×10^{-5}	1.75	1.00	0.750×10^{-3}
110	143.27	950.6	0.8263	2230	4229	2071	0.682	0.0262	0.255×10^{-3}	1.261×10^{-5}	1.58	1.00	0.798×10^{-3}
120	198.53	943.4	1.121	2203	4244	2120	0.683	0.0275	0.232×10^{-3}	1.296×10^{-5}	1.44	1.00	0.858×10^{-3}
130	270.1	934.6	1.496	2174	4263	2177	0.684	0.0288	0.213×10^{-3}	1.330×10^{-5}	1.33	1.01	0.913×10^{-3}
140	361.3	921.7	1.965	2145	4286	2244	0.683	0.0301	0.197×10^{-3}	1.365×10^{-5}	1.24	1.02	0.970×10^{-3}
150	475.8	916.6	2.546	2114	4311	2314	0.682	0.0316	0.183×10^{-3}	1.399×10^{-5}	1.16	1.02	1.025×10^{-3}
160	617.8	907.4	3.256	2083	4340	2420	0.680	0.0331	0.170×10^{-3}	1.434×10^{-5}	1.09	1.05	1.145×10^{-3}
170	791.7	897.7	4.119	2050	4370	2490	0.677	0.0347	0.160×10^{-3}	1.468×10^{-5}	1.03	1.05	1.178×10^{-3}
180	1,002.1	887.3	5.153	2015	4410	2590	0.673	0.0364	0.150×10^{-3}	1.502×10^{-5}	0.983	1.07	1.210×10^{-3}
190	1,254.4	876.4	6.388	1979	4460	2710	0.669	0.0382	0.142×10^{-3}	1.537×10^{-5}	0.947	1.09	1.280×10^{-3}
200	1,553.8	864.3	7.852	1941	4500	2840	0.663	0.0401	0.134×10^{-3}	1.571×10^{-5}	0.910	1.11	1.350×10^{-3}
220	2,318	840.3	11.60	1859	4610	3110	0.650	0.0442	0.122×10^{-3}	1.641×10^{-5}	0.865	1.15	1.520×10^{-3}
240	3,344	813.7	16.73	1767	4760	3520	0.632	0.0487	0.111×10^{-3}	1.712×10^{-5}	0.836	1.24	1.720×10^{-3}
260	4,688	783.7	23.69	1663	4970	4070	0.609	0.0540	0.102×10^{-3}	1.788×10^{-5}	0.832	1.35	2.000×10^{-3}
280	6,412	750.8	33.15	1544	5280	4835	0.581	0.0605	0.094×10^{-3}	1.870×10^{-5}	0.854	1.49	2.380×10^{-3}
300	8,581	713.8	46.15	1405	5750	5980	0.548	0.0695	0.086×10^{-3}	1.965×10^{-5}	0.902	1.69	2.950×10^{-3}
320	11,274	667.1	64.57	1239	6540	7900	0.509	0.0836	0.078×10^{-3}	2.084×10^{-5}	1.00	1.97	
340	14,586	610.5	92.62	1028	8240	11,870	0.469	0.110	0.070×10^{-3}	2.255×10^{-5}	1.23	2.43	
360	18,651	528.3	144.0	720	14,690	25,800	0.427	0.178	0.060×10^{-3}	2.571×10^{-5}	2.06	3.73	
374.14	22,090	317.0	317.0	0	—	—	—	—	0.043×10^{-3}	4.313×10^{-5}			

Note 1: Kinematic viscosity ν and thermal diffusivity α can be calculated from their definitions, $\nu = \mu/\rho$ and $\alpha = k/\rho c_p = \nu/\mathrm{Pr}$. The temperatures 0.01°C, 100°C, and 374.14°C are the triple-, boiling-, and critical-point temperatures of water, respectively. The properties listed above (except the vapor density) can be used at any pressure with negligible error except at temperatures near the critical-point value.

Note 2: The unit kJ/kg · °C for specific heat is equivalent to kJ/kg · K, and the unit W/m · °C for thermal conductivity is equivalent to W/m · K.

Source: Viscosity and thermal conductivity data are from J. V. Sengers and J. T. R. Watson, *Journal of Physical and Chemical Reference Data* 15 (1986), pp. 1291–1322. Other data are obtained from various sources or calculated.

TABLE A–22

Properties of air at 1 atm pressure

Temp. T, °C	Density ρ, kg/m^3	Specific Heat c_p, J/kg · K	Thermal Conductivity k, W/m · K	Thermal Diffusivity α, m^2/s	Dynamic Viscosity μ, kg/m · s	Kinematic Viscosity ν, m^2/s	Prandtl Number Pr
−150	2.866	983	0.01171	4.158×10^{-6}	8.636×10^{-6}	3.013×10^{-6}	0.7246
−100	2.038	966	0.01582	8.036×10^{-6}	1.189×10^{-6}	5.837×10^{-6}	0.7263
−50	1.582	999	0.01979	1.252×10^{-5}	1.474×10^{-5}	9.319×10^{-6}	0.7440
−40	1.514	1002	0.02057	1.356×10^{-5}	1.527×10^{-5}	1.008×10^{-5}	0.7436
−30	1.451	1004	0.02134	1.465×10^{-5}	1.579×10^{-5}	1.087×10^{-5}	0.7425
−20	1.394	1005	0.02211	1.578×10^{-5}	1.630×10^{-5}	1.169×10^{-5}	0.7408
−10	1.341	1006	0.02288	1.696×10^{-5}	1.680×10^{-5}	1.252×10^{-5}	0.7387
0	1.292	1006	0.02364	1.818×10^{-5}	1.729×10^{-5}	1.338×10^{-5}	0.7362
5	1.269	1006	0.02401	1.880×10^{-5}	1.754×10^{-5}	1.382×10^{-5}	0.7350
10	1.246	1006	0.02439	1.944×10^{-5}	1.778×10^{-5}	1.426×10^{-5}	0.7336
15	1.225	1007	0.02476	2.009×10^{-5}	1.802×10^{-5}	1.470×10^{-5}	0.7323
20	1.204	1007	0.02514	2.074×10^{-5}	1.825×10^{-5}	1.516×10^{-5}	0.7309
25	1.184	1007	0.02551	2.141×10^{-5}	1.849×10^{-5}	1.562×10^{-5}	0.7296
30	1.164	1007	0.02588	2.208×10^{-5}	1.872×10^{-5}	1.608×10^{-5}	0.7282
35	1.145	1007	0.02625	2.277×10^{-5}	1.895×10^{-5}	1.655×10^{-5}	0.7268
40	1.127	1007	0.02662	2.346×10^{-5}	1.918×10^{-5}	1.702×10^{-5}	0.7255
45	1.109	1007	0.02699	2.416×10^{-5}	1.941×10^{-5}	1.750×10^{-5}	0.7241
50	1.092	1007	0.02735	2.487×10^{-5}	1.963×10^{-5}	1.798×10^{-5}	0.7228
60	1.059	1007	0.02808	2.632×10^{-5}	2.008×10^{-5}	1.896×10^{-5}	0.7202
70	1.028	1007	0.02881	2.780×10^{-5}	2.052×10^{-5}	1.995×10^{-5}	0.7177
80	0.9994	1008	0.02953	2.931×10^{-5}	2.096×10^{-5}	2.097×10^{-5}	0.7154
90	0.9718	1008	0.03024	3.086×10^{-5}	2.139×10^{-5}	2.201×10^{-5}	0.7132
100	0.9458	1009	0.03095	3.243×10^{-5}	2.181×10^{-5}	2.306×10^{-5}	0.7111
120	0.8977	1011	0.03235	3.565×10^{-5}	2.264×10^{-5}	2.522×10^{-5}	0.7073
140	0.8542	1013	0.03374	3.898×10^{-5}	2.345×10^{-5}	2.745×10^{-5}	0.7041
160	0.8148	1016	0.03511	4.241×10^{-5}	2.420×10^{-5}	2.975×10^{-5}	0.7014
180	0.7788	1019	0.03646	4.593×10^{-5}	2.504×10^{-5}	3.212×10^{-5}	0.6992
200	0.7459	1023	0.03779	4.954×10^{-5}	2.577×10^{-5}	3.455×10^{-5}	0.6974
250	0.6746	1033	0.04104	5.890×10^{-5}	2.760×10^{-5}	4.091×10^{-5}	0.6946
300	0.6158	1044	0.04418	6.871×10^{-5}	2.934×10^{-5}	4.765×10^{-5}	0.6935
350	0.5664	1056	0.04721	7.892×10^{-5}	3.101×10^{-5}	5.475×10^{-5}	0.6937
400	0.5243	1069	0.05015	8.951×10^{-5}	3.261×10^{-5}	6.219×10^{-5}	0.6948
450	0.4880	1081	0.05298	1.004×10^{-4}	3.415×10^{-5}	6.997×10^{-5}	0.6965
500	0.4565	1093	0.05572	1.117×10^{-4}	3.563×10^{-5}	7.806×10^{-5}	0.6986
600	0.4042	1115	0.06093	1.352×10^{-4}	3.846×10^{-5}	9.515×10^{-5}	0.7037
700	0.3627	1135	0.06581	1.598×10^{-4}	4.111×10^{-5}	1.133×10^{-4}	0.7092
800	0.3289	1153	0.07037	1.855×10^{-4}	4.362×10^{-5}	1.326×10^{-4}	0.7149
900	0.3008	1169	0.07465	2.122×10^{-4}	4.600×10^{-5}	1.529×10^{-4}	0.7206
1000	0.2772	1184	0.07868	2.398×10^{-4}	4.826×10^{-5}	1.741×10^{-4}	0.7260
1500	0.1990	1234	0.09599	3.908×10^{-4}	5.817×10^{-5}	2.922×10^{-4}	0.7478
2000	0.1553	1264	0.11113	5.664×10^{-4}	6.630×10^{-5}	4.270×10^{-4}	0.7539

Note: For ideal gases, the properties c_p, k, μ, and Pr are independent of pressure. The properties ρ, ν, and α at a pressure P (in atm) other than 1 atm are determined by multiplying the values of ρ at the given temperature by P and by dividing ν and α by P.

Source: Data generated from the EES software developed by S. A. Klein and F. L. Alvarado. Original sources: Keenan, Chao, Keyes, Gas Tables, Wiley, 1984; and Thermophysical Properties of Matter. Vol. 3: Thermal Conductivity, Y. S. Touloukian, P. E. Liley, S. C. Saxena, Vol. 11: Viscosity, Y. S. Touloukian, S. C. Saxena, and P. Hestermans, IFI/Plenun, NY, 1970, ISBN 0-306067020-8.